Incineration Systems for Sludge Wastes

Assisted by the massive construction grant program of the Water Pollution Control Act, a large number of primary and secondary municipal and industrial wastewater treatment plants were brought on-line in the United States between 1970 and 1990. The plants generate large quantities of waste solids (402). These “biosolids” include several waste streams:

- “Primary” sludge (produced by gravity settling in the first stage of treatment)
- “Biological” (“secondary”) sludge produced in the activated sludge process
- “Advanced” wastewater treatment sludge from tertiary treatment process
- Other treatment plant solids such as the “screenings” recovered by the mechanical removal of large-dimension solids, “grit” recovered by settling sand and other coarse granular solids, and “scum” or “skimmings” recovered as the floatable solids skimmed from clarifiers

At the points of their generation in the treatment plant, these wastes vary in solids content over a very wide range (from 1% to 8% solids). If incineration is to be considered as an affordable process alternative, the sludge must first be dewatered to 20% solids or more. Note that in accordance with common usage, the relative proportions of dry solids and moisture in sludge are referenced in most of the wastewater treatment literature as “percent solids” rather than “percent moisture.”

The cooling effect of free water, by greatly slowing the overall combustion rate, is a key process characteristic of these high-moisture sludges. The outer layer of sludge dries and chars on introduction into a hot environment. The ash and char layer insulate the surface, thus reducing the rate of heat transfer to the interior. The high latent heat of evaporation of water in the interior further extends the time required for complete drying and combustion. The net impact of these effects is that the sludge incinerator must either (1) provide effective means to manipulate or abrade the sludge mass to disturb and/or wear off the protective ash/char layer and expose the wet interior to heat or (2) provide an extensive solids residence time.
Sludge is significantly different from municipal solid waste in both chemical and physical properties. In general, mixed wastewater treatment sludge is a relatively homogeneous, pumpable mass that exhibits the rheological properties of a “Bingham plastic.” It is characterized by a very high ratio of water to solids. Its chemical makeup is predominantly carbon, hydrogen, and oxygen but includes significant fractions of phosphorous and nitrogen. From a health effects point of view, biological sludge is relatively benign, with the exception of potential problems with pathogenic organisms and/or heavy metals.

A second class of sludge waste is generated in the course of a wide variety of organic chemical synthesis and other processing operations, petroleum product handling and refining, coke manufacture, etc. Although generated in lesser volume than the wastewater sludge on a national basis, the quantities of these sludges and pastes can also be significant. The characteristics of these sludges include (1) medium to low ratio of water to solids, (2) carbon and hydrogen as the dominant elemental constituents, (3) a high room-temperature viscosity (usually temperature-dependent), and (4) health effects that are often adverse from both the raw sludge and its combustion products.

The primary application of the incineration technologies discussed in Chapter 9 is for the management of wastewater treatment plant sludge. The multiple-hearth furnace (MHF) and the fluidized bed (FB) are the incinerator concepts most commonly used to address the unique requirements of this service. Other furnace designs discussed in this chapter were originally developed for domestic or industrial solid wastes and have since been adapted to burn biological sludge. Conversely, the MHF and FB systems can be, and sometimes are, used for the incineration of industrial and/or domestic solid wastes, although with some processing of the solids before feeding. Also, note that modified versions of the FB concept are used for industrial sludge not derived from wastewater treatment (e.g., their application in many refineries to burn API separator sludge and tank bottoms).

Rotary kilns are infrequently used for biological sludge. Kilns see their greatest application for some industrial wastes and for hazardous wastes (including the high heat content sludge) and are discussed in Chapter 11.

Because (most) sludge can be pumped, many of the problems of storing and feeding that are experienced with solid wastes are greatly simplified (although odor can be a severe problem after prolonged storage of biological wastewater treatment sludge). Conversely, the flow characteristics and small ash particle sizes of sludge make the use of a grate-type support during burning unacceptable. For this reason, sludge is burned on a hearth or in suspension.

Table 1  Estimated Incinerator Area Requirements

<table>
<thead>
<tr>
<th>System</th>
<th>Burning rate (ton/day)</th>
<th>Plant area (m²)</th>
<th>Specific area (m²/TPD)</th>
</tr>
</thead>
<tbody>
<tr>
<td>MHF, oxidizing mode</td>
<td>40</td>
<td>1000</td>
<td>25.0</td>
</tr>
<tr>
<td></td>
<td>300</td>
<td>3850</td>
<td>12.8</td>
</tr>
<tr>
<td>MHF, pyrolysis mode</td>
<td>12.5</td>
<td>500</td>
<td>40.0</td>
</tr>
<tr>
<td></td>
<td>200</td>
<td>1730</td>
<td>8.7</td>
</tr>
<tr>
<td>Fluid bed</td>
<td>20</td>
<td>400</td>
<td>20.0</td>
</tr>
<tr>
<td></td>
<td>120</td>
<td>400</td>
<td>8.3</td>
</tr>
<tr>
<td>Fluid bed and dryer</td>
<td>20</td>
<td>510</td>
<td>25.5</td>
</tr>
<tr>
<td></td>
<td>158</td>
<td>1500</td>
<td>9.5</td>
</tr>
</tbody>
</table>

Source: From (396).
The facilities used to burn sludge are relatively large. This reflects the need for redundancy (to provide high reliability in the volume reduction function) and the generally large dimensions of ductwork, scrubbers, conveyors, and other feed systems and bunkers and, of course, the furnaces themselves. Typical floor space requirements for sludge combustion facilities are shown in Table 1. Naturally, the plant area in a particular facility depends on the specific associated equipment, structural and layout interactions with other structures and facilities, etc.

I. MULTIPLE-HEARTH FURNACE (MHF) SYSTEMS

A. Process Characteristics

1. Fully Oxidizing Mode

The multiple-hearth furnace (Fig. 1) operated in the fully oxidizing mode is the most widely used wastewater treatment sludge incinerator in the United States. These units consist of a vertical cylindrical shell containing 4 to 14 firebrick hearths. A hollow cast-
iron or steel shaft is mounted in the center of the shell. The center shaft is rotated at 0.5 to 1.5 rpm. Two to four opposed arms are attached to the shaft and are cantilevered out over each hearth. A series of wide, spade-like teeth are mounted on the arms. As the arm rotates, the sludge is plowed or "rabbled" toward the center (on an "in-feed hearth") or toward the outside wall (on an "out-feed hearth").

The center shaft and rabble arms are insulated with refractory and cooled by air forced through a central "cold air tube" by a blower. The air returns via the annular space between the cold air tube and the outer walls: the hot air compartment (Fig. 2). The center shaft is provided with a rotating sand seal on the top and bottom to avoid infiltration of tramp air into the furnace.

Standard units can be purchased in diameters of 1.4 to 8.8 m. The capacity of these furnaces ranges from 100 to 3600 kg/hr of dry sludge. Each hearth has an opening ("drop-

Figure 2  Shaft cooling air arrangement in a multiple-hearth furnace.
hallow'') through which the sludge falls from hearth to hearth. The range of common (U.S. practice) furnace dimensions are given in Table 2 to illustrate typical furnace geometry. The drop-hole penetrations alternate in location from a wide clearance around the central shaft (on an “in-feed hearth”) to spaced holes at the periphery (an “out-feed hearth”). The sequence is set such that the top hearth is usually fed at the periphery and the bottom hearth discharges at the periphery. The mean residence time of solids in the furnace is variable but approximates 0.75 to 1.25 hours.

The rabbling process acts not only to move the sludge but also to cut, furrow, and open the surface as it passes through the drying, burning or combustion zone, and cooling zone. Observations of the furnace show a significant “freshening” of the luminous diffusion flame rising from the sludge bed on passage of the rabble arm. The flame dies down to a flicker in only a few moments, only to be rekindled with the passage of the next arm. The angle of the rabble teeth on the rabble arms is set both to move the sludge toward the exit opening and to generate ridges, thus increasing the effective exposed area to up to 130% of the plan area. An optimum rabble arm speed is where the width of the level portion in the valley of the furrows is approximately three cm. When rabble speed is too fast, this width will increase. When it is too slow, it will fill in with sludge.

Table 2  Typical Dimensions of Multiple-Hearth Incinerators

<table>
<thead>
<tr>
<th>Furnace O.D. (meters)</th>
<th>Cooling air (m³/min)</th>
<th>Outhearth area (m²)</th>
<th>Number of dropholes</th>
<th>Drophole area (m²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.98</td>
<td>20</td>
<td>—</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>2.29</td>
<td>25</td>
<td>—</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>2.59</td>
<td>30</td>
<td>2.26</td>
<td>24</td>
<td>0.57</td>
</tr>
<tr>
<td>3.28</td>
<td>60</td>
<td>3.38</td>
<td>18</td>
<td>0.85</td>
</tr>
<tr>
<td>3.89</td>
<td>70</td>
<td>5.59</td>
<td>20</td>
<td>0.93</td>
</tr>
<tr>
<td>4.34</td>
<td>85</td>
<td>6.88</td>
<td>24</td>
<td>1.30</td>
</tr>
<tr>
<td>5.11</td>
<td>115</td>
<td>9.19</td>
<td>18</td>
<td>2.93</td>
</tr>
<tr>
<td>5.71</td>
<td>140</td>
<td>14.05</td>
<td>30</td>
<td>2.30</td>
</tr>
<tr>
<td>6.78</td>
<td>200</td>
<td>19.18</td>
<td>30</td>
<td>4.51</td>
</tr>
<tr>
<td>7.85</td>
<td>280</td>
<td>28.96</td>
<td>36</td>
<td>5.42</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Furnace O.D. (meters)</th>
<th>Inhearth area (m²)</th>
<th>Drophole area (m²)</th>
<th>Insulated shaft O.D. (cm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.98</td>
<td>—</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>2.29</td>
<td>—</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>2.59</td>
<td>0.89</td>
<td>0.19</td>
<td>31.8</td>
</tr>
<tr>
<td>3.28</td>
<td>1.37</td>
<td>0.22</td>
<td>31.8</td>
</tr>
<tr>
<td>3.89</td>
<td>2.18</td>
<td>0.31</td>
<td>48.3</td>
</tr>
<tr>
<td>4.34</td>
<td>2.76</td>
<td>0.42</td>
<td>48.3</td>
</tr>
<tr>
<td>5.11</td>
<td>3.6</td>
<td>1.03</td>
<td>63.5</td>
</tr>
<tr>
<td>5.71</td>
<td>5.19</td>
<td>0.82</td>
<td>71.2</td>
</tr>
<tr>
<td>6.78</td>
<td>7.25</td>
<td>1.59</td>
<td>83.8</td>
</tr>
<tr>
<td>7.85</td>
<td>9.99</td>
<td>2.21</td>
<td>83.8</td>
</tr>
</tbody>
</table>

Refractory thickness approximately 0.33 meters.
In some instances, the angle of the rabble teeth can be reversed (“back rabbling”) to increase residence time and to control the location of the combustion hearth. Back rabbling can approximately double the residence time on the hearth.

Attempts to “move the fire” by changing the rabble arm speed has the secondary effect of either building or stripping the inventory of sludge on the hearths. Excessive sludge inventory can lead to a runaway condition where relatively large quantities of sludge begin to burn at one time. This overloads the air supply and leads to smoking and excessive hydrocarbon emissions. As a generalization, the use of rabbling speed as a day-to-day operating parameter is unwise.

Commonly, each hearth is equipped with two access doors. The doors have fitted cast-iron frames with machined faces to provide a reasonably gastight closure. In circumstances where air inleakage is critical (e.g., starved-air combustion, activated carbon regeneration, or charcoal manufacture), the doors may be gasketed and latched closed.

Several of the intermediate hearths are equipped with burners supplied with natural gas or No. 2 fuel oil. The air for the burners is supplied at a high pressure such that the flame and the excess air are forcibly projected into the overhearth volume. The air supplied through the burners typically supplies the air for sludge combustion. The overall excess air level used in multiple hearth systems varies widely. A good operator working with a feed system that supplies the sludge at a steady rate and consistent moisture content can run the units at 75% to 85% excess air. More typically, one finds the units operated at 100% to 125% excess air. Highly irregular feed conditions and/or indifferent operation leads to excess air levels well over 125%, with consequent excessive fuel and power consumption.

The temperature profile in the countercurrent sludge and flue gas flows depends on the relative magnitude of the following energy terms:

- Sludge character and feed rate (moisture content, dry solids heating value, and ash content).
- Combustion air quantities and temperature: In some units the warmed air from the hot air compartment (typically 120°C–175°C) is used for combustion air. The unused arm cooling air is often blended with the exhaust gas from the scrubber to minimize the persistence of the visible plume. Also, the hearth doors leak some air.
- Fuel firing rate: Usually, the burners are provided on the upper and several lower hearths. Also, screenings, grease, and scum are often added to one of the lower hearths.
- Heat loss from the outer shell.

In the idealized case, the temperature profile is as shown in Table 3.

<table>
<thead>
<tr>
<th>Table 3</th>
<th>Idealized Temperature Profile in Multiple-Hearth Furnace Burning Municipal Wastewater Treatment Sludge</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Drying zone</td>
</tr>
<tr>
<td>Sludge</td>
<td>70°C</td>
</tr>
<tr>
<td>Flue gases</td>
<td>425°C+</td>
</tr>
</tbody>
</table>
In theory, the MHF can be operated without generating an odorous off-gas: Little odoriferous matter is distilled until 80% to 90% of the water has been driven off (a sludge solids content of, say, 70%), and, at this point in the furnace, flue gas temperatures are high enough to burn out the odor. In practice, uncompensated variations in sludge moisture and/or heat content, inattentive or untrained operators, inadequate mixing and/or residence time of odorous off-gas, and other factors occur with sufficient frequency to almost ensure that odor will be a problem, at least from time to time. Protection from such problems includes the use of the top hearth as a secondary combustion chamber (with auxiliary fuel firing as needed) or the installation of a separate afterburner chamber.

In the design of MHF units for municipal sludge service, the parameters listed in Table 4 are reported design or operating values for basic parameters.

MHFs may also be used for the incineration of industrial wastes that have a high moisture content. This could include industrial biological wastewater treatment sludge, spent grains, or biomass from fermentation operations and the like. High organic content sludge (e.g., tank bottoms or tarry residues) are more efficiently and completely burned in other incinerator configurations.

MHFs have been tested as a device in which to co-incinerate mixtures of municipal sewage sludge and prepared municipal refuse. Refuse preparation includes shredding and removal of much of the glass and metals (142) (especially wire that tangles in the rabble plows). In this concept, sludge is fed to the top hearth and prepared refuse is fed either to the top or to an intermediate hearth. Several European plants where refuse is fed to the intermediate hearth experienced unacceptable odor emission problems. The problem apparently results from the variability in refuse heat content and consequent swings in the temperature of zones critical to odor destruction. No afterburner was installed. Details on these operations are available (143–147).

Data from the Uzwil co-incineration plant in Switzerland (147) indicate an estimated mean hearth heat release of 57,000 kcal hr⁻¹ m⁻² with a hearth loading of 31.3 to 33.8 kg hr⁻¹ m⁻² and a volumetric heat release of 42,700 kcal hr⁻¹ m⁻³. The exit gas temperature was 880 °C. The top two (of 12 original) hearths were removed to provide additional combustion space over the feed hearth.

### Table 4 Typical Design Parameters for Multiple-Hearth Incinerators

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Units</th>
<th>Low</th>
<th>Mean</th>
<th>High</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hearth area burning rate</td>
<td>kg dry solids hr⁻¹ m⁻²</td>
<td>7.2</td>
<td>9.8</td>
<td>16.2</td>
</tr>
<tr>
<td>Excess air percent overall</td>
<td></td>
<td>20</td>
<td>50</td>
<td>80</td>
</tr>
<tr>
<td>Cooling air exit temperature °C</td>
<td></td>
<td>95</td>
<td>150</td>
<td>195</td>
</tr>
<tr>
<td>Discharge ash temperature °C</td>
<td></td>
<td>38</td>
<td>160</td>
<td>400</td>
</tr>
<tr>
<td>Off-gas temperature °C</td>
<td></td>
<td>360</td>
<td>445</td>
<td>740</td>
</tr>
<tr>
<td>Sludge properties</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>heat content kcal/kg volatile</td>
<td></td>
<td>5300</td>
<td>5550</td>
<td>7760</td>
</tr>
<tr>
<td>volatile content % of dry solid</td>
<td></td>
<td>43.4</td>
<td>54.2</td>
<td>71.8</td>
</tr>
<tr>
<td>Total energy input (fuel plus sludge) kcal/kg</td>
<td></td>
<td>810</td>
<td>1100</td>
<td>1355</td>
</tr>
<tr>
<td>(for ~25% solids sludge)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(For ~48% solids sludge)</td>
<td></td>
<td>1595</td>
<td>1730</td>
<td>1922</td>
</tr>
<tr>
<td>Volumetric heat release (fuel plus sludge) kcal hr⁻¹ m⁻³</td>
<td></td>
<td>67,600</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
2. Starved-Air (Pyrolysis) Mode

MHFs can be operated with a restricted air supply (starved-air or pyrolysis mode) to give several desirable operating characteristics within a hardware system well proven in conventional, fully oxidizing service. Importantly, the MHF can be configured to permit switching back to the full combustion mode as a process backup.

In conventional MHFs, the system is allowed to operate with full burning. Thus, the gases leaving the furnace are primarily carbon dioxide, water vapor, and the nitrogen and excess oxygen from the air. In the pyrolysis mode, there is a deficiency of air relative to the stoichiometric requirement. In the starved-air mode, perhaps 80% of theoretical air is supplied to the furnace. This releases approximately 80% of the heat of combustion in the sludge. The off-gas contains a mixture of nitrogen, (mostly) low- and high-molecular-weight hydrocarbons, hydrogen, and considerable carbon monoxide. Because of the reducing conditions, one also finds HCN and NH₃ at high concentrations.

The first product of the pyrolysis process is a mixture of carbon and ash. With time, the carbon is gasified by reaction with oxygen, carbon dioxide, and steam according to

\[
\begin{align*}
C + O_2 & \rightarrow CO \text{ and/or } CO_2 \\
C + CO_2 & \rightarrow 2CO \\
C + H_2O & \rightarrow CO + H_2
\end{align*}
\]

Data from pilot studies (397) presented in Table 5 show the effect of temperature and gas composition on the hearth gasification rate.

The primary beneficial features of the MHF system in the starved-air mode relate to energy conservation. Lower excess air operation (only at 30% to 40% excess) can reduce fuel use. Clinker formation is reduced to a minimum, and refractory maintenance should be reduced due to less severe operating temperatures.

Air emissions (particulate, heavy metals, and organic emissions) are significantly lower in the pyrolysis mode than for MHF systems operating under fully oxidizing conditions (395). Pilot data on pyrolysis zone gas composition from Japan (394) show that from 43% to 87% of the sulfur in the cake remains in the residue. Seventeen percent to 51% of the sulfur was converted into SOₓ and between 0% and 7% into H₂S. Fifty-five percent to 87% of the chlorine in the cake remains in the residue, and the remainder converts into HCl. As discussed below, NOₓ generation was almost nil at the 60% stoichiometric air mode of the pyrolysis environment.

<table>
<thead>
<tr>
<th>Hearth temperature (℃)</th>
<th>Oxygen concentration (%)</th>
<th>CO₂ + steam concentration (%)</th>
<th>Gasification rate (kg/hr m²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>995</td>
<td>0</td>
<td>42–47</td>
<td>4.21</td>
</tr>
<tr>
<td>950</td>
<td>4</td>
<td>25–30</td>
<td>3.28</td>
</tr>
<tr>
<td>869</td>
<td>0</td>
<td>42–47</td>
<td>1.47</td>
</tr>
<tr>
<td>876</td>
<td>4</td>
<td>25–30</td>
<td>1.47</td>
</tr>
<tr>
<td>864</td>
<td>4</td>
<td>25–30</td>
<td>1.22</td>
</tr>
<tr>
<td>871</td>
<td>9</td>
<td>16</td>
<td>1.96</td>
</tr>
</tbody>
</table>

Source: From (398).
The oxidation of chromium from the relatively benign trivalent form to the carcinogenic hexavalent form is reduced almost to nil for pyrolysis mode operation. Data from Japanese researchers (393) showed that only 0.08% to 0.09% of the total chromium in the solid residue following pyrolysis mode decomposition was hexavalent in contrast to as much as 34% in the full incineration case. Indeed, the Cr\textsuperscript{6+} content of the ash was lower than that of the inlet sludge. A further benefit was the low solubility of all forms of chromium in the residue.

Data from Concord, California, tests (224) showed an off-gas higher heating value of about 800 kcal per dry standard cubic meter (approximately 5,200 kcal/kg). Tests in Japan (393) showed about 400 to 500 kcal/m\textsuperscript{3}. The off-gases in the Japanese tests contained about 60% of the fuel value of the original sludge. Thus, in the pyrolysis system, an “afterburner chamber” is required wherein a burner is mounted (to ensure ignition) and additional air (to, perhaps, only 40% excess overall) is added. Since most sludge incinerators are energy-deficient (the sludge fuel energy is insufficient to supply all needed heat), the savings in fuel from operation at this lower level of excess air can be substantial. About 5% of the sludge fuel value can be lost in unburned carbon in the ash unless the lower hearths are operated above the stoichiometric level to ensure burn-out.

Data on the composition of pyrolysis mode off-gas are summarized in Table 6. Some operators have registered concern about the potential for explosions in MHF systems operated in the pyrolysis mode. The expressed opinion is that since the furnace is filled with combustible gases, inadvertent opening of hearth doors will provide the air for an explosive event. To date, such an eventuality has not been realized in the limited operational experience of sludge incineration in the starved-air mode and, also, not in the operational records of the hundreds of MHF furnaces used for activated charcoal manufacture, charcoal regeneration, etc., where the furnace is maintained in a starved-air condition. In general, operator feedback from pyrolysis mode operations suggests that stable conditions could more easily be maintained under pyrolysis than in the oxidizing mode. Further, lower hearth temperatures almost ensure freedom from clinker formation.

### B. Process Relationships

#### 1. Retention Time

The retention time of solids in the multiple hearth furnace is proportional to shaft rpm, the number of rabble arms, and the plow angle settings of the rabble teeth. An average of 80

<table>
<thead>
<tr>
<th>Pollutant</th>
<th>Volume % in off-gas</th>
<th>Volume % after SCC</th>
<th>Pollutant</th>
<th>ppmdv in off-gas</th>
<th>ppmdv after SCC</th>
</tr>
</thead>
<tbody>
<tr>
<td>H\textsubscript{2}</td>
<td>4.27</td>
<td>ND</td>
<td>NO\textsubscript{x}</td>
<td>26</td>
<td>94</td>
</tr>
<tr>
<td>CO</td>
<td>5.69</td>
<td>ND</td>
<td>NO</td>
<td>25</td>
<td>87</td>
</tr>
<tr>
<td>CH\textsubscript{4}</td>
<td>0.79</td>
<td>ND</td>
<td>HCl</td>
<td>690</td>
<td>87</td>
</tr>
<tr>
<td>C\textsubscript{2}H\textsubscript{4}</td>
<td>0.5</td>
<td>ND</td>
<td>HCN</td>
<td>524</td>
<td>15</td>
</tr>
<tr>
<td>C\textsubscript{2}H\textsubscript{6}</td>
<td>0.016</td>
<td>ND</td>
<td>NH\textsubscript{3}</td>
<td>7125</td>
<td>11</td>
</tr>
<tr>
<td>CO\textsubscript{2}</td>
<td>8.80</td>
<td>2.95</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>N\textsubscript{2}, etc.</td>
<td>81.78</td>
<td>97.05</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

*Source: From (393).*
minutes retention time is required for a 20% solids sludge. The mean solids residence time \( t \) on a hearth is given by

\[
    t = \frac{r_h}{E \omega}
\]

where

\( r_h = \) hearth radius

\( E = \) a factor characterizing the “slip” between the solid material and the movement of the rabble teeth. An efficiency of about 50% is often used for calculations.

\( a = \) distance from the center of the furnace to the innermost rabble teeth

\( \omega = \) angular velocity

2. Heat Transfer

The rate of heat transfer from the furnace walls and the hot flue gases is a complex function of gas temperature, the concentration of radiating gases \((\text{e.g., CO}_2\) and \(\text{H}_2\text{O}\)) and soot and ash particles in the gas (affecting the mean gas emissivity), the relative velocity of gases over the solids, and so forth. Workers in the vendor community (195) suggest the use of an empirical overall heat transfer coefficient \((U')\) given by the following dimensional equation:

\[
    U' = \frac{T}{100} \text{ Btu/hr ft}^2\text{°F}
\]

where \( T \) is the average gas temperature in degrees Fahrenheit and \( U' \) is to be used (after conversion of units) with the log mean temperature difference in estimating the heat flux \( Q \) as follows:

\[
    Q = 35.28 U'A \Delta T_{\text{lm}} \text{ kcal/hr}
\]

where \( A \) is the area \((\text{m}^2)\) and the log mean temperature difference \((\text{°C})\) is given by

\[
    \Delta T_{\text{lm}} = \frac{(T_{g2} - T_{s1}) - (T_{g1} - T_{s2})}{\log_e \left( \frac{(T_{g2} - T_{s1})}{(T_{g1} - T_{s2})} \right)}
\]

where the subscripts on temperature refer to gas \((g)\) or solid \((s)\) temperatures at their respective inlet \((1)\) or outlet \((2)\) conditions for the hearth under study.

3. Pyrolysis and Combustion Processes

Pyrolysis reactions of sludge organic matter begin at about 200°C. The reactions are a temperature-dependent chemical reaction with a modest endothermic, then exothermic heat effect and are substantially complete when the temperature reaches about 650°C. Although not rigorous, the degree of pyrolysis effected at a given temperature \( T \) may be roughly estimated by

\[
    \text{fraction pyrolyzed} = 1.0663 \left(1 - e^{-0.7} \right)
\]

Pyrolysis results in partial gasification of the sludge. Approximately 85% of the sludge combustible is pyrolyzed to a gaseous product. The remaining 15% ends up as a char. The char burns in the solid phase or is subsequently gasified to \(\text{CO}\) and \(\text{CO}_2\). Ultimate burnout
of residual fuel values occurs in the gas phase. A typical rate of gasification of the char for temperatures above about 850°C in the fully oxidizing mode is about 2 kg hr⁻¹ m⁻².

Data from Japanese MHF pyrolysis experiments showed a two-step process beginning with pyrolytic breakdown of complex organic matter (393). This was followed by gasification of the solid char product by reactions with water vapor, carbon dioxide, and/or oxygen. The pyrolytic step at 800°C brought the overall degradation to 98.17% complete in 7.5 min (linear in ln[x − 1] versus time, where x is the decimal percent reacted). The slower char gasification process was also linear in the same coordinates: continuing from 98.17% degradation at 7.5 min to 99.75% at 20 min.

The gaseous products of pyrolysis range broadly and include simple, low-molecular-weight hydrocarbons, complex polyaromatics, and a wide variety of partially oxidized alcohols, aldehydes, ketones, and the like. Higher temperatures favor the simpler, low-molecular-weight compounds, and lower temperatures favor the tarry heavy oils.

As the sludge moves through the furnace, the average moisture content drops. However, the onset of combustion comes before the total sludge mass is thoroughly dried. Based on an analysis by Lewis and Lundberg (222), that presumption is in general agreement with observations of operating furnaces where vigorous combustion ensues when the mean moisture content has fallen to 53% (47% solids).

4. Heat and Material Balance Characteristics

The key process feature of the multiple-hearth furnace is the regenerative exchange of heat between the incoming sludge drying on the top hearths and the countercurrent rising flow of gases heated by combustion and ash cooling on the lower hearths. In order to realize this energy economy, the top hearth temperatures must be allowed to fall considerably below levels where combustion takes place (say, as low as from 300°C to 350°C). If the incoming sludge contains significant amounts of greases and oils, this may lead to emission of hydrocarbon-rich aerosols from the top hearths.

Regulatory trends in the United States increasingly require those gases in contact with raw sludge to be reheated (say, to 815°C), both to ensure sterilization and deodorization of the off-gas and to burn hydrocarbon vapors. If reheating is a firm requirement, the intrinsic energy efficiency of the multiple hearth is lost. The best-case fuel efficiency in this circumstance matches the multiple hearth with conventional fluid bed units but without the potential for energy recovery that can be achieved in the fluid bed with hot windbox designs.

The overall heat and material balance on a MHF is straightforward. The techniques and results are similar to those presented in Chapter 2. One problem with the furnace, however, is not revealed by such analyses: the sensitivity of the furnace to rapid temperature excursions. The sensitivity arises from the inherent structural design of the hearths, which are very flat, self-supporting refractory arches. Such a structure cannot tolerate the dimensional changes and expansion/contraction stresses associated with a rapid change in temperature. Consequently, relatively large quantities of fuel are used to sustain furnace temperature during no-feed periods, to very slowly bring up the furnace temperature on startup, and to drop the furnace temperature on shutdown.

Perhaps the greatest benefit of the MHF is its energy economy (when environmental regulations permit). Also, the inherent inertia of the system provides a flywheel effect, which allows absorption of feed fluctuations. The problems of the system, aside from the environmental emissions and clinkering problems noted above, result from the complexity and structural sensitivity of the design. The complexity requires considerable operator skill.
and, often, results in the simple expedient of turning up the air to wash out the need for fine trim. The structural problem arises due to the flatness of the hearths such that careful control over the rate of temperature rise and fall must be observed to avoid catastrophic hearth failures.

In the starved-air mode, the furnace itself is operated at between 30% and 90% of theoretical air. In an external afterburner chamber, sufficient air is added to bring the overall stoichiometry to 25% to 50% excess air. Clear advantages in fuel requirements are obvious in comparison to typical operation at 125% excess air and, with indifferent or inattentive operators, up to 200% excess air.

5. Operating Characteristics

Table 7 indicates typical production rates for MHFs burning a variety of sludge feeds. It can be seen that loading rates reflect a balance between the evaporative load and the sludge heat content (roughly scaled by the combustible content). There are significant differences in the observed typical furnace capacity between small and large incineration plants. This difference reflects heat losses, instrumentation and control systems and, to a degree, operator sophistication and training.

Table 8 shows the typical operating labor requirement for MHF installations. The table is based on systems with a high degree of automation. The labor budget includes labor for operation of the furnace, the scrubber, and the ash handling functions.

The multiple hearth is a large furnace with considerable thermal inertia. Thus, it can absorb substantial swings in feed sludge quantity and quality without producing unrecoverable upsets. However, the furnace is complex, with each hearth’s processes reflecting the contribution of other hearths’ off-gases, burners, and the sludge’s own contribution of heat absorption (evaporation) or heat release (combustion/pyrolysis). Because of the interplay of processes from hearth to hearth, the importance of feed stability to achieve fully satisfactory, energy-efficient, and environmentally acceptable system performance cannot be overstated. The staged character of the MHF hearth process means that cycling in feed rate results in the development of several combustion zones in the furnace. This can result in the discharge of still-burning sludge into the residue.

Table 7  Typical Hearth Loading Rates for Multiple-Hearth Furnaces

<table>
<thead>
<tr>
<th>Sludge type (<strong>), (</strong>)</th>
<th>Percent solids</th>
<th>Percent combustibles (dry basis)</th>
<th>Typical wet sludge loading rate (kg/m²/hr) (*)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Primary</td>
<td>30</td>
<td>60</td>
<td>34.3–58.8</td>
</tr>
<tr>
<td>Primary + FeCl₃</td>
<td>16</td>
<td>47</td>
<td>29.4–49.0</td>
</tr>
<tr>
<td>Primary + Low lime</td>
<td>35</td>
<td>45</td>
<td>39.2–58.8</td>
</tr>
<tr>
<td>Primary + WAS</td>
<td>16</td>
<td>69</td>
<td>29.4–49.0</td>
</tr>
<tr>
<td>Primary + WAS + FeCl₃</td>
<td>20</td>
<td>54</td>
<td>31.8–53.9</td>
</tr>
<tr>
<td>Primary + WAS + FeCl₃</td>
<td>16</td>
<td>53</td>
<td>29.4–49.0</td>
</tr>
<tr>
<td>WAS</td>
<td>16</td>
<td>80</td>
<td>29.4–49.0</td>
</tr>
<tr>
<td>WAS + FeCl₃</td>
<td>16</td>
<td>50</td>
<td>29.4–49.0</td>
</tr>
<tr>
<td>Anaerobically digested primary</td>
<td>30</td>
<td>43</td>
<td>34.3–58.8</td>
</tr>
</tbody>
</table>

(*): Lower number is applicable to small plants, higher number to large plants.
(**): WAS = Waste Activated Sludge.
Source: From (223).
conveyors with consequent equipment damage. Further, operator response to the situation where multiple combustion zones have emerged is often to increase the air supply. Increasing air results in greater fuel consumption and cost.

The use of high-pressure piston pumps as a means to feed multiple hearth systems at a steady rate and without introduction of tramp air has shown great success in stabilizing the burning front and improving environmental emissions and fuel economy. With stable feed, the multiple hearth system can be an effective and reliable operating system with few operating problems.

If the combustion-zone temperature becomes excessive relative to the sludge ash fusion point, sintered masses or ash or “clinkers” form that can block dropholes, bind rabble arm movement, and foul or jam the ash conveyor. This problem becomes excessive at temperatures above 1000°C. As noted above, the clinkering problem has been associated with the use of ferric chloride (with lime) as a conditioning aid for sludge dewatering. In the hydrocarbon- and CO/H₂-rich atmosphere within the mass of gasifying solids, ferric oxide formed by dehydration of ferric hydroxide is reduced. The shift in iron oxidation state decreases the ash fusion temperature. High-phosphorus-content sludge also exhibits low ash fusion temperatures.

Problem areas in MHF operation in sludge burning service (beyond the odor problems described above) include high refractory and rabble arm maintenance and the long time required to bring the units into service. Refractory and rabble arm problems arise from rapid temperature excursions (a too-rapid rise to temperature) and high temperatures due to localized overfiring. Ideally, the furnace should be brought to temperature (or cooled down) over a 24- to 30-hr period.

A second class of problems with MHF operation concerns air leaks (350). Since the furnaces are normally operated under draft, air leaks cool the furnace, increase fuel use, and, sometimes with crucial significance, add to the load on the induced draft fan. The four major areas for leaks include the upper shaft sand or water seal, the several furnace doors, the sludge feed, and the emergency bypass damper. The first two leakage points are readily correctable by proper maintenance and operator attention. The latter problems can be serious and are difficult and costly to resolve. The combination of damper and sludge feed leaks at typical furnace draft of 0.75 cm H₂O can lead to infiltration of almost 100 Nm³/min (350) so remedies will be important contributors to plant efficiency and capacity.

### Table 8 Labor Requirements for Multiple-Hearth Furnace Installations

<table>
<thead>
<tr>
<th>No. of furnaces</th>
<th>Operation</th>
<th>Maintenance</th>
<th>Total</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>2,920</td>
<td>1,460</td>
<td>4,380</td>
</tr>
<tr>
<td>3</td>
<td>8,760</td>
<td>4,380</td>
<td>13,140</td>
</tr>
<tr>
<td>5</td>
<td>14,600</td>
<td>7,300</td>
<td>21,900</td>
</tr>
</tbody>
</table>

Note: Assumes fulltime operation (7 days/week, 52 weeks/year) for highly automated process.

*Includes operation of the furnace, scrubber, and ash handling units.
The sludge feed system using a belt conveyor–chute feed is often a problem. Most sludge feed points use counterweighted, flap gates (some use double gates with the goal of having one gate close and seal before the second gate opens). The stickiness of the sludge often keeps the gates from closing fully. This can result in leakage areas of about 650 cm². Significant improvement has been observed in plants where the feed system has been converted to high-pressure diaphragm pumps that completely eliminate the leak or to the use of covered screw conveyors and chutes.

The emergency bypass butterfly damper, though made of high-alloy stainless steels, is susceptible to warping. Warping of the damper quickly leads to loss of the ceramic rope caulking, resulting in a leak area of about 350 cm². Damper leakage is a difficult problem, and no satisfactory solution for the butterfly damper has emerged. Converting to a sealing stack cap (equipped with a hydraulic opener) of the type used in cupolas has been suggested.

In the pyrolysis mode, there is considerably less operating experience than in the fully oxidizing condition. It should be noted that the oxygen-deficient pyrolysis mode requires a reversal of the “normal” response of the operator to temperature rise. In the fully oxidizing mode, temperature can be reduced by increasing the air supply. In the pyrolysis mode, however, adding oxygen increases temperature. Operator training was needed to develop this reversed “sense” for pyrolysis mode operation. Hearth-by-hearth temperature control (using air supply as the control means) should be installed for optimum performance. However, data from the Concord, California, plant (224) and from extensive pilot plant studies at Nichols’ Research and Engineering, Inc. (225) support the general conclusion that the multiple hearth in the pyrolysis mode is much easier to control than in the oxidizing mode. Further, due to the low hearth temperatures, no clinkering problems were observed.

The ash product from the pyrolysis mode (without special features) is much higher in combustible content than conventional MHF ash. Total combustible ranges from 3% to 30% including up to 20% elemental carbon (224). A portion of the combustible can be gasified by the addition of air to the lower hearths. Char burnout can also be effected by sparging steam into the furnace on the lower hearths. The water–gas reaction gasifies the carbon char (producing CO and hydrogen, which burn in the afterburner) without adding diluent nitrogen, which has an adverse impact on the overall heat balance.

6. General Environmental Considerations

The conventional MHF (with feed to the top hearth and no afterburner) can present problems with air emissions. The inherent regenerative feature of the multiple hearth can lead to excessive distillation of greases in a low-temperature zone with no subsequent means for control. At times, these greases are partially burned, leading to finely divided soot (opacity) and odor emissions. Also, incomplete burning on the top hearth can result in hydrocarbon emissions that exceed regulations. Raw sludge can be carried off by entrainment in the gas flow rising through the drop holes.

If the combustion and char burning hearth temperatures exceed 850° to 1000°C, volatilization of the oxides and chlorides of several toxic and/or carcinogenic metals (especially mercury, lead, arsenic, cadmium, chromium, and nickel) increases markedly. Subsequent cooling of the combustion gases in the upper hearths results in enrichment of the hard-to-collect fine-particle fraction of the particulate matter by these elements. The fine particles present a large fraction of the available surface area (if not the mass fraction) of the particulate matter on which condensation occurs. Since the fine particulate is the
hardest to collect in air pollution control devices, increased emissions of these environmentally important elements result in increased health risk in the area downwind of the incinerator. This topic is considered in greater detail in Chapter 13.

Other than the limited tests in Concord, California, the Nichols Research and Engineering investigations for the ISIS, and the Japanese testing (393), there are only limited data on the MHF pyrolysis operating configuration. Two full-scale commercial plants (in Alexandria, Virginia, and Cranston, Rhode Island) were designed and constructed to operate in the pyrolysis mode but have not done so. Several incinerators were run in starved-air mode for a short time when high heat content, low moisture sludge was being processed and temperatures became excessive in the fully oxidizing mode. Data indicate that the environmental performance of the multiple hearth when operated in the pyrolysis mode is more stable than in the oxidizing mode. If, for example, feed rates increase suddenly but the air rate is kept constant, furnace heat release does not change. By contrast, in the oxidizing mode, air is available to oxidize the excess combustible associated with increases in feed rate and furnace heat release and temperature will directly track feed-rate pulses.

Due to reduced and more stable heat release, pyrolysis mode hearth loading rates can be higher than for a comparable optimum incineration mode furnace. Wet-weight rate data (397, 399) when processing sludges ranging in solids content from 22% to 35% in a one-meter diameter pilot furnace showed 64 kg/hr-m² in the mode producing low-temperature char (HTC). HTC material is similar to charcoal (about 20% carbon, 80% ash), has an average bulk density of 0.26 g/cm³, and results from processing with a maximum solids temperature of 700°C. When producing high-temperature char (HTC), the processing rate rose to 78 kg/hr-m². HTC is a mixture of ash and fixed carbon (about 6.5% carbon, 93.5% ash) and is generated when the solids temperature is allowed to rise to 870°C. When operating to consume all of the carbon, where the average bulk density of the residue reaches 0.53 g/cm³, the rate falls to 59 kg/hr-m². Note that the wet processing rate was almost constant over the range of solids content, whereas the solids processing rate varied significantly.

Due to reduced gas velocities, particulate carryover from pyrolysis operations is reduced almost 50% in comparison to the fully oxidizing mode. Also, minimization of hearth temperatures reduces heavy metal emissions significantly, as predicted from theory. Although data are limited, reducing conditions in the furnace under pyrolysis mode conditions inhibit oxidation of chromium in the sludge from the relatively innocuous Cr³⁺ to the carcinogenic Cr⁶⁺ valence state form. Significant chromium oxidation (over 30% conversion in some furnaces) has been observed in fully oxidizing multiple hearth systems. Also, fuel nitrogen NOₓ generation appears reduced (224) in the pyrolysis operating mode. More data are required to fully characterize and demonstrate the environmental characteristics of the starved-air mode.

II. FLUID BED SYSTEMS

The fluidized bed furnace (FB), as applied in sludge incineration, is an inherently simple combustor. Air at high pressure is forced through a bed of sand. The sand particles become suspended in the rising gas and take on the behavior of a turbulent liquid: bubbling and flowing to maintain especially uniform temperatures throughout the bed volume. Typically, gas temperatures vary less than 5°C to 8°C between any one location in the bed and another. The gas velocities under these conditions average between 0.7 and 1.0 m/sec.
The top of the bed is relatively well defined, and the gas rising through the bed includes clearly defined gas bubbles. The hydraulic behavior of the fluidized bed is as though it held an ordinary liquid: solids with a lower density float; the upper surface is well defined and remains horizontal when the bed is tipped; the surface levels equalize when two chambers are interconnected; solids will overflow if the upper surface is higher than a drain point in the sidewall. This is the conventional, “bubbling fluid bed” mode of operation.

Above the fluid bed is a large, cylindrical disengaging space known as the freeboard. The freeboard usually provides about 3 to 4 sec of residence time for final burnout of combustible material. The freeboard operates at or slightly above the bed temperature. The finely divided ash is swept out of the bed and collected in a scrubber or other air pollution control system. Coarse or heavy particles remain in the bed: either decrepitating with time and blowing out or requiring removal through a drain.

At a given time, only a small portion (usually less than 1%) of the bed mass is combustible matter. The large mass of the bed gives it thermal inertia so that the bed can absorb fluctuations in feed characteristics without the problematical upset conditions that affect the MHF. Solids or sludge fed into the bed or into the freeboard are rapidly heated by radiation and intense convection. The rapid heat and mass transfer between bed constituents results in high-temperature uniformity, usually with not more than a few degrees Centigrade differential between any two parts of the bed.

Sludge cake introduced into the hot bed is abraded by the scrubbing action of the sand grains. The sludge particles rapidly dry and then burn, releasing most of their fuel value in the bed. A 5- to 8-cm-diameter plug of sludge discharged into the bed requires 20 to 30 sec to volatilize (343), while more dispersed feeds may gasify within only a few seconds.

The basic FB design concepts and control system logic draw from the petroleum and ore processing industries where the technology originated. The FB system was developed to provide a means to contact finely divided catalysts with high-molecular-weight petroleum feedstocks in a cracking or reforming process. Subsequent development led to a wide range of applications where the objectives included drying of high moisture content solids (e.g., ores), rapid heat and mass transfer to the feed (solid or gas), uniform temperatures, high thermal inertia (insensitivity to minor variations in feed character), and/or relatively short residence times (less than 5 sec).

FBs were introduced for the combustion of sewage sludge in 1962. Although its application in the United States to date has been dwarfed by MHF installations, the energy-efficient “hot windbox design” embodiment of the fluid bed is the system of choice in most new installations. This reflects the greater degree of control available and improved air emissions due to lower, more controllable temperatures (reduced NOx and heavy metal volatilization) and superior mixing (very low CO and total hydrocarbon emissions). Over 75 furnaces are operating in North America and many more in Europe. The capacities of fluid bed furnaces (in all services) range from 250,000 to 60 million kcal/hr and diameters from slightly over 1 m to 15 m.

FB technology rapidly penetrated the sludge incineration market when fuel was relatively cheap. However, as fuel costs rose steeply in the late 1970s, there were severe cutbacks in the rate of construction of new FB installations, and many existing units were shut down. The energy problem, importantly, derived from two factors. First, the early FBs were simple, plug-flow reactors. Without any regenerative heat feedback, the full price of the heat content corresponding to peak combustion temperatures was paid for with fuel.
Second, the mechanical dewatering equipment in wide use (primarily, the vacuum filter) was not very effective, so the evaporative energy demand was high. This combination made the FB very expensive to operate. The energy efficiency of the MHF, unencumbered with penalties (at that time) for hydrocarbon or carbon monoxide emissions, appeared attractive.

In response to the challenge of burgeoning energy costs, the vendor community developed a new, energy-conserving modification, the “hot windbox fluid bed.” Here, the hot off-gas from the freeboard is passed through a shell and tube heat exchanger to preheat the incoming combustion air. Initial designs heated the air to 500°C but, in modern plants, reheat to 850°C is achieved. Recycling heat substantially decreases the net fuel used in the FB.

Beyond the incineration of biological sludge, fluid beds also have been used for a wide variety of wastes including chemical process wastes, petroleum tank bottoms, coffee grounds and tea leaves, sludge from pharmaceutical, pulp and paper, and nylon manufacturing operations, waste plastics, carbon black waste, spent activated carbon, precoat filter cake, and waste oils and solvents. With preshredding and removal of a large fraction of the metal and glass, FBs have also been used to burn dry and wet-processed municipal refuse.

Another embodiment of the fluid bed concept uses higher gas velocities—high enough to continuously sweep the particles from the bed. Downstream, the larger particles are disengaged from the gas stream, passed to a return line, and reinjected into the bed until their combustion is complete. This “circulating fluid bed” design is generally not used for sludge but has found many applications in burning processed solid waste, coal, wood chips, anthracite culm, and other solid fuels and biomass in steam-raising and power generation installations. The circulating bed has also proved useful for the remediation of contaminated soils and other hazardous wastes. These applications are discussed in Chapter 11.

A. Process Characteristics

The conventional “bubbling fluidized bed” sludge incinerator (Fig. 3) is a vertical, cylindrical, refractory-lined vessel with a perforated grid in the lower section supporting a bed of graded sand. The depth of the static, “slumped” bed is usually 0.9 to 1.2 m. Consider the response of such an arrangement, initially at rest, as the flow rate of gas upward through the bed increases:

1. Initially, friction produces a pressure drop across the bed that increases with velocity.
2. At some point the pressure drop has increased until it is equal to the sum of the bed weight per unit cross-sectional area plus the friction of the bed against the walls. Beyond this velocity the bed either lifts and moves as a piston up and out of the chamber (unlikely with loose, granular solids) or the bed expands or fluidizes so the gas can pass without the pressure drop exceeding the bed weight. Pores and channels appear in the early stages of bed expansion.
3. With further increases in gas velocity to about 0.6 to 1.0 m/sec, particle-to-particle spacing increases and the violence of movement fills and eliminates the gross channels. The expanded bed height is, approximately, a linear function of the superficial gas velocity. In conventional biological sludge applications, a bed expansion about 30% to 60% in volume from the slumped bed is typical. Gas
and particles circulate in the bed with some transient streams of extra-high or extra-low particulate concentration but, in general, a high degree of bed uniformity. Bubbles form at the distribution plate and move rapidly through the bed, erupting at the bed surface. Note that it is not necessary to taper the sidewalls to achieve this behavior. Commercial practice includes both flare-then-straight and all-straight wall designs. The beds with higher velocities at the distribution plate (see following paragraph) favor the flared designs to avoid excessive sand loss.

The superficial gas velocity (the gas volumetric flow rate divided by the furnace cross-sectional area) varies with the particle size of the bed material. Conservative design practice suggests velocities at the distribution plate of 0.5 to 1.0 m/sec. Some recent designs, however, are based on velocities as high as 1.3 m/sec. The gas velocity in the freeboard is typically kept lower, averaging about 0.75 m/sec. Heat release rates in the bed range from 900,000 to 1,800,000 kcal/hr/m³. FB sludge combustors are normally sized at about 500,000 kcal/hr/m² of freeboard area. High-velocity FBs achieve over one million kcal/hr/m². Overall residence times of the gas at full temperature depend on the volume of the freeboard and ductwork but usually range between 4 and 8 sec. This extended plug-flow zone contributes to excellent burnout of gas phase and solid organic matter. In many
cases, equivalent destruction efficiencies are achieved in fluid beds to rotary kilns operated at 100° to 150°C lower temperatures.

In most instances, the fluid bed is of the “push through” design, where all gas handling is provided by the fluidizing blower. When one or more of the downstream systems (heat recovery, air pollution control, etc.) are such that they must operate under draft, a second induced fan is added ahead of the stack. A balance is struck between the fluidizing blower and the induced draft fan to create a null point (equal to atmospheric pressure) at a target location in the system. Often, the balance point is set just above the bed, thus providing a useful zone where difficult-to-feed materials can be readily charged using a chute or rotary “flinger” without problems with hot furnace gases leaks or excessive cold air in-leakage.

The FB incinerator is well suited to the drying and combustion of a wide variety of wastes. The sand in the bed is in a state of violent recirculating motion that maintains a remarkable uniformity in bed temperature and gas composition. The large mass of the sand, heated to bed temperature, provides a large thermal flywheel effect: protecting against rapid fluctuations in temperature if the feed rate or net heating value changes. For example, a 5-m-diameter freeboard FB reactor, during normal operation, contains about two million Kcal in the bed. As grains of hot sand strike the surface of the incoming sludge, a quantity of water is flashed off. The surface is rapidly dried and abraded so that no insulating layer of ash can develop. Particles of sludge that are sloughed off or abraded into the hot gas stream are ignited and begin to burn almost instantly.

Attention must be given to corrosion of the mild steel shell of the FB reactor when high concentrations of acid gas formers (e.g., sulfides, organic chlorides) are in the feed. Normally, a rubber lining between the steel and acid brick is provided. Alternatively, conditions can be set to ensure that the steel wall temperature is held high enough to prevent condensation. Wall temperatures are maintained at no less than 230° to 260°C where high sulfur concentrations exist and not below 110°C where HCl is present (401).

Other than the bed itself, key equipment includes:

- The fluidizing air turbine blower servicing air to the system. In some plants, a boiler is installed following the bed and the blower is powered with a steam turbine drive.
- The freeboard zone wherein combustion is completed. Sometimes overfire air jets assist through stimulation of mixing processes.
- An (optional) heat exchanger. The exchanger preheats fluidizing air countercurrent to the hot gases leaving the furnace.
- An (optional) boiler. The boiler generates hot water or steam from the hot gases leaving the furnace or the heat exchanger.
- The APC equipment. This commonly includes the Venturi and tray scrubber combination (especially in older plants), a dry electrostatic precipitator (used more often in Europe), or a combination of a Venturi scrubber (with or without a tray scrubber) and a wet electrostatic precipitator (ESP).

One of the most useful characteristics of the FB derives from its simple geometry. The cylindrical construction of the FB, the freeboard, and the associated ducting is associated with very robust refractory design and inherent resistance to refractory problems due to temperature excursions. This means that the FB can be shut down for short or long periods of time at the will of the operator. The shutdown (slumped) state loses heat to the surroundings very slowly (usually only 3° to 8°C per hour). Startup is therefore
rapid with only minimal (if any) fuel use to regain operating temperature. This contrasts with the MHF, where fuel firing is required during all standby periods because of the sensitivity of the refractory hearths to uneven and or rapidly changing temperature. Such flexibility is compatible with, say, five-day-per-week or 12-hour-per-day operating schedules except, possibly, at the “wet–dry” interface at the entrance to the scrubber.

It should be noted that the prolonged shutdown of a FB unit can result in severe corrosion problems with dry air pollution control systems. For example, a mild steel, dry ESP would be vulnerable to rapid wastage after the metal temperature passes through the dew point. Isolation and heating of the precipitator after shutdown are conceptually effective to combat this problem but, obviously, present risk. Hazardous waste incinerators in Denmark equipped with mild steel ESPs were successful in avoiding corrosion by washing down the units on shutdown. This problem is not relevant for systems equipped with wet scrubbers.

The ability to shut down the FB from time to time without a significant fuel or maintenance penalty helps to compensate for the limited turndown capability of the furnace. Because of the inherent dynamics of the FB, the gas flow must be within relatively narrow bounds or the bed will collapse. In practice, the units cannot be run at much less than 80% or more than 110% of design capacity. If the shortfall in feed will continue for an extended period, one or more rings of brick can be laid in the furnace to reduce the effective diameter and, in effect, de-rate the furnace. This “turndown” is, clearly, not readily changed but can be useful in those cases where the sludge quantity increases significantly over several years and only one furnace is desired.

Tapered beds facilitate turndown. If operation at low rates is to continue over a long period of time, consideration should be given to changing to a finer grade of bed sand. Increasing the excess air level allows operation at lower feed rates but is often uneconomical. In rectangular beds (more common for circulating-type fluid beds boilers in industrial or power generation service than for sludge incineration), turndown has been achieved by compartmentalizing the windbox and then allowing one compartment of the bed to slump.

1. Fully Oxidizing Mode

Almost all FBs are operated in the oxidizing mode. Excess air levels are maintained at 20% to 40% to minimize fuel costs but ensure complete oxidation of all volatile solids in the sludge cake. The burnout of sludge volatile matter is excellent with, typically, far less than 0.1% unburned matter in the fly ash.

2. Starved-Air (Pyrolysis) Mode

The Hyperion treatment plant in Los Angeles was the first to operate in the starved-air mode. Dewatering is effected using a special, multiple-effect evaporation method (Carver–Greenfield technology) to develop an almost completely dry powder. The dried sludge is then burned in a four-stage, starved-air system. The bed is operated at 30% of theoretical air. Additional air is added in the freeboard to bring the total to 80% of theoretical air. At this point, the sludge volatile matter is fully gasified to include, importantly, the nitrogen compounds in the sludge. The gases then pass to the first of two afterburners, where the air supply is brought to stoichiometric and cooled in a boiler. The final stage is a fully oxidizing afterburner where the total air reaches 135% of theoretical. This design limits NOX formation from the oxidation of sludge-based nitrogenous matter as well as thermal NOX.
The operation of a FB unit under substoichiometric conditions is not without precedent. Reducing conditions are used for the processing of iron ores, for iron reduction, and for activated carbon regeneration in fluid beds. With steady sludge flow, the excellent mixing in the fluid bed unit is, indeed, conducive to the attainment of a uniform product gas having the predominant form of unrealized fuel value as hydrogen and carbon monoxide (plus small quantities of low-molecular-weight hydrocarbons). The secondary (afterburner) chamber can involve high turbulence (with air jets and chamber design combining) to ensure complete burnout.

B. Process Relationships (Oxidizing Mode)

1. Heat and Material Balance Characteristics

Combustion occurs largely within the bed. This is important since the heat released in the freeboard is not available to dry the incoming sludge. In most FBs burning sludge, the bed is run between 730°C and 840°C. Also, in the general case, the freeboard temperature is, perhaps, 50°C or more higher. This differential is normal, but higher levels should be avoided. Overall excess air levels between 30% and 50% are normal.

For FB sludge incineration at 760°C and 40% excess air, combustion is autogenous (no fuel is required) at an energy parameter (EP) of 425 kg H₂O/million kcal. As the feed sludge becomes dryer and "hotter" (lower EP), the bed temperature will rise. The maximum bed temperature is set by a combination of factors: materials of construction, design features (especially the ability of the distribution plate to accommodate thermal expansion), and the desire to keep the bed cool enough to prevent an approach to the ash fusion temperature of the bed material (rapidly leading to bed defluidization) or excessive volatilization of heavy metals.

Ideally, combustion is completed within the sand bed such that a maximum of energy from sludge combustion is available to dry the incoming sludge. The fraction of the burning that takes place in the freeboard increases as the sludge grease-content increases and when the sludge is dropped through the freeboard space. This starves the bed for drying energy and may lead to overheating (slag formation and adhesion) in the freeboard and outlet flues.

Auxiliary fuels used in sludge incineration operation include light fuel oil (No. 2) and coal. Coal proved especially effective in a West Virginia plant (474). There, a 22% solids at 45% volatile solids mixed primary and secondary sludge was burned using 0.635 cm × zero, 1% sulfur, 15% ash coal (HHV 6,111 kcal/kg) to reach the target bed temperature of 785°C to 850°C using 650°C preheated air. The coal was drawn from a silo and fed through a metering screw to two tapered-barrel feed screws along with the sludge. The screws fed the 4.88-m inside-diameter incinerator (6.1-m ID freeboard); no other mixing device was needed. All air permit requirements were met. Freeboard temperatures have averaged only 30°C above the bed temperature (a smaller difference than was observed with oil firing).

Natural gas firing in the bed can be a problem from several standpoints: For good combustion of methane, the bed must be maintained above about 800°C. Also, the discharge of a jet of raw gas tends to overtax the air supply as the gas is swept upward and, often, a significant fraction of the heating value of the gas is released in the freeboard. Thus, when gas is the preferred fuel, it is generally best to consider gas firing of the windbox (to its temperature limits).
Because of the good sludge burnout effected by the FB, boilers can be added without undue risk of fouling. Steam is used in many plants to drive the forced draft fan, the induced draft fan, and/or the feedwater pumps or for general plant hot water, heating, etc. Consideration must be given to reducing the gas velocity in the boiler-tube banks to control erosion. In some plants, boilers are used in addition to the hot windbox design to cool the gases to approximately 260°C prior to an electrostatic precipitator for particulate control.

Boiler-tube surface can also be incorporated into the walls or tube U-bends inserted into the bubbling bed as a means of controlling bed temperature. Clearly, as the level of dewatering increases, the flame temperature will increase until, at some point, there is a risk of defluidization. Also, high bed temperatures (say, >875°C) are undesirable due to the excessive volatilization of heavy metals (especially cadmium, arsenic, lead, etc.). In such circumstances or when burning any fuel with a high flame temperature, one can consider the use of bed tubes to moderate temperatures.

In some circumstances, economic analysis has shown that the economic benefits of reduced or even zero (autogenous combustion) fuel use can be most cost effectively obtained by thermal drying of the sludge prior to incineration (rather than by upgrading dewatering facilities). Such systems use a boiler to recover heat from the fluid bed off-gas. The steam is then passed to an indirect dryer system. Several dryer types are available, but all have the characteristic that the feed sludge to the dryer should be about 60% to 65% solids. At this level of dryness, the rheological characteristics of the sludge are similar to damp sawdust or earth. If the sludge is fed at, say, 30% solids, there is a region in the dryer where the sludge moisture content ranges between 40% and 55% solids. In this range, the sludge may become very sticky and viscous. Then it adheres to the heating surfaces (reducing the heat transfer rate) and greatly increases the shaft power requirement. To resolve this problem, a portion of the dried sludge can be recycled and blended with the freshly dewatered sludge to develop the 60% to 65% solids feedstock.

In broad terms, the heat balance in the fluid bed defines the size of the unit. This is shown in the general capacity data in Tables 9 and 10.

2. Bed Solids and Bed Defluidization

a. Bed Solids. In some systems, the majority of the bed material is purchased, graded sand. The maximum superficial space velocity in the bed (the gas velocity calculated from

<table>
<thead>
<tr>
<th>Reactor diameter (m)</th>
<th>Cold windbox heat release ( \times 10^6 ) kcal/hr</th>
<th>Hot windbox (*) water evaporation (kg/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>3.05</td>
<td>2.420</td>
<td>1,590</td>
</tr>
<tr>
<td>4.25</td>
<td>4.760</td>
<td>3,068</td>
</tr>
<tr>
<td>5.5</td>
<td>7.890</td>
<td>5,045</td>
</tr>
<tr>
<td>6.7</td>
<td>11.590</td>
<td>7,545</td>
</tr>
<tr>
<td>7.6</td>
<td>15.195</td>
<td>9,727</td>
</tr>
<tr>
<td>8.5</td>
<td>19.075</td>
<td>12,182</td>
</tr>
</tbody>
</table>

*Preheat to 540°C.

Source: From (334).
the open bed diameter and with cognizance of the temperature and static pressure) increases as the mean particle size increases. A common sludge incineration sand uses 10 to 80 mesh alumina or ceramic material. With biological sludges, sand replacement rates vary with the rate of “inherent makeup” from the sludge itself, but a rate of 1% per day is common. If grit is burned along with the sludge, there may be an accumulation of solids in the bed such that solids removal is required from time to time. Most plants include a pneumatic sand makeup system as well as a bed drain.

The slumped bed depth ranges from 0.8 to 1 m deep and undergoes an expansion by 80% to 100% during fluidization. The size of solids that may be fluidized varies from less than 1 μm to 10 cm, with the best (most uniform) operation for particles between 10 and 250 μm. A range of particles is preferred. Mono-disperse (narrow \( d_p \) range) beds with large particles are prone to slugging and can have poor top-to-bottom circulation. This results in increased elutriation of incompletely burned solids, higher freeboard burning, and, potentially, greater fuel use.

A particle will remain in the bed as long as the gas flow rate is less than the terminal velocity of the particle as given by

\[
U_t = \frac{4g(\rho_p - \rho_g)d_p}{3\rho_gC_x} \tag{6}
\]

where

- \( U_t \) = terminal velocity (the minimum fluidizing velocity)
- \( g \) = gravitational constant
- \( \rho_p \) = density of particle
- \( \rho_g \) = density of the gas (varies with temperature and pressure)
- \( d_p \) = particle diameter
- \( C_x \) = drag coefficient (a dimensionless constant that is a function of the particle shape and Reynolds number based on particle dimensions. See Eq. 2 of Chapter 13 and Chapter 13, Section I.A.3.a.).
The lowest operating velocity for sludge incineration beds begins at about $1.2U_t$ but ranges to as much as $2U_t$ to $10U_t$ \cite{343}. Conservative sludge bed designs are based on a space velocity (based on freeboard diameter) between 0.8 and 0.9 m/sec, which corresponds to a maximum heat release between 350,000 and 500,000 kcal/h/m². Some vendors recommend higher velocities to increase heat release rate (reducing capital cost). As the velocities increase, however, sand losses increase, fluidization may become unstable, and the potential for increased freeboard burning increases.

The materials chosen for the bed sand are selected in view of their resistance to abrasion and melting temperature and, of course, cost. In recent years, some concern has been expressed regarding the potentially adverse health effects on plant personnel of silica-containing sand. Alternatives are readily available, however. Sand replacement rates vary depending on the degree that the inert solids in the feed contribute useful bed material. Typical makeup rates approximate one percent per week.

\textbf{b. Bed Defluidization.} Bed defluidization is a critical failure mode in FBs. The characteristics of the problem are obvious: The bed temperature rises such that the bed sand reaches a condition of incipient fusion. The resulting stickiness of the sand particles leads to agglomeration and, rapidly, to solidification of the bed. Tests of sludge ash residues using differential thermal analysis methods \cite{151} or by placing the residue in combustion boats and holding the mass at progressively higher temperatures in a muffle furnace will help to identify the conditions where clinker formation is likely. Also, the “initial deformation temperature” in the ASTM Ash Fusion Temperature test (with, perhaps, a safety factor of at least 100°C) can be used to anticipate this type of problem.

Although defluidization is driven by a physical phenomena (melting), the cause and control often derive from the chemistry of the ash. The sludge ash components of particular importance in producing low-melting compounds or eutectics include iron, potassium, sodium, phosphorous, and the chloride and sulfate anions. Biological or chemical treatment that fixes phosphorous and potassium compounds in the sludge biosolids can exacerbate ash fusion problems.

For example, some wastewater treatment plants add soluble ferric compounds (e.g., ferric chloride) at the plant headworks for the removal of phosphates according to

\[
\text{PO}_4^{3-} + \text{Fe}^{3+} \rightarrow \text{FePO}_4
\]

As the weight fraction of the ferric phosphate precipitate in the total sludge ash increases, the ash fusion temperature decreases. In some plants, this has caused bed defluidization. The resolution of this problem came through addition of calcium values (as lime) to the sludge such that the total overall iron-to-calcium ratio became at least equal to the proportions given by

\[
3\text{Ca}^{2+} + 2\text{FePO}_4 \rightarrow \text{Ca}_3(\text{PO}_4)_2 + 2\text{Fe}^{3+}
\]

Reaction of the lime with the ferric phosphate occurs in the bed (i.e., the lime could be added as late as the dewatering step). Thus, lime addition need not be made (at substantially higher cost) earlier in the plant in the dilute wastewater.

Another defluidization and slagging problem can occur when high concentrations of alkali metal salts are found in the sludge \cite{226}. This occurs, for example, in neutral sulfite semi-chemical (NSSC) waste liquor from paper mills or for petroleum refinery sludge where the salt content in the crude is high. In these circumstances, defluidization problems arise both from eutectics of the alkali metal salts and, also, with silica bed sands, from the
formation of very sticky, viscous, sodium-silicate glasses. For example, the combination of silica, sodium chloride, and water vapor from the bed will react:

\[3\text{SiO}_2 + 2\text{NaCl} + \text{H}_2\text{O} \rightarrow \text{Na}_2\text{O} \cdot 3\text{SiO}_2 + 2\text{HCl}\]

Similarly, for potassium chloride:

\[3\text{SiO}_2 + 2\text{KCl} + \text{H}_2\text{O} \rightarrow \text{K}_2\text{O} \cdot 3\text{SiO}_2 + 2\text{HCl}\]

Some of the sodium oxide–silica mixtures have melting points as low as 635°C. The ash fusion characteristics of alkali metal salts are discussed further in Chapter 4, and fusion point data are reported in Table 4 of that chapter.

**METAL OXIDES FOR DEFLUIDIZATION CONTROL.** Metal oxides such as CaO, Fe$_2$O$_3$, and Al$_2$O$_3$ can convert the low-melting silicate glasses into high-melting compounds (476). These metal oxides can be used to devitrify glasses after a duct or freeboard buildup problem has occurred. For example, the sodium glass noted above reacts with lime as follows:

\[\text{Na}_2\text{O} \cdot 3\text{SiO}_2 + 3\text{CaO} + 3\text{SiO}_2 \rightarrow \text{Na}_2\text{O} \cdot 3\text{CaO} \cdot 6\text{SiO}_2\]

When excess silica is absent, the following reaction occurs:

\[\text{Na}_2\text{O} \cdot 3\text{SiO}_2 + 2\text{CaO} \rightarrow \text{Na}_2\text{O} \cdot 2\text{CaO} \cdot 3\text{SiO}_2\]

Both of these complex products melt above 1000°C, which is usually considerably above the fluid bed operating temperature. If lime presents problems (e.g., scale in a Venturi scrubber or hygroscopic clumping in a bag filter), finely divided iron oxide or alumina also can neutralize alkali silicate glasses by the reactions

\[\text{Na}_2\text{O} \cdot 3\text{SiO}_2 + \text{Fe}_2\text{O}_3 + \text{SiO}_2 \rightarrow \text{Na}_2\text{O} \cdot \text{Fe}_2\text{O}_3 \cdot 4\text{SiO}_2\]

\[\text{Na}_2\text{O} \cdot 3\text{SiO}_2 + \text{Al}_2\text{O}_3 + 3\text{SiO}_2 \rightarrow \text{Na}_2\text{O} \cdot \text{Al}_2\text{O}_3 \cdot 6\text{SiO}_2\]

The iron-based product (acmite) has a melting point of 955°C and the aluminum product (albite) a melting point of 1118°C.

Problems with phosphorous eutectics have also been addressed with additives (338). For example, in the presence of calcium chloride, ferric phosphate will react at only 815°C to form calcium phosphate as a solid (s) and release ferric chloride as gas (g) at these temperatures:

\[2\text{FePO}_4(\text{s}) + 3\text{CaCl}_2(\text{s}) \rightarrow \text{Ca}_3(\text{PO}_4)_2(\text{s}) + 2\text{FeCl}_3(\text{g})\]

The ferric chloride reacts with water vapor to form specular hematite and releases HCl:

\[2\text{FeCl}_3(\text{g}) + 3\text{H}_2\text{O}(\text{g}) \rightarrow \text{Fe}_2\text{O}_3(\text{s}) + 6\text{HCl(\text{g})}\]

The hematite can form scales and layers of red deposits on the dome of the fluidized bed system. Lime can be used to avoid this scaling problem by mixing lime with the feed sludge, converting iron phosphate to iron hydroxide and calcium phosphate in the feed hopper. It has been shown that 2.71 kg of lime (100% soluble CaO basis) are needed per kg of soluble phosphorous. Note that lime values (as soluble calcium) in the sludge act to reduce the requirement for added lime. The reaction proceeds according to

\[2\text{FePO}_4(\text{s}) + 3\text{Ca(OH)}_2(\text{s}) \rightarrow \text{Ca}_3(\text{PO}_4)_2(\text{s}) + 2\text{Fe(OH)}_3(\text{s})\]

In the fluid bed, at temperatures of 800°C, the hydroxide dehydrates according to

\[2\text{Fe(OH)}_3(\text{s}) \rightarrow \text{Fe}_2\text{O}_3(\text{s}) + 3\text{H}_2\text{O}(\text{g})\]
The dehydration occurs in the bed, and therefore gaseous ferric chloride is not formed and the hematite scale does not form.

**CLAY (KAOLIN) FOR DEFLUIDIZATION CONTROL.** Kaolin clay (a natural mixture of hydrous aluminum silicates) has been used to prevent glassification or bed defluidization before the problem occurs by neutralizing the alkali salts before a low-melting mixture forms. The kaolin, ground very finely and well-mixed with the feed sludge, reacts to form mixed, high-melting sodium–potassium aluminum silicate compounds (226, 337) with a much higher melting point. The reaction sequence is as follows:

First, the kaolin clay dehydrates according to

\[
\text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 \cdot 2\text{H}_2\text{O} \rightarrow \text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 + 2\text{H}_2\text{O}
\]

The dehydrated clay neutralizes sodium (or potassium) chloride by forming nepheline (melting point 1526°C) and releasing HCl according to

\[
\text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 + 2\text{NaCl} + \text{H}_2\text{O} \rightarrow \text{Na}_2\text{O} \cdot \text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 + 2\text{HCl}
\]

If Na₂SO₄ is present in the bed, it also reacts with the clay to form nepheline but releasing oxygen and SO₂ according to

\[
\text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 + \text{Na}_2\text{SO}_4 \rightarrow \text{Na}_2\text{O} \cdot \text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2 + \text{SO}_2 + \frac{1}{2} \text{O}_2
\]

When using kaolin to neutralize sodium and potassium, the clay is added at a kaolin-to-alkali weight ratio of 5.6 and 3.3, respectively, based on the soluble portion of the alkali metals. In this calculation, one must determine the percent purity of the clay (often about 90%) and, prudently, allow a 25% to 30% safety margin. These rates assume that there is no aluminum and silicon in the sludge. If these elements are present (and they often are), the clay dose can be reduced in proportion. In most cases, silica is present in excess, so for every kilogram of alumina (Al₂O₃) in the dry sludge solids, 258/54 or 4.78 kg less kaolin are needed (476).

**BED TEMPERATURE FOR DEFLUIDIZATION CONTROL.** An alternative strategy to deal with defluidization is strict temperature control. This approach has been used with the sludges from some petroleum refineries burning API separator sludge, water–oil emulsions from slop-oil recovery, tank cleanings and flotation clarifier sludges from wastewater treatment, low-strength caustic, and other waste liquids. Bed temperatures were maintained at only 720°C and, other than a slow buildup of soft ash on the heat exchanger tubes (hot windbox design), operation has been acceptable. A similar approach for spent NSSC liquor has proven successful by controlling bed temperature and limiting waste chloride content (226).

3. **The Distribution Plate**

The distribution plate (227) serves several functions, the most important of which is the uniform distribution of the combustion air over the entire bed. The plate includes an array of tuyers (the metal devices that penetrate the distribution plate and pass the combustion air from the windbox to the active bed). The tuyers are fabricated in several designs but have the common characteristic that when the bed is either active or shut down (slumping), the sand does not drain into the windbox. The pressure ratio across the tuyers varies from 10 : 1 to 3 : 1, with the higher ratios providing the greatest degree of uniformity in flow throughout the furnace. Common pressure drop ranges between 0.035 and 0.055 atm.
4. Sludge Feed

Sludge feed systems have been designed to inject the sludge into the bed, to drop the sludge onto the bed from the freeboard dome, or to cast out the sludge over and into the bed from strategically located sidewall locations. In general, feed systems are chosen on the basis of the solids content of the waste stream. Each approach presents certain advantages and disadvantages.

a. Bed Injection. Direct injection into the bed was the dominant approach used in the early days of FB sludge incineration and remains the preference in the United States. From 1960 to 1970, most of the sludge was dewatered using centrifuges or vacuum filters and, at best, was a comparatively liquid 16% to 18% solids. The injectors were, in essence, shallow flight screw extruders that forced the sludge into the bed against a back pressure of 0.1 to 0.2 atm. The high moisture content of the sludge helped to cool the barrel of the injectors, but there was only a modest extension of the injector pipe into the bed in most cases.

In the late 1970s, improved dewatering led to abrasion problems with the screw injectors. Also, the reduced moisture content gave little leeway before an injector that had stopped for even a short period was almost solidified. The same time period saw the development of high-pressure, piston sludge pumps that could handle stiff sludge (over 35% solids). Although these pumps are costly, they are a reliable means to introduce the sludge, feed at a steady and controlled rate, and provide considerable design and layout flexibility.

The injection of sludge at a given point in the sidewall leads to a measure of “focus” of the heat demand for the evaporation of moisture in the bed sector above the injection point. As the feed rate increases, eventually the sector becomes overloaded and a portion of the combustion heat release moves to the freeboard. A useful design basis suggests that for a freeboard temperature of 815°C, there should be one feed point for each 1,130 kg of feed water. For a freeboard working temperature of 925°C, there should be about one feed point for each 2,275 kg of feed water.

b. Freeboard Injection. An alternative feed method used in the early days involved injection of the sludge at the roof of the freeboard dome. The sludge fell against the rising bed gases and into the bed. This approach was low in cost and simple and worked well when the sludge was wet. However, as the sludge moisture dropped, the surface of the sludge “chunk” falling through the freeboard dried enough to flash off greases and oils and to allow a portion of the volatile combustion to occur in the freeboard. Freeboard temperatures rose. This acted to starve the bed of heat. Compensating fuel firing in the bed raised freeboard temperatures even more. Bed defluidization became a continuing threat. Few U.S. plants are operated with feed injection through the freeboard, although this approach is preferred in Europe.

c. Sidewall Chute or Rotary Vane Feeders. When the furnace diameter exceeds 5 to 6 m, one must be increasingly concerned with the effective use of the entire bed volume if conventional design practice is followed and all of the sludge is fed from the wall. The use of long, cantilevered feed points is conceptually feasible but presents structural problems at high bed temperatures and requires uninterrupted cooling for survival.

For very large beds where one wishes to inject at least some portion of the feed sludge deep into the bed and one has a highly dewatered sludge, an alternative feeding strategy can be used involving the use of a “flinger” feeder. The flinger, a device similar to
a mechanical spreader stoker, is then mounted at the balance point just above the top of the expanded bed. These feed devices can easily cast the sludge 3 to 5 m into the center of the bed. This device can also be used to feed other relatively dry materials such as screenings, grit, and other irregular or abrasive material that is not compatible with a pump feed.

With a flinger or with a gravity chute, the fluid bed should be designed with both forced and induced draft fans to define a plane at, say, one meter above the top of the expanded bed that can be held at atmospheric pressure (see below). An opening for a gravity chute or a flinger-type feeder can then be made without undue concern for either blowout of furnace gases or unreasonable in-leakage of cold air.

5. Air Supply and Flow Balance

Air at 0.2 to 0.35 atm is forced into the windbox and passes into the cylindrical furnace through a refractory or refractory-lined “distribution plate” or “constriction plate.” The pressure drop across the distribution plate and sand bed is about 0.15 atm, and the remainder of the total static pressure provides for the exhaust gas treatment (scrubber) system. Initially, the sand bed rests on the distribution plate. As the rate of air flow increases, the sand bed expands. The bed growth results in a density high enough that sludge will not float to the top of the bed, yet with insufficient air flow to blow the sand out of the reactor. Typically, the superficial air velocity at the bottom of the bed (based on the bed diameter and windbox temperature) is 0.5 to 1.0 m/sec. The air supply system is usually designed with a 10% to 15% safety factor. More detailed information on fluid bed design and operational characteristics may be found in (148–150, 343).

The general design regarding air supply and flue gas handling can be of two types: the push-through concept and the push-and-pull concept. For push-through, there is only one blower. This fan pushes the air into the system with sufficient static pressure to meet the distribution plate and scrubber pressure drop requirements and to accommodate the duct pressure drop and discharge kinetic energy. With this approach, the static pressure in the ductwork is above atmospheric at all points.

In the push-and-pull concept, the combustion air is provided with one blower and an induced draft fan is provided after the scrubber. With this system, the static pressure in the ductwork can be controlled to meet special needs. These can include a neutral point where the internal static pressure is just balanced with atmospheric pressure (useful at an injection point for solids). Also, since the ductwork is generally all below atmospheric pressure, there is no discharge of dust through small holes, at imperfectly sealed hatches, etc.

6. Combustion Air Preheat

To improve energy economy, recent designs often include a combustion air preheater. If only a modest amount of preheat is required, the energy for preheat can come from steam heating coils (the “warm windbox design”). This approach is useful to achieve preheat levels up to 175°C. Ideally, the steam is supplied from a boiler fitted to the FB.

When higher levels of preheat are required, air from the combustion air blower is passed on the shell side of a stainless steel shell and tube heat exchanger to preheat it prior to introduction to the windbox. Flue gases from the furnace are passed through the tube side. Typically, the tube diameter is about 75 cm and gas velocities range from 30 to 40 m/sec. With this technique, preheat levels of up to 650°C are reliably achievable. The heat exchanger used in this service is of special design and is fabricated of carefully selected materials; often high-nickel alloys such as Alloy 625. The expansion joints on the
tubes must work over a wide temperature range and under moist, abrasive conditions. The heat exchanger is the most vulnerable device in the incineration train and is critical to system availability. This piece of equipment is not the place to save money. Even with sound design, replacement of the expansion joints (the weak link) should be expected at five-year intervals. Complete tube bundle replacement every 10 years is normal.

Preheating the combustion air makes significant demands on the design of the windbox. The plenum, the distribution plate, and the supporting structure are simple for cold windbox FB systems. However, increasing the air temperature demands careful design to provide adequate structural strength and to compensate for the significant thermal expansion. Some vendors use proprietary refractory arch designs to achieve these ends. Others use metal ducts to service the tuyers and run with a relatively cold backside to the distribution plate. Both techniques are satisfactory although they generate different capital and operating cost tradeoff scenarios.

7. The Freeboard

The freeboard is a relatively simple, cylindrical chamber. Functionally, however, the freeboard volume is very important to incinerator performance. Most important, the freeboard provides the residence time for ultimate burnout of the combustible matter (both in the gaseous flow and in the solid particulate). The freeboard provides 4 to 6 sec of residence time in many designs, sufficient for complete burnout of almost all organic material. This accounts for the low carbon monoxide and hydrocarbon pollutant emission from the FB and for the low percent combustible in the fly ash.

The diameter of the freeboard bears a relationship to the constriction plate diameter. However, since more water is evaporated per unit volume of air in a hot windbox unit, the constriction-plate-to-freeboard-area ratios are different for the two configurations. The specifics vary between incinerator manufacturers, but common ratios would be 1.66 for the cold windbox design and 1.86 for the hot windbox.

When the material being burned has an exceptionally high volatile content or the degree of burnout must be exceptionally high (as in hazardous waste incineration applications), it may be desirable to incorporate overfire jets in the freeboard to stimulate mixing and, in some cases, to add additional air. The jets can use steam or air for this purpose. Some designs use two levels of jets. The lower bank uses jets aimed at an angle such as to induce, say, a clockwise flow. The upper bank jets are aimed to induce a counterrotating flow so the net swirl effect is null but vigorous mixing is induced. The mass addition in these jets should be minimized to avoid chilling the combustion gas flow.

C. Operating Characteristics

The FB furnace is stable in operation. Commonly, oxygen instrumentation is used to adjust fuel feed to maintain a relatively constant excess air level and bed temperature. Because of the simple refractory design, furnace maintenance costs are low. Maintenance problems with the heat exchanger used in the hot windbox design have been experienced due to thermal stresses and fouling. In general, these problems can be handled with proper design and operation (especially avoidance of excessive temperatures).

Electrical usage by the fluid bed is high due to the energy used in the high-pressure fluidizing blower. Typical power usage is summarized graphically in Fig. 4.

As sludge burns out in the bed, the finer ash particles are swept from the bed. As a consequence, highly efficient air pollution control devices should be used with these units.
The coarser particles in the sludge residue accumulate in the bed. On balance, however, there is often a net loss in the sand bed due to abrasion and disintegration such that periodic sand addition is required. Accumulation of bed solids may occur, however, and means for withdrawal of material should be provided. Generally, only a simple downcomer is needed to drain bed solids. Data on the particle-size distribution of the fly ash elutriated from fluid bed incinenders burning domestic wastewater treatment sludge are given in Table 11.

Problem areas with fluid beds for sewage sludge or other liquid or sludge waste disposal center in two areas:

- Control of the temperature throughout the bed and flues
- Clinker (slag) formation

The former problem is particularly serious if the waste is fed into the freeboard. If continuous or periodic ignition occurs in this area (where mixing is very imperfect), overheating of regions of the flue may occur, causing slag buildup and necessitating shutdown. Problems with bed defluidization are serious and, if not resolved, render continuing operation of the FB problematical. Often, additive approaches have been successful, though at a cost for both the additive and increased residue disposal.

An unexpected problem can occur when firing natural gas through a lance to make up for energy deficiencies. In this instance, the natural gas appears to accelerate rapidly to the surrounding gas velocity, and thereafter mixing stops. Consequently, the gas passes up through the bed and, ultimately, burns in the freeboard: not where the energy release was needed.

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**Figure 4** Typical power usage in a fluid bed incinerator.
Published data on the Hyperion plant performance in the starved-air mode have been limited as of this writing. However, it is reported that the performance of the FB portion of the plant has been very satisfactory. It should be noted, however, that the Hyperion FB is fed with dried sludge and, therefore, its behavior may be atypical. At Hyperion, pneumatic feeder problems have been noted with its dry (>95% solids) feed.

D. General Environmental Considerations

Because of the excellent mixing in the bed and the long residence times of gases in the freeboard, emissions of hydrocarbons and carbon monoxide from the FB are small. If excessive emission of hydrocarbons becomes a problem, addition of air jets in the freeboard is often effective in realizing acceptable burnout. Fluid beds have achieved 99.9994% DREs for toluene at a bed temperature of only 700°C due to the excellent mixing and long (8-sec) residence times. Trial burns of various fluid bed incinerators have met or exceeded 99.99% DRE targets for chlorobenzene, aniline, carbon tetrachloride, chloroform, cresol, para-dichlorobenzene, methyl methacrylate, perchloroethylene, phenol 1,1,1-trichloroethane, and other hazardous chemicals (334). In most plants all of the sludge ash is emitted from the bed suspended in the flue gas. Thus, high-efficiency particulate control systems are necessary. The low bed temperatures of the FB allow the operator to avoid the heavy metal volatilization problem. Although most U.S. systems use the Venturi-tray scrubber system, many European plants, in combination with energy recovery in a boiler, have applied the electrostatic precipitator for particulate removal. Excellent particulate resistivity properties lead to a collection efficiency that meets or exceeds U.S. standards.

Data on the particle size of ash emitted from the furnace (402) indicate the following:

<table>
<thead>
<tr>
<th>Median diameter (μ)</th>
<th>Poughkeepsie, NY</th>
<th>Somerset-Raritan Valley, NJ</th>
<th>Liberty, NY</th>
<th>Bayshore, NJ</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.8</td>
<td>1.5</td>
<td>1.4</td>
<td>1.5</td>
<td>1.5</td>
</tr>
<tr>
<td>2.8</td>
<td>2.0</td>
<td>2.0</td>
<td>2.5</td>
<td>2.0</td>
</tr>
<tr>
<td>5.4</td>
<td>8.5</td>
<td>6.5</td>
<td>8.5</td>
<td>4.0</td>
</tr>
<tr>
<td>11.5</td>
<td>14.0</td>
<td>10.0</td>
<td>16.0</td>
<td>5.0</td>
</tr>
<tr>
<td>19.0</td>
<td>16.2</td>
<td>23.7</td>
<td>15.0</td>
<td>16.3</td>
</tr>
<tr>
<td>28.0</td>
<td>12.5</td>
<td>23.0</td>
<td>12.2</td>
<td>15.0</td>
</tr>
<tr>
<td>31.0</td>
<td>2.5</td>
<td>7.5</td>
<td>3.2</td>
<td>8.0</td>
</tr>
<tr>
<td>Residue</td>
<td>42.8</td>
<td>25.9</td>
<td>41.4</td>
<td>48.2</td>
</tr>
<tr>
<td>Total</td>
<td>100.0</td>
<td>100.0</td>
<td>100.3</td>
<td>100.0</td>
</tr>
</tbody>
</table>

Source: Courtesy of Dorr-Oliver, Inc. and participating communities.
**Operation of the starved-air mode fluid bed in a two-stage manner with intermediate heat removal is very beneficial from an environmental viewpoint.** As noted above, NO\textsubscript{x}, and heavy metal enrichment of fines (2) are minimized due to the low bed temperatures and the reducing gas environment. Also, the reducing environment would be expected to minimize chromium oxidation.

**III. SLAGGING COMBUSTION SYSTEMS FOR BIOLOGICAL SLUDGE**

Concern regarding the leaching of heavy metals from sludge ash has led to the development of several combustion systems wherein the sludge ash is heated beyond the point of fusion. Then, a glassy slag is formed. Leaching tests on the slag show limited to no solvation of metals. The Kubota Corporation of Osaka, Japan, drew on a furnace design developed at the Volkswagenwerks in the 1960s in Wolfsberg, Germany, to offer a “melting furnace” to achieve this functional behavior. Another Japanese firm, Itoh Takuma, uses a reciprocating grate furnace-boiler quite similar to those used for solid waste to achieve a similar endpoint.

**A. Kubota System**

The primary furnace used by Kubota makes use of a slowly rotating, refractory-lined cylindrical, cuplike chamber with an outlet tap in the center of the floor of the furnace (Fig. 5). The furnace is driven from the periphery. Mounted above the discharge tap is a fixed, relatively flat, refractory-lined, reverse-conical roof equipped with one or more down-firing burners. The feed of predried sludge (>85% solids) moves by gravity through the annulus between the roof and lower chamber into the cavity between the roof and the lower chamber. Movement of the sludge is “encouraged” by the slow rotation of the lower body. In the internal cavity, the burners, supplied with preheated air (about 85% of theoretical air overall), reduce the sludge to a molten slag. A portion of the combustion heat in the sludge solids is released by furnishing excess air (relative to the burner fuel requirements) through the burner ports. The size of the cavity may be adjusted to increase (larger cavity) or decrease (smaller cavity) the processing rate. The primary furnace cavity is maintained slightly substoichiometric. This produces maximum temperatures and minimizes the gas flow, thus reducing particle entrainment.

The molten slag and off-gas from the primary chamber flow down into a secondary combustion chamber. The slag continues to fall to a solidification area or directly into a water quench. Slow cooling favors crystal growth and forms a dense, obsidianlike black glass. Fast cooling in the water quench results in a granular black glass frit with poorer structural characteristics than the slow-cooled product. Addition of secondary air completes the combustion in the tunnel-type secondary chamber or afterburner.

The hot gases from the secondary are used to preheat the combustion air for the primary furnace (using a heat exchanger similar in design to the hot windbox fluid bed) and then are tempered and used in a direct contact tray or rotary dryer to dry the incoming sludge. Off-gas from the dryer is cleansed with an electrostatic precipitator, condenser, and scrubber.
The use of fossil fuel is minimized by the low overall excess air of the unit (10% to 30%). Temperatures in the primary furnace are maintained at approximately 1450°C for most sludge to yield acceptable sludge fluidity. Data indicate that acceptable long-term operation is possible. Several plants are now operating in Japan, the United States, and Germany using the Kubota melting process. The first plant came on-line in 1975. The furnace operation is stable, as seen from strip charts of temperature and gas composition. The operation of the primary chamber substoichiometric greatly limits the formation of fuel-nitrogen NO\(_x\). Net NO\(_x\) in the flue gases from several plants averaged 100 to 150 ppm. Although one might expect the high primary furnace temperatures to result in emission of heavy metals, data indicate that the majority of the metals are bound into the fluid slag that blankets the incoming sludge and do not show up as heightened metal emissions in the small particles.

The key environmental characteristic of the Kubota furnace is the ash, which exhibits essentially no leaching (below the detection limits by atomic absorption) for cadmium, chromium +6 and +3, arsenic, mercury, and lead. There was also no detection of PCBs or cyanide ion, although these species appear in the raw sludge. The slag formed has a specific gravity of 2.4 to 2.7 and appears useful as a clean fill or road bed material. Fuel requirements are modest, but the plant is complex, including dryers, several air cleansing systems, and a complex, high-temperature furnace and heat recovery system.

**B. Itoh Takuma System**

The Itoh Takuma system also requires a dried sludge feed (>85% solids), but the sludge is burned with controlled and limited air on a reciprocating grate. The grate is enclosed in a waterwall boiler with a configuration not unlike a refuse-fired incinerator. Steam raised in
the boiler is used to dry the sludge using an indirect dryer. The temperatures in the bed are high enough to sinter the sludge ash to produce a clinker similar in appearance to coal ash. Although the ash is not a true glass as is the Kubota product, the leaching characteristics are almost as good and the furnace system is of significantly lower cost and simplicity of operation.