Chapter 6

Design of Biomass Gasifiers

6.1 INTRODUCTION

The design of a gasification plant includes the gasifier reactor as well as its auxiliary or support equipment. A typical biomass gasification plant design comprises the following systems:

- Gasifier reactor
- Biomass-handling system
- Biomass-feeding system
- Gas-cleanup system
- Ash or solid residue-removal system

This chapter deals with the design of the gasifier reactor alone. Chapter 8 discusses the design of the handling and feeding systems. Gas-cleaning systems are briefly discussed in Chapters 4 and 9.

As with most process plant equipment, the design of a gasifier may be divided into three major phases:

Phase 1. Process design and preliminary sizing
Phase 2. Optimization of design
Phase 3. Detailed mechanical design

For cost estimation and/or for submission of initial bids, most manufacturers use the first step of sizing the gasifier. The second step is considered only for a confirmed project—that is, when an order is placed and the manufacturer is ready for the final stage of detailed mechanical or manufacturing design.

This chapter mainly concerns the first phase and, briefly, the second phase (design optimization). To set the ground for design methodologies, a short description of different gasifier types is presented, followed by a discussion of design considerations and design methodologies.

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TABLE 6.1 Comparison of Some Commercial Gasifiers

Parameters	Fixed/Moving Bed	Fluidized Bed	Entrained Bed
Feed size	<51 mm	<6 mm	<0.15 mm
Tolerance for fines	Limited	Good	Excellent
Tolerance for coarse	Very good	Good	Poor
Exit gas temperature	450–650 °C	800–1000 °C	>1260 °C
Feedstock tolerance	Low-rank coal	Low-rank coal and excellent for biomass	Any coal including caking but unsuitable for biomass
Oxidant requirements	Low	Moderate	High
Reaction zone temperature	1090 °C	800–1000 °C	1990 °C
Steam requirement	High	Moderate	Low
Nature of ash produced	Dry	Dry	Slagging
Cold-gas efficiency	80%	89%	80%
Application	Small capacities	Medium-size units	Large capacities
Problem areas	Tar production and utilization of fines	Carbon conversion	Raw-gas cooling

Source: Data compiled from Basu, 2006.

6.1.1 Gasifier Types

Gasifiers are classified mainly on the basis of their gas–solid contacting mode and gasifying medium. Based on the gas–solid contacting mode, gasifiers are broadly divided into three principal types (Table 6.1): (1) fixed or moving bed, (2) fluidized bed, and (3) entrained flow. Each is further subdivided into specific types as shown in Figure 6.1. Major western technology providers, as listed in the figure, supply their gasification technologies as per one of these.

One gasifier type is not necessarily suitable for the full range of gasifier capacities. There is an appropriate range of application for each. For example, the moving-bed (updraft and downdraft) type is used for smaller units (10 kWth–10 MWth); the fluidized-bed type is more appropriate for intermediate units (5 MWth–100 MWth); entrained-flow reactors are used for large-capacity units (>50 MWth). Figure 6.2 shows the overlapped range of application for different

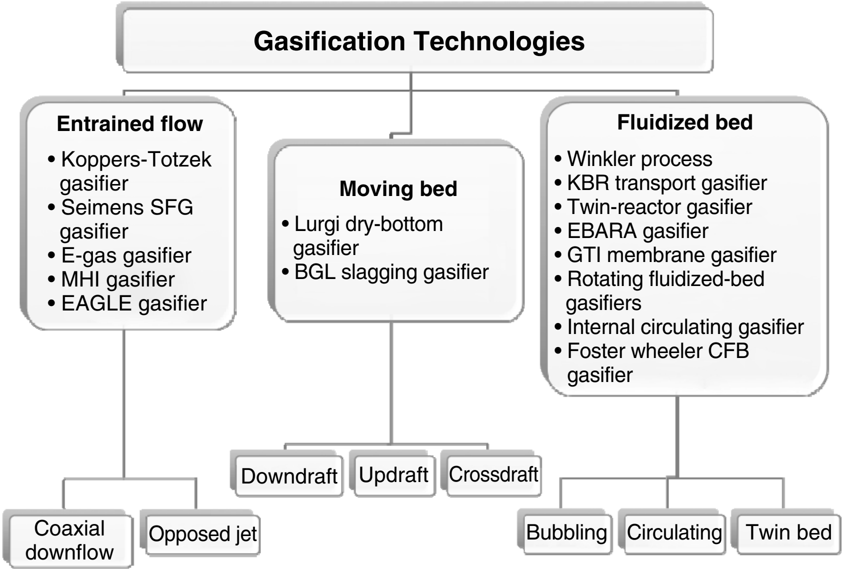

FIGURE 6.1 Gasification technologies and their commercial suppliers.

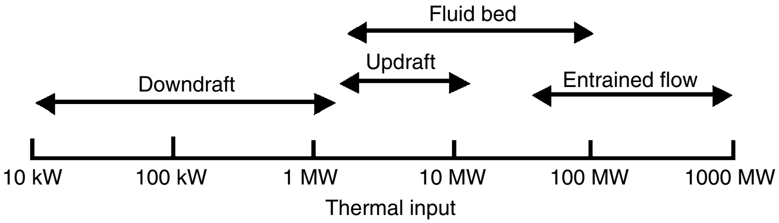

FIGURE 6.2 Range of applicability for biomass gasifier types.

types of gasifiers developed with data from Maniatis (2001) and Knoef (2005). Crossdraft gasifiers are for the smallest size while entrained flow are the largest size gasifiers.

6.2 FIXED-BED/MOVING-BED GASIFIERS

In entrained-flow and fluidized-bed gasifiers, the gasifying medium conveys the fuel particles through the reactor, but in a *fixed-bed* (also known as *moving-bed*) gasifier the fuel is supported on a grate (hence its name). This type is also called *moving-bed* because the fuel moves down in the gasifier as a plug. Fixed-bed gasifiers can be built inexpensively in small sizes, which is one of their major attractions. For this reason, large numbers of small-scale moving-bed biomass gasifiers are in use around the world.

TABLE 6.2 Characteristics of Fixed-Bed Gasifiers

Fuel (wood)	Updraft	Downdraft	Crossdraft
Moisture wet basis (%)	60 max	25 max	10–20
Dry-ash basis (%)	25 max	6 max	0.5–1.0
Ash melting temperature (°C)	>1000	>1250	
Size (mm)	5–100	20–100	5–20
Application range (MW)	2–30	1–2	
Gas exit temperature (°C)	200–400	700	1250
Tar (g/Nm^3)	30–150	0.015–3.0	0.01–0.1
Gas LHV (MJ/Nm^3)	5–6	4.5–5.0	4.0–4.5
Hot-gas efficiency (%)	90–95	85–90	75–90
Turn-down ratio (–)	5–10	3–4	2–3
Hearth load (MW/m^2)	<2.8		

Source: Adapted from Knoef, 2005, p. 26.

Both mixing and heat transfer within the moving (fixed) bed are rather poor, which makes it difficult to achieve uniform distribution of fuel, temperature, and gas composition across the cross-section of the gasifier. Thus, fuels that are prone to agglomeration can potentially form agglomerates during gasification. This is why fixed-bed gasifiers are not very effective for biomass fuels or coal with a high caking index in large-capacity units.

There are three main types of fixed- or moving-bed gasifier: (1) updraft, (2) downdraft, and (3) crossdraft. Table 6.2 compares their characteristics.

6.2.1 Updraft Gasifiers

An updraft gasifier is one of the oldest and simplest of all designs. Here, the gasification medium (air, oxygen, or steam) travels upward while the bed of fuel moves downward, and thus the gas and solids are in countercurrent mode. The product gas leaves from the top as shown in Figure 6.3. The gasifying medium enters the bed through a grate or a distributor, where it meets with the hot bed of ash. The ash drops through the grate, which is often made moving (rotating or reciprocating), especially in large units to facilitate ash discharge. Chapter 5 describes this process in more detail.

Updraft gasifiers are suitable for high-ash (up to 25%), high-moisture (up to 60%) biomass. They are also suitable for low-volatile fuels such as charcoal. Tar production is very high (30–150 g/nm^3) in an updraft gasifier, which makes

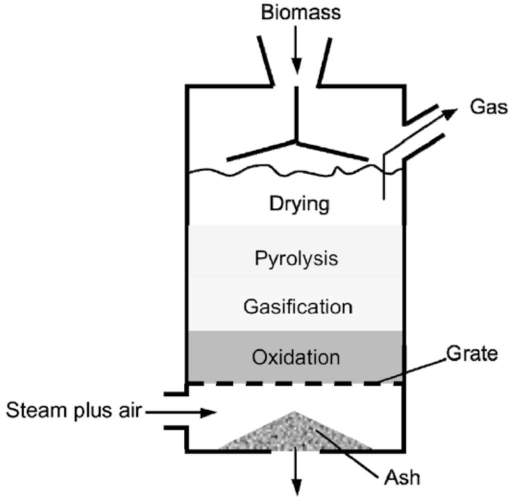

FIGURE 6.3 Schematic of an updraft gasifier.

it unsuitable for high-volatility fuels. On the other hand, as a countercurrent unit, an updraft gasifier utilizes combustion heat very effectively and achieves high cold-gas efficiency (Section 6.11.1). Therefore, it is more suitable for direct firing, where the gas produced is burnt in a furnace or boiler with no cleaning or cooling required. Since the gas is not fired in an engine or stored, the tar produced does not have to be cleaned.

Updraft gasifiers find commercial use in small units like cooking stoves in villages and in large units like South African Synthetic Oils (SASOL) for production of gasoline from coal. The following is a brief description of two important large-scale commercial updraft gasifier technologies.

Dry-Ash Gasifier

Lurgi, a process development company, developed a pressurized dry-ash updraft gasifier. It is called *dry ash* because the ash produced is not molten. One that produces molten ash is called a *slagging* gasifier.

Though the peak temperature (in the combustion zone) is 1200 °C, the maximum gasification temperature is 700 to 900 °C. The reactor pressure is in the neighborhood of 3 MPa, and the residence time of coal in the gasifier is between 30 and 60 minutes (Ebasco, 1981). The gasification medium is a mixture of steam and oxygen, steam and air, or steam and oxygen-enriched air. It uses a relatively high steam/fuel carbon ratio (~1.5).

The coal is first screened to between 3 and 40 mm (Probstein and Hicks, 2006, p. 162) and then fed into a lock hopper. The gasifying agent moves upward in the gasifier while the solids descend. The reactor is a double-walled pressure vessel. Between the two walls lies water that quickly boils into steam under pressure, utilizing the heat loss from the reactor. As the coal travels down the reactor, it undergoes drying, devolatilization, gasification, and combustion. Typical residence time in the gasifier is about an hour (Probstein and Hicks, 2006, p. 162). In a dry-ash gasifier, the temperature is lower than the melting point of ash, so the solid residue dries and is removed from the reactor by a rotating grate.

The dry-ash technology has been used at SASOL in South Africa, the world's biggest gasification complex. SASOL produces 55 million Nm^3/day of syngas, which is used to produce 170,000 bbl/day of Fischer-Tropsch liquid fuel.

Slagging Gasifier

The British Gas/Lurgi consortium developed a moving-bed gasifier that works on the same principle as the dry-ash gasifier, except a much higher temperature (1500–1800 °C) is used in the combustion zone to melt the ash (hence its name, *slagging gasifier*). Such a high temperature requires a lower steam-to-fuel ratio (~0.58) than that used in dry-ash units (Probstein and Hicks, 2006, p. 169).

Coal crushed to 5 to 80 mm is fed into the gasifier through a lock hopper system (Minchener, 2005). The gasifier's tolerance for coal fines is limited, so briquetting is used in places where the coal carries too many of them. Gasification agents, oxygen and steam, are introduced into the pressurized (~3 MPa) gasifier vessel through sidewall-mounted tuyers (lances) at the elevation where combustion and slag formation occur.

The coal introduced at the top gradually descends through several process zones. The feed is first dried in the top zone and then devolatilized. The descending coal is transformed into char and then passes into the gasification (reaction) zone. Below this zone, any remaining carbon is oxidized, and the ash content is liquefied, forming slag. Slag is withdrawn from the slag pool through an opening in the hearth plate at the bottom of the gasifier vessel. The product gas leaves from the top, typically at 400 to 500 °C (Minchener, 2005).

6.2.2 Downdraft Gasifiers

A downdraft gasifier is a co-current reactor where air enters the gasifier at a certain height below the top. The product gas flows downward (giving the name *downdraft*) and leaves through a bed of hot ash (Figures 6.4 and 6.5). Since it passes through the high-temperature zone of hot ash, the tar in the product gas finds favorable conditions for cracking (see Chapter 4). For this reason, a downdraft gasifier, of all types, has the lowest tar production rate.

Air from a set of nozzles, set around the gasifier's periphery, flows downward and meets with pyrolyzed char particles, developing a combustion zone (zone III shown schematically in Figure 6.5 and described in the discussion of throatless downdraft gasifiers that follows) of about 1200 to 1400 °C. Then the

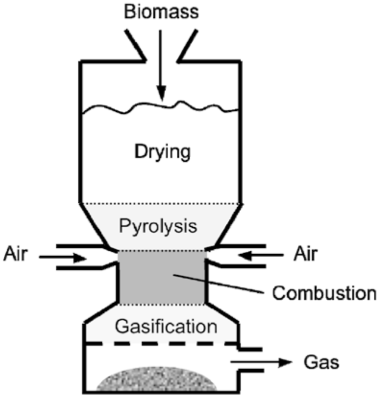

FIGURE 6.4 Schematic of a throated-type downdraft gasifier.

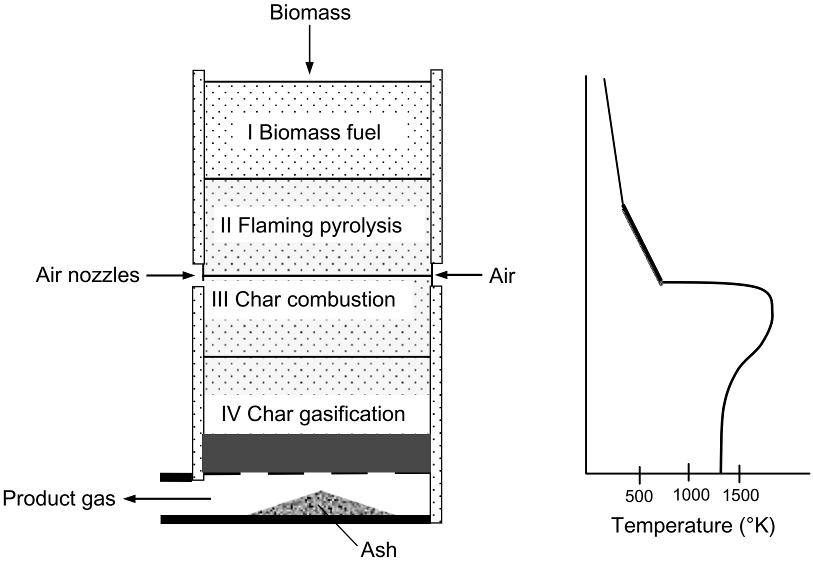

FIGURE 6.5 Schematic of the operation of a throatless downdraft gasifier. Temperature gradient along the height shown at the right.

gas descends further through the bed of hot char particles (zone IV), gasifying them. The ash produced leaves with the gas, dropping off at the bottom of the reactor.

Downdraft gasifiers work well with internal-combustion engines. The engine suction draws air through the bed of fuel, and gas is produced at the end. Low tar content (0.015–3 g/nm^3) in the product gas is another motivation for their use with internal-combustion engines. A downdraft gasifier requires a shorter time (20–30 minutes) to ignite and bring the plant up to working temperature compared to the time required by an updraft gasifier.

There are two principal types of downdraft gasifier. The throatless (or open core) type is illustrated in Figure 6.5. Reactions in different zones and at different temperatures are plotted on the right. The throated (or constricted) type is shown in Figure 6.4.

Throatless Gasifier

This gasifier type is also called *open top*, or *stratified throatless.* Here, the top is exposed to the atmosphere, and there is no constriction in the gasifier vessel because the walls are vertical. Figure 6.5 shows that a throatless design allows unrestricted movement of the biomass down the gasifier, which is not possible in the throated type shown in Figure 6.4. The absence of a throat avoids bridging or channeling. Open-core is another throatless design, but here air is not added from the middle as in other types of downdraft gasifiers. Air is drawn into the gasifier from the top by the suction created downstream of the gasifier. Such gasifiers are suitable for finer fuels—for example, lighter biomass such as rice husk.

The following are some of the shortcomings of a downdraft gasifier:

- It operates best on pelletized fuel instead of fine light biomass.
- The moisture in the fuel must not exceed 25%.
- A large amount of ash and dust remains in the product gas.
- As a result of its high exit temperature, it has a lower gasification temperature.

Operating Principle

Because an open-top, or a throatless, gasifier is simple in construction, it is used to describe the gasification process in the downdraft gasifier (Figure 6.5). The throatless process can be divided into four zones (Reed and Das, 1988, p. 39). The first, or uppermost, zone receives raw fuel from the top that is dried in air drawn through the first zone. The second zone receives heat from the third zone principally by thermal conduction.

During its journey through the first zone, the biomass heats up (zone I in Figure 6.5). Above 350 °C, it undergoes pyrolysis, breaking down into charcoal, noncondensable gases (CO, H_2, CH_4, CO_2, and H_2O), and tar vapors (condensable gases). The pyrolysis product in zone II receives only a limited supply of air from below and burns in a fuel-rich flame. This is called *flaming pyrolysis*. Most of the tar and char produced burn in zone III, where they generate heat for pyrolysis and subsequent endothermic gasification reactions (Reed and Das, 1988, p. 28).

Zone III contains ash and pyrolyzed char produced in zone II. While passing over the char, hot gases containing CO_2 and H_2O undergo steam gasification and Boudouard reactions, producing CO and H_2. The temperature of the down-flowing gas reduces modestly, owing to the endothermic gasification reactions, but it is still above 700 °C.

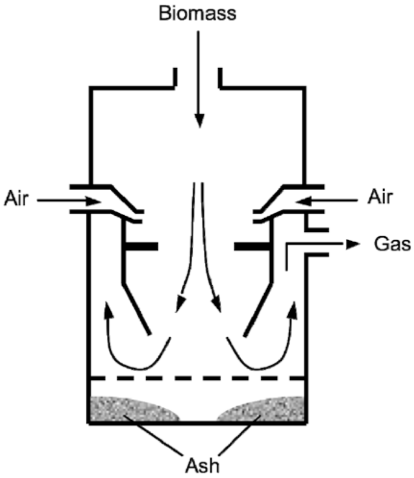

FIGURE 6.6 Constricted downdraft gasifier (Imbert type). Air/oxygen is added through nozzles around the vessel just above the constriction. *(Source: Adapted from Reed and Das, 1988, p. 39.)*

The bottommost layer (zone IV) consists of hot ash and/or unreacted charcoal, which crack any unconverted tar in this layer. Figure 6.5 shows the reactions and temperature distribution along the gasifier height. In one version of the throatless downdraft gasifier, the *open-core* type, the air enters from the top along with the feed. This type is free from some of the problems of other downdraft gasifiers.

Throated Gasifier

The cross-sectional area of a throated (also called *constricted*) gasifier is reduced at the throat and then expanded, as shown in Figure 6.4. The purpose is for the oxidation (combustion) zone to be at the narrowest part of the throat and to force all of the pyrolysis gas to pass through this narrow passage. Air is injected through nozzles just above the constriction. The height of the injection is about one-third of the way up from the bottom (Reed and Das, 1988, p. 33).

The movement of the entire mass of pyrolysis product through this hot and narrow zone results in a uniform temperature distribution over the cross-section and allows most of the tar to crack there. In the 1920s, a French inventor, Georges Imbert, developed the original design, which is popularly known as an *Imbert gasifier* (Figure 6.6).

The fuel, fed at the top, descends along a cylindrical section that serves as storage. The air pyrolyzes the biomass and burns the pyrolysis product or some charcoal. The hot char and the pyrolysis product pass through the throat, where most of the tar is cracked and the char is gasified. Figure 6.6 showed a flat-type throat construction, but it can be a V-type like in Figure 6.4.

Throated downdraft gasifiers are not suitable for scale-up to larger sizes because they do not allow for uniform distribution of flow and temperature in the constricted area. Beyond 1 MWth capacity, an annular constriction can be employed, but this has not been the practice to date.

6.2.3 Crossdraft Gasifiers

A crossdraft gasifier is a co-current moving-bed reactor, in which the fuel is fed from the top and air is injected through a nozzle from the side (Figure 6.7). It is primarily used for gasification of charcoal with very low ash content. Unlike the downdraft and updraft types, it releases the product from its side wall opposite to the entry point of the air for gasification. Because of this configuration, the design is also referred to as *sidedraft*. High-velocity air enters the gasifier through a nozzle set at a certain height above the grate. Excess oxygen in front of the nozzles facilitates combustion (oxidation) of part of the char, creating a very-high-temperature (>1500 °C) zone. The remaining char is then gasified to CO downstream in the next zone (Figure 6.7). The product gas exits from the opposite side of the gasifier. Heat from the combustion zone is conducted around the pyrolysis zone, so the fresh biomass is pyrolyzed while passing through it.

This type of gasifier is generally used in small-scale biomass units. One of its important features is a relatively small reaction zone with low thermal capacity, which gives a faster response time than that of any other moving-bed type. Moreover, startup time (5–10 minutes) is much shorter than in downdraft and updraft units. These features allow a sidedraft gasifier to respond well to load changes when used directly to run an engine. Because its tar production is low (0.01–0.1 g/nm^3), a crossdraft gasifier requires a relatively simple gas-cleaning system.

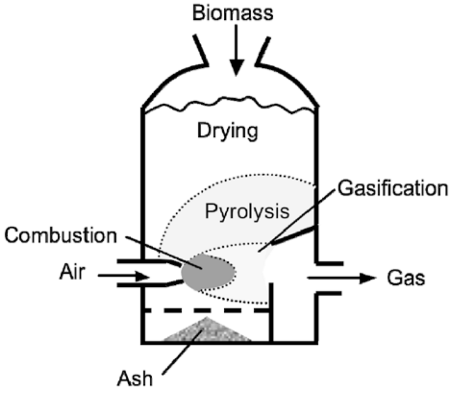

FIGURE 6.7 Schematic of a crossdraft gasifier.

Crossdraft gasifiers can be very light and small (<10 kWe). Since layers of fuel and ash insulate the walls from the high-temperature zone, the gasifier vessel can be constructed of ordinary steel with refractory linings on the nozzle and gas exit zone.

The crossdraft design is less suitable for high-ash or high-tar fuels, but it can handle high-moisture fuels if the top is open so that the moisture can escape. Particle size should be controlled, as unscreened fuel runs the risk of bridging and channeling. Crossdraft gasifiers work better with charcoal or pyrolyzed fuels. For unpyrolyzed fuels, the height of the air nozzle above the grate becomes critical (Reed and Das, 1988, p. 32).

6.3 FLUIDIZED-BED GASIFIERS

Fluidized-bed gasifiers are noted for their excellent mixing and temperature uniformity. A fluidized bed is made of granular solids, called *bed materials*, that are kept in a semi-suspended condition (*fluidized state*) by the passage of the gasifying medium through them at the appropriate velocities. The excellent gas–solid mixing and the large thermal inertia of the bed make this type of gasifier relatively insensitive to the fuel's quality (Basu, 2006). Along with this, the temperature uniformity greatly reduces the risk of fuel agglomeration.

The fluidized-bed design has proved to be particularly advantageous for gasification of biomass. Its tar production lies between that for updraft (~50 g/nm^3) and downdraft gasifiers (~1 g/nm^3), with an average value of around 10 g/nm^3 (Milne et al., 1998, p. 14). There are two principal fluidized-bed types: bubbling and circulating.

6.3.1 Bubbling Fluidized-Bed Gasifier

The bubbling fluidized-bed gasifier, developed by Fritz Winkler in 1921, is perhaps the oldest commercial application of fluidized beds; it has been in commercial use for many years for the gasification of coal (Figure 6.8); for biomass gasification, it is one of the most popular options. A fairly large number of bubbling fluidized-bed gasifiers of varying designs have been developed and are in operation (Lim and Alimuddin, 2008; Narváez et al., 1996).

Because they are particularly suitable for medium-size units (<25 MWth), many biomass gasifiers operate on the bubbling fluidized-bed regime. Depending on operating conditions, bubbling-bed gasifiers can be grouped as low-temperature and high-temperature types. They can also operate at atmospheric or elevated pressures.

In the most common type of fluidized bed, biomass crushed to less than 10 mm is fed into a bed of hot materials. These bed materials are fluidized with steam, air, or oxygen, or their combination, depending on the choice of

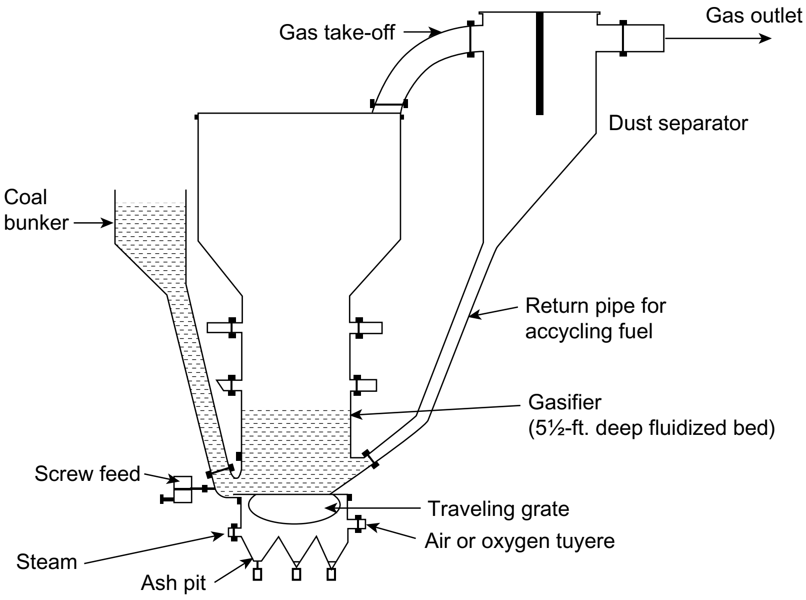

FIGURE 6.8 A sketch of the original Winkler bubbling fluidized-bed gasifier.

gasification medium. The ash generated from either the fuel or the inorganic materials associated with it is drained easily from the bottom of the bed. The bed temperature is normally kept below 980 °C for coal and below 900 °C for biomass to avoid ash fusion and consequent agglomeration.

The gasifying medium may be supplied in two stages. The first-stage supply is adequate to maintain the fluidized bed at the desired temperature; the second-stage supply, added above the bed, converts entrained unreacted char particles and hydrocarbons into useful gas.

High-temperature Winkler (HTW) gasification is an example of high-temperature, high-pressure bubbling fluidized-bed gasification for coal and lignite. Developed by Rheinbraun AG of Germany, the process employs a pressurized fluidized bed operating below the ash-melting point. To improve carbon conversion efficiency, small char particles in the raw gas are separated by a cyclone and returned to the bottom of the main reactor (Figure 6.9).

The gasifying medium (steam and oxygen) is introduced into the fluidized bed at different levels as well as above it. The bed is maintained at a pressure of 10 bars while its temperature is maintained at about 800 °C to avoid ash fusion. The overbed supply of the gasifying medium raises the local temperature to about 1000 °C to minimize production of methane and other hydrocarbons.

The HTW process produces a better-quality gas compared with the gas that is produced by traditional low-temperature fluidized beds. Though originally

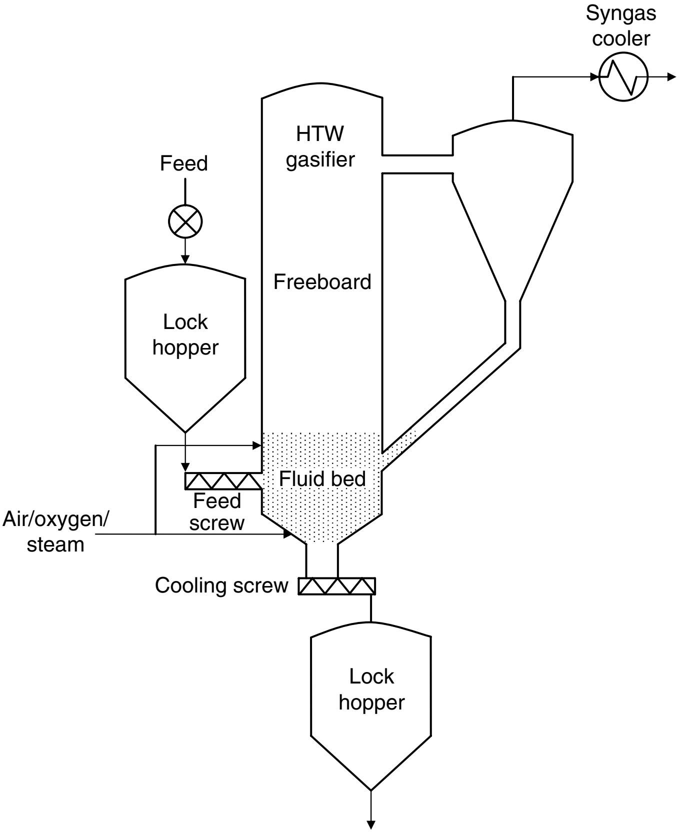

FIGURE 6.9 High-temperature Winkler (HTW) bubbling fluidized-bed gasifier.

developed for coal, it is suitable for lignite and other reactive fuels like biomass and treated municipal solid waste (MSW).

6.3.2 Circulating Fluidized-Bed Gasifier

A circulating fluidized-bed (CFB) gasifier has a special appeal for biomass gasification because of the long gas residence time it provides. It is especially suitable for fuels with high volatiles. A CFB typically comprises a riser, a cyclone, and a solid recycle device (Figure 6.10). The riser serves as the gasifier reactor.

Although the HTW process (Figure 6.9) appears similar to a CFB, it is only a bubbling bed with limited solid recycle. The circulating and bubbling fluidized beds are significantly different in their hydrodynamic. In a CFB, the solids are dispersed all over the tall riser, allowing a long residence time for the gas as well as for the fine particles. The fluidization velocity in a CFB is much

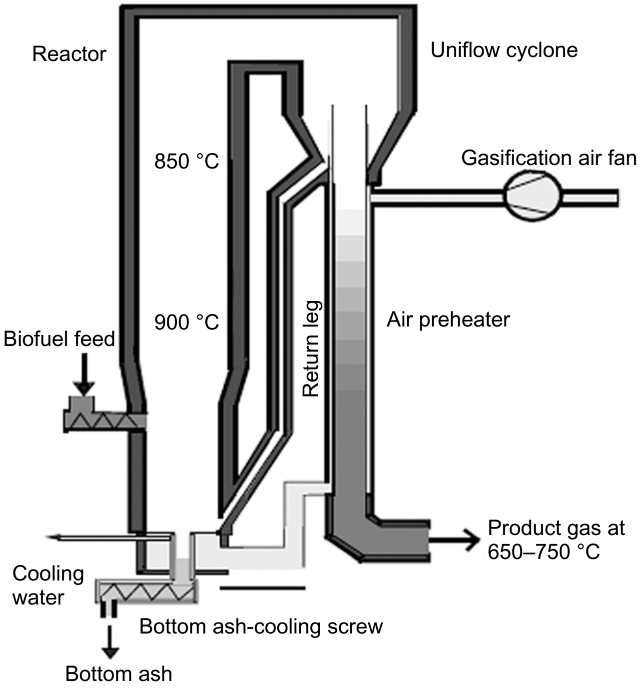

FIGURE 6.10 Circulating fluidized-bed gasifier. *(Source: Adapted from Foster Wheeler.)*

higher (3.5–5.5 m/s) than that in a bubbling bed (0.5–1.0 m/s). Also, there is large-scale migration of solids out of the CFB riser. These are captured and continuously returned to the riser's base. The recycle rate of the solids and the fluidization velocity in the riser are sufficiently high to maintain the riser in a special hydrodynamic condition, known as *fast fluidized bed*. Depending on the fuel and the application, the riser operates at a temperature of 800 to 1000 °C.

The hot gas from the gasifier passes through a cyclone, which separates most of the solid particles associated with it, and the loop seal returns the particles to the bottom of the gasifier. Foster Wheeler developed a CFB gasifier where an air preheater is located in the standpipe below the cyclone to raise the temperature of the gasification air and indirectly raise the gasifier temperature (Figure 6.10).

Many commercial gasifiers of this type have been installed in different countries. At the time of writing, the biggest among these is a 60-MWth unit in a coal-fired and natural-gas-fired power plant in Lahti, Finland, to provide a cheap supplementary fuel by gasifying waste wood and refuse-derived fuels (RDFs). Several manufacturers around the world have developed versions of the CFB gasifier that work on the same principle and vary only in engineering details.

TABLE 6.3 Comparison of Hydrodynamic Operating Conditions of Commercial Transport Gasifier and Circulating Fluidized Bed of Fluid Catalyst Cracking Units

Parameter	Smith et al., 2002	Peterson and Werther, 2005	Zhu and Venderbosch, 2005
Particle size (μm)	200–350	180–230	20–150
Riser velocity (m/s)	12–18	3.5–5.0	6–28
Circulation rate (kg/m^2·s)	730–3400	2.5–9.2*	400–1200
Riser temperature (°C)	910–1010	800–900	500–550
Riser pressure (bar)	140–270 psig	1 bar	150–300 kPa
Reactor	KBR gasifier	CFB gasifier	FCC cracker

*Computed from comparable units.

Transport Gasifier

This type of gasifier has the characteristics of both entrained-flow and fluidized-bed reactors. The hydrodynamics of a transport gasifier is similar to that of a fluid catalytic cracking reactor. A transport gasifier operates at circulation rates, velocities, and riser densities considerably higher than those of a conventional circulating fluidized bed. This results in higher throughput, better mixing, and higher mass and heat-transfer rates. The fuel particles are also very fine (Basu, 2006) and as such it requires a pulverizer or a hammer mill. A comparison of typical hydrodynamic operating conditions in a transport gasifier and in a fluid catalytic cracking unit is given in Table 6.3.

A transport gasifier consists of a mixing zone, a riser, a disengager, a cyclone, a standpipe, and a J-leg. Coal, sorbent (for sulfur capture), and air are injected into the reactor's mixing zone. The disengager removes the larger carried-over particles, and the separated solids return to the mixing section through the J-valve located at the base of the standpipe (Figure 6.11). Most of the remaining finer particles are removed by a cyclone located downstream, from which the gas exits the reactor. The reactor can use either air or oxygen as the gasification medium.

Use of oxygen as the gasifying medium avoids nitrogen, the diluting agent in the product gas. For this purpose, air is more suitable for power generation, while oxygen is more suitable for chemicals production. The transport gasifier has proved to be effective for gasification of coal, but it is yet to be proven for biomass.

Twin Reactor System

One of the major problems in air gasification of coal or biomass is the dilution of product gas by the nitrogen in the air used for the exothermic combustion

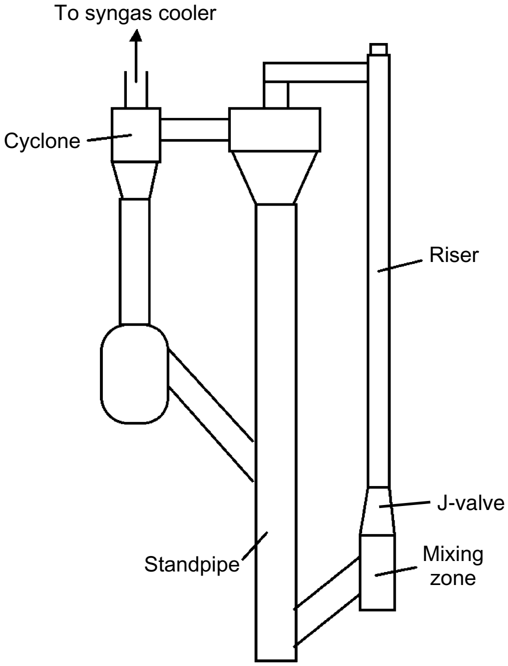

FIGURE 6.11 A sketch of a typical transport fluidized-bed gasifier.

reaction necessary in a self-sustained gasifier. To avoid this, oxygen is used instead, but oxygen gasification is expensive and highly energy intensive (see Example 6.5 later in chapter). A twin reactor (e.g., a dual fluidized bed) overcomes this problem by separating the combustion reactor from the gasification reactor such that the nitrogen released in the air combustion does not dilute the product gas. Twin reactor systems are used for coal and biomass. They are either externally circulating or internally circulating.

This type of system has some limitations; for example, Corella et al. (2007) identified two major design issues with the dual fluidized-bed system:

- Biomass contains less char than coal contains; however, if this char is used for gasification the amount of char available may not be sufficient to provide the required endothermic heat to the gasifier reactor to maintain a temperature above 900 °C. External heating may be necessary.
- Though the gasifier runs on steam, only a small fraction (<10%) of the steam participates in the gasification reaction; the rest of it simply leaves the gasifier, consuming a large amount of heat and diluting the gas.

The Technical University of Vienna used the externally circulating system to gasify various types of biomass in an industrial plant in Gussing, Austria.

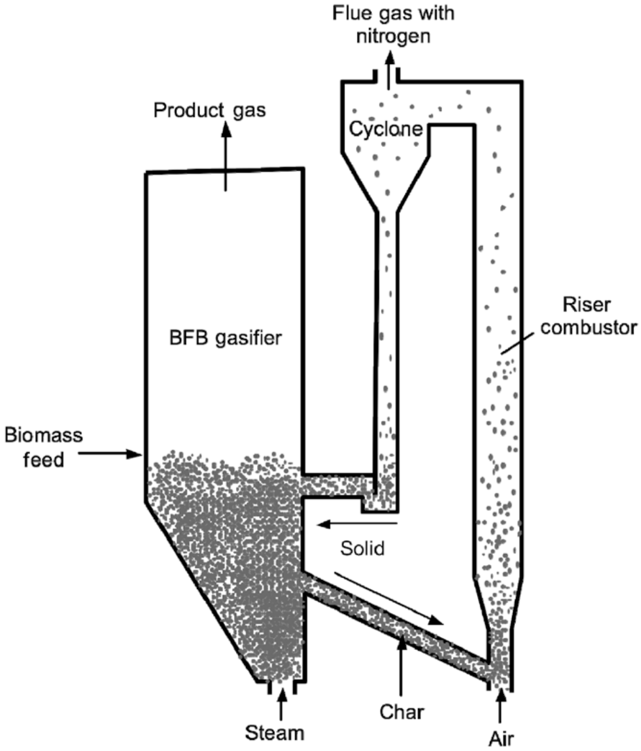

FIGURE 6.12 Twin reactor (dual fluidized-bed) gasifier.

The system is comprised of a bubbling fluidized-bed gasifier and a circulating fluidized-bed combustor (Figure 6.12). The riser in a CFB operates as a combustor; the bubbling fluidized bed in the return leg operates as a gasifier. Pyrolysis and gasification take place in the bubbling fluidized bed, which is fluidized by superheated steam. Unconverted char and tar move to the riser through a nonmechanical valve. The riser is fluidized by air.

Tar and gas produced during pyrolysis are combusted in the riser's combustion zone. Heat generated by combustion raises the temperature of the inert bed material to around 900 °C. This material leaves the riser and is captured by the cyclone at the riser exit. The collected solids drop into a standpipe and are then circulated into the bubbling fluidized-bed reactor to supply heat for its endothermic reactions. The char is gasified in the bubbling bed in the presence of steam, producing the product gas. This system overcomes the problem of tar by burning it in the combustor. In this way, a product gas relatively free of tar can be obtained.

The Rentech-Silvagas process is also based on the externally circulating principle. Here, both the combustor and the gasifier work on circulating fluidized-bed principles. In the internally circulating design, the fluidized-bed

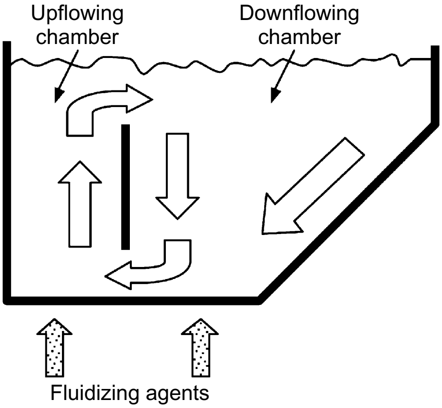

FIGURE 6.13 Internally circulating dual fluidized-bed gasifier.

reactor is divided into two chambers and connected by a window at the bottom of the division wall separating them. The chambers are fluidized at different velocities (Figure 6.13), which result in their having varying bed densities. As the bed height is the same in both, the hydrostatic pressure at the bottom of the two chambers is different. The biomass and sand thus flow from the higher-density chamber to the lower-density chamber, creating a continuous circulation of bed materials similar to the natural circulation in a boiler. This helps increase the residence time of solids in the fluidized bed.

Such an arrangement can provide a more uniform distribution of biomass particles in the reactor, with increased gasification yield and decreased tar and fine solids (char) in the syngas (Freda et al., 2008). A special feature of the twin reactor is that more air or oxygen can be added in one part of the bed to encourage combustion, and more steam can be added in another part to encourage gasification.

Chemical Looping Gasifier

Chemical looping is a relatively new concept. Its primary motivation is production of two separate streams of gases—a product gas rich in hydrogen and a gas stream rich in carbon dioxide—such that the CO_2 can be sequestrated while the hydrogen can be used for applications that require hydrogen-rich gas. The system uses calcium oxide as a carrier of carbon dioxide between two reactors: a gasifier (bubbling fluidized bed) and a regenerator (circulating fluidized bed). The CO_2 produced during gasification is captured by the CaO and released in a second reactor during sorbent regeneration.

Figure 6.14 is a schematic of the chemical looping process. Biomass is fed into the gasifier that receives calcium oxide from the regenerator and superheated steam from an external source. During gasification, the carbon dioxide produced is captured by the calcium oxide that makes up the bubbling fluidized bed (Acharya et al., 2009), as follows:

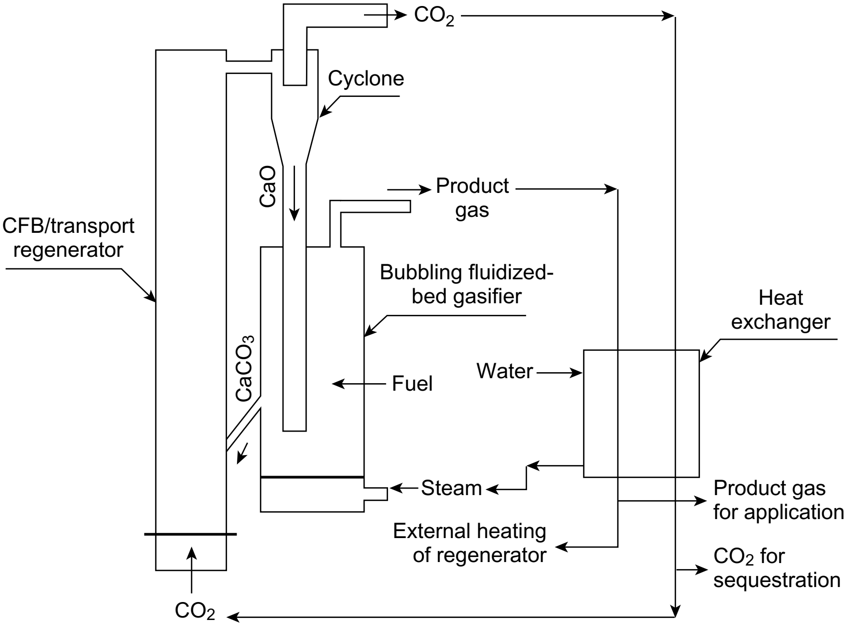

FIGURE 6.14 Chemical looping gasification with CaO as the carrier of CO_2 between the gasifier and the regenerator.

$$\text{Gasification reaction: } C_nH_hO_o + (2n-p)H_2O + n\text{CaO} \leftrightarrow n\text{CaCO}_3 + \left(\frac{h}{2} + 2n - o\right)H_2 \tag{6.1}$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 \tag{6.2}$$

$$CO_2 \text{ removal reaction: CaO} + CO_2 \rightarrow \text{CaCO}_3 \tag{6.3}$$

Immediate removal of the reaction product, CO_2, from the system increases the rate of forward reaction (Eq. 6.2), enhancing the water–gas shift reaction, therefore yielding more hydrogen in the product gas. The calcium carbonate formed in the gasifier (Eq. 6.3) is transferred to a circulating/transport regenerator, where it is calcined into calcium oxide and carbon dioxide.

$$\text{Regeneration: CaCO}_3 \rightarrow \text{CaO} + CO_2 + 178.3 \text{ kJ/mol} \tag{6.4}$$

The carbon dioxide and the product gas leave the regenerator and gasifier, respectively, at a high temperature. The hot product can be used for generation of steam needed for gasification.

6.4 ENTRAINED-FLOW GASIFIERS

Entrained flow is the most successful and widely used gasifier type for large-scale gasification of coal, petroleum coke, and refinery residues. It is ideally suited to most types of coal except low-rank coal, which, like lignite and biomass, is not attractive because of its large moisture content. High-ash coal is also less suitable because cold-gas efficiency decreases with increasing ash content. For slurry-fed coal, the economic limit is 20% ash; for dry feed it is 40% (Higman and Burgt, 2008, p. 122).

The suitability of entrained-flow gasification for biomass is questionable for a number of reasons. Owing to a short residence time (a few seconds) in entrained-flow reactors, the fuel needs to be very fine, and grinding fibrous biomass into such fine particles is difficult. For biomass with CaO but no alkali, the ash-melting point is high, and therefore it has a higher oxygen requirement. The melting point of biomass ash with a high alkali content is much lower than that of coal. This reduces the oxygen required to raise the temperature of the ash above its melting point. However, molten biomass ash is highly aggressive, which greatly shortens the life of the gasifier's refractory lining.

For these reasons entrained-flow reactors are not preferred for biomass gasification. Still, they have the advantage of easily destroying tar, which is very high in biomass and is a major problem in biomass gasification.

Entrained-flow gasifiers are essentially co-current plug-flow reactors, where gas and fuel travel. The hydrodynamics is similar to that of the well-known pulverized-coal (PC) boiler, where the coal is ground in a pulverizing mill to sizes below 75 micron and then conveyed by part of the combustion air to a set of burners suitably located around the furnace. The reactor geometry of the entrained-flow gasifier is much different from the furnace geometry of a PC boiler. Additionally, an entrained-flow gasifier works in a substoichiometric supply of oxygen, whereas a PC boiler requires excess oxygen.

The gasification temperature of an entrained-flow gasifier generally well exceeds 1000 °C. This allows production of a gas that is nearly tar-free and has a very low methane content. A properly designed and operated entrained-flow gasifier can have a carbon conversion rate close to 100%. The product gas, being very hot, must be cooled in downstream heat exchangers that produce the superheated steam required for gasification.

Figure 6.15 describes the working principle of an entrained-flow gasifier by means of a simplified sketch. The high-velocity jet forms a recirculation zone near the entry point. Fine fuel particles are rapidly heated by radiative heat from the hot walls of the reactor chamber and from the hot gases downstream, and start burning in excess oxygen. The bulk of the fuel is consumed near the entrance zone through devolatilization; here the temperature may rise to as high as 2500 °C.

The combustion reaction consumes nearly all of the oxygen feed, so the residual char undergoes gasification reactions in CO_2 and H_2O environments downstream of this zone. These reactions are relatively slow compared to the devolatilization reaction, so the char takes much longer to complete its conversion to gases. For this reason, a large reactor length is required.

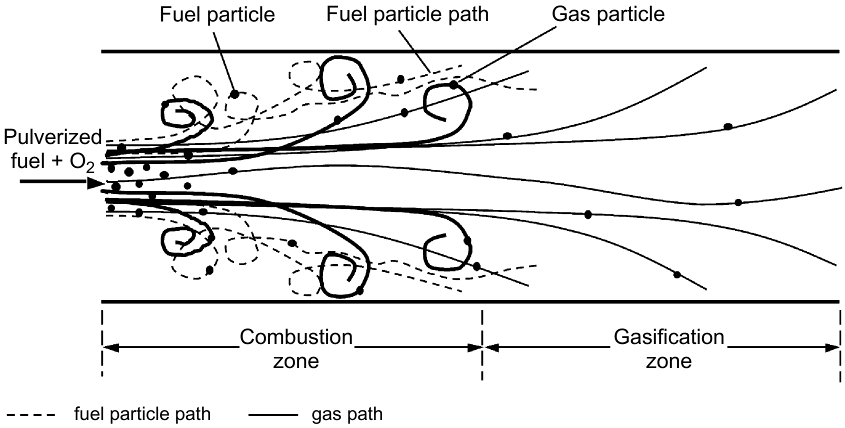

FIGURE 6.15 Simplified sketch of gas–solid flow in an entrained-flow gasifier.

Entrained-flow gasifier design may be classified into two broad groups: (1) the top-fed downflow (used by GE Energy and Siemens SFG), shown in Figure 6.16; and (2) the side-fed upflow (used by Koppers-Totzek, the Shell gasification process, Prenflo, and the Lurgi multipurpose), shown in Figure 6.17.

6.4.1 Top-Fed Gasifier

Top-fed gasifiers use a vertically cylindrical reactor vessel, into which pulverized fuel (biomass or coal) and gasifying agent(s) are conveyed by oxygen and injected from the top. This vessel resembles a vertical furnace with a downward burner (Figure 6.16). The fuel and the gasifying agent(s) are injected into the reactor through a jet that generally sits at the reactor's middle section.

The fuel gasifier (SFG) process of Siemens uses a top-fired reactor design, in which the reactants are introduced through the single centrally mounted burner. This has several advantages. First, it is of an axisymmetrical construction, reducing equipment costs; second, the flow of reactant occurs from a single burner, reducing the number of burners to be controlled; finally, the product gas and the slag flow in the same direction, which reduces any potential blockage in a slag trap (Higman and van der Burgt, 2008, p. 132).

6.4.2 Side-Fed Gasifier

In side-fed gasifiers, powdered fuel is injected through horizontal nozzles set opposite each other in the reactor's lower section (Figure 6.17). Jets of fuel and gasifying agents form a stirred-tank reactor characterized by a high degree of mixing. The product gas moves upward and exits through the top. Because of the high oxygen availability in this mixing zone, rapid exothermic reactions take place, raising the gas temperature to well above the ash-melting point

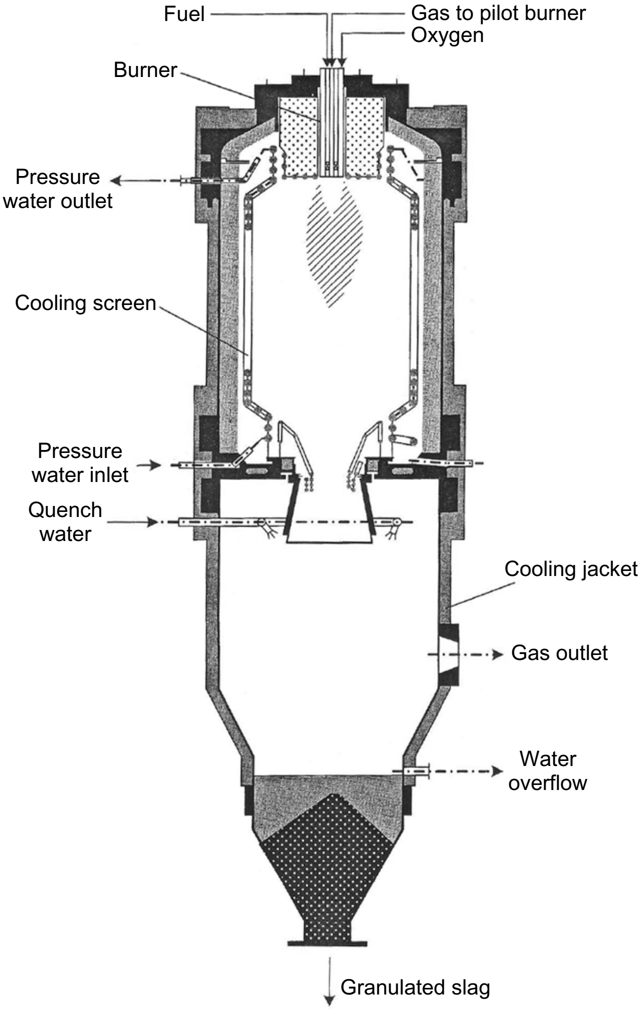

FIGURE 6.16 A schematic of a top-fed downflow entrained-flow gasifier.

(>1400 °C). Thus, the ash, instead of traveling up, is separated in this zone as slag from the fuel, and drained. Some gasifier designs (e.g., E-gas, MHI, Eagle) inject additional fuel further downstream from the main reaction zone.

The Koppers-Totzek atmospheric pressure gasifier also uses side feeding. It consists of two side-mounted burners where a mixture of coal and oxygen is injected. The gas leaves from the top of the gasifier at temperatures around 1500 °C and is quenched with water downstream. The reactor has a steam jacket to protect its shell from high temperatures (Higman and van der Burgt, 2008, p. 129).

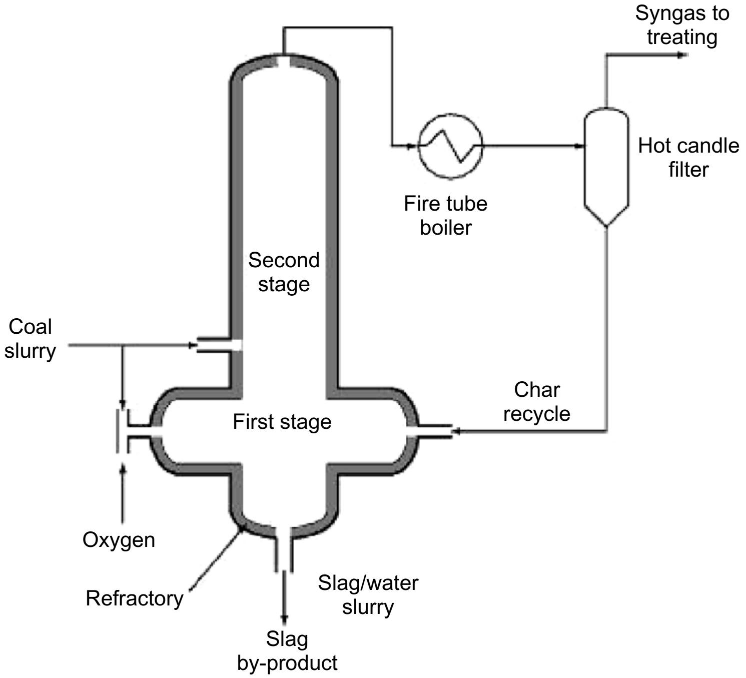

FIGURE 6.17 A schematic of a side-fed entrained-flow gasifier.

The E-gas gasifier is a side-fed two-stage entrained-flow slagging gasifier with a coal–water slurry feed. It is designed to use sub-bituminous coal (Figure 6.17). The coal slurry is fed at the nonslagging stage, where the upflowing gas heats it. Thus, the gas exits at a lower temperature and then passes through a fire-tube boiler and is filtered in a hot candle filter. The char, separated out by the filter, is taken back to the slagging zone. The slag is quenched in a water bath at the bottom of the slagging reactor.

6.4.3 Advantages of Entrained-Flow Gasifiers

Entrained-flow gasifiers have several advantages over other types:

- Low tar production
- A range of acceptable feed
- Ash produced as slag
- High-pressure, high-temperature operation
- Very high conversion of carbon
- Low methane content well suited for synthetic gas production

6.4.4 Entrained-Flow Gasification of Biomass

For thermal gasification of the refractory components of biomass (those difficult to gasify) such as lignin, the minimum temperature requirement is similar to that for coal (~900 °C) (Higman and van de Burgt, 2008, p. 147). Entrained-flow gasification of biomass is therefore rather limited and has not been seen on a commercial scale for the following reasons:

- The residence time in the reactor is very short. For the reactions to complete, the biomass particles must be finely ground. Being fibrous, biomass cannot be pulverized easily.
- Molten ash from biomass is highly aggressive because of its alkali compounds and can corrode the gasifier's refractory or metal lining.

Given these shortcomings, entrained-flow gasifiers are not popular for biomass. However, there is at least one successful entrained-flow biomass gasifier, known as the Choren process.

Choren Process

The Choren entrained-flow biomass gasifier is comprised of three stages (Figure 6.18). The first stage receives biomass in a horizontal stirred-type low-temperature reactor for pregasification at 400 to 500 °C in a limited supply of air. This produces solid char and a tar-rich volatile product. The latter flows into the second chamber (stage 2), an entrained-flow combustor where oxygen and the product gas from the first stage are injected downward into the reactor. Combustion raises the temperature to 1300 to 1500 °C and completely cracks the tar. The hot combustion product flows into the third chamber (stage 3), where the char is gasified.

The solid char received from the first stage is pulverized and fed into the third stage of the Choren process. It is gasified in the hot gasification medium produced in the second stage. Endothermic gasification reactions reduce the

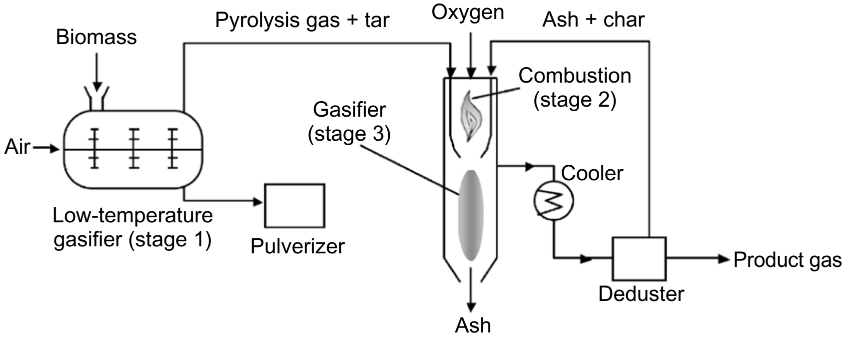

FIGURE 6.18 Choren process. The biomass is gasified in an entrained-flow gasifier, facilitated by a rotary-type partial gasifier (stage 1).

temperature to about 800 °C. Char and ash from the product gas are separated and recycled into the second-stage combustor. The ash melts at the high temperature in the combustor and is drained from the bottom. Now molten, the ash freezes, forming a layer on the membrane wall that protects the wall against the corrosive action of fresh molten biomass ash. The product gas is processed downstream for Fisher-Tropsch synthesis or other applications.

6.5 PLASMA GASIFICATION

In plasma gasification, high-temperature plasma helps gasify biomass hydrocarbons. It is especially suitable for MSW and other waste products. This process may also be called "plasma pyrolysis" because it essentially involves thermal disintegration of carbonaceous material into fragments of compounds in an oxygen-starved environment. The heart of the process is a plasma gun, where an intense electric arc is created between two electrodes spaced apart in a closed vessel through which an inert gas is passed (Figure 6.19).

Though the temperature of the arc is extremely high (~13,000 °C), the temperature downstream, where waste products are brought in contact with it, is much lower (2700–4500 °C). The downstream temperature is still sufficiently high, however, to pyrolyze complex hydrocarbons into simple gases such as CO and H_2. Simultaneously, all inorganic components (e.g., glass, metals,

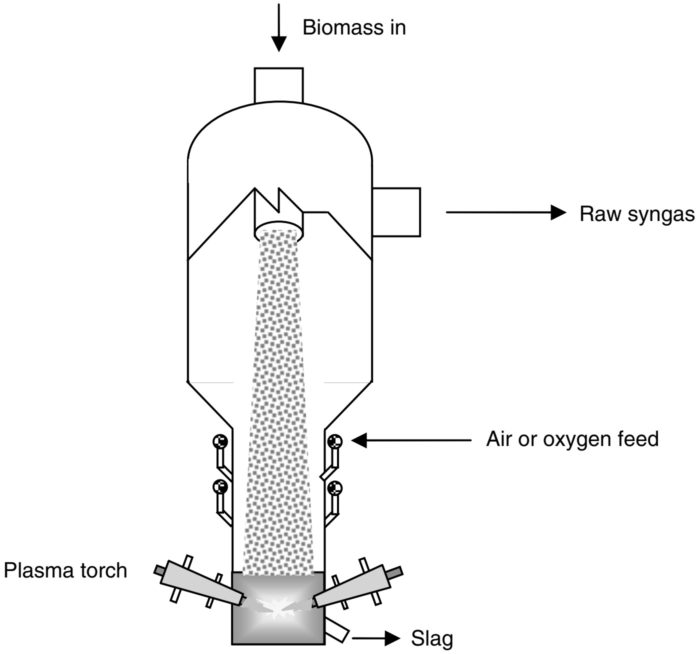

FIGURE 6.19 Plasma gasification of solid waste.

silicates, heavy metals) are fused into a volcanic-type lava, which after cooling forms a basaltic slag. The product gas leaves the gasifier at very high temperatures (1000–1200 °C).

A typical plasma reactor provides a relatively long residence time for the gas in the gasifier. This and the high temperature cause the tar products to be cracked and harmful products like dioxin and furan to be destroyed.

Owing to the high reactor temperature and the presence of chlorine in wastes, the life of the reactor liner is an issue. However, an attractive feature is that plasma gasification is relatively insensitive to the quality of the feedstock. This is the result of an independent energy source run by electricity instead of partial combustion of the gasification product.

6.6 PROCESS DESIGN

The design of a gasifier involves both process and hardware. The process design gives the type and yield of the product, operating conditions, and the basic size of the reactor. The hardware design involves structural and mechanical components, such as grate, main reactor body, insulation, cyclone, and others, that are specific to the reactor type. This section focuses on gasifier process design.

6.6.1 Design Specification

For any design, specification of the plant is very important. The input includes the specification of the fuel, gasification medium, and product gas. A typical fuel specification will include proximate and ultimate analysis, operating temperatures, and ash properties. The specification of the gasifying medium is based on the selection of steam, oxygen, and/or air and their proportions.

These parameters could influence the design of the gasifier, as follows:

- The desired heating value of the product gas dictates the choice of gasification medium. Table 6.4 gives typical ranges of heating value for different mediums.

TABLE 6.4 LHV of Product Gas Ranges and Choice of Gasifying Medium

Gasification Medium	Range of Heating Value of Product Gas (MJ/Nm^3)
Air gasification	4–7
Steam gasification	10–18
Oxygen gasification	12–28

- Hydrogen can be maximized with steam, but if it is not a priority, oxygen or air is a better option, as it reduces the energy used in generating steam and the energy lost through unutilized steam.
- If nitrogen in the product gas is not acceptable, air cannot be chosen.
- Capital cost is lower for air, followed by steam. A much larger investment is needed for an oxygen plant, which also consumes a large amount of auxiliary power.
- Equivalence ratio

For the product gas, the specification includes:

- Desired gas composition
- Desired heating value
- Desired production rate (Nm^3/s or MWth produced)
- Yield of the product gas per unit fuel consumed
- Required power output of the gasifier, Q

The design outputs of process design include geometry and operating and performance parameters.

Basic size includes reactor configuration, cross-section area, and height (hardware design). Important operating parameters are: (1) reactor temperature; (2) preheat temperature of the steam, air, or oxygen; and (3) amount (i.e., steam/biomass ratio) and relative proportion of the gasifying medium (i.e., steam/oxygen ratio). Performance parameters of a gasifier include carbon conversion and cold-gas efficiency.

A typical process design starts with a mass balance followed by an energy balance. The following subsections describe the calculation procedures for these.

6.6.2 Mass Balance

Basic mass and energy balance is common to all types of gasifiers. It involves calculations for product gas flow and fuel feed rate.

Product Gas Flow Rate

The gasifier's required power output, Q (MWth), is an important input parameter specified by the client. Based on this, the designer makes a preliminary estimation of the amount of fuel to be fed into the gasifier and the amount of gasifying medium. The volume flow rate of the product gas, V_g (Nm^3/s), from its desired lower heating value, LHV_g (MJ/Nm^3), is found by

$$V_g = \frac{Q}{LHV_g} \mathrm{Nm}^3/\mathrm{s} \tag{6.5}$$

The net heating value or lower heating value (LHV) can be calculated from its composition. The composition may be predicted by the equilibrium calculations, described later, or by sophisticated kinetic modeling of the gasifier, as discussed in Chapter 5. In the absence of these, a reasonable guess can be made. either from published data on similar fuels in similar gasification conditions or from the designer's experience.

For example, for air-blown fluidized-bed biomass gasifiers, the LHV is in the range 3.5 to 6 MJ/Nm^3 (Enden and Lora, 2004). For oxygen gasification, it is in the range 10 to 15 MJ/Nm^3 (Ciferno and Marano, 2002). So, for an air-blown gasifier, we start with a value of 5 MJ/Nm^3 as a reasonable guess (Quaak et al., 1999).

Fuel Feed Rate

To find the biomass feed rate, M_f , the required power output is divided by the LHV of the biomass (LHV_{bm}) and by the gasifier efficiency, η_{gef}.

$$M_f = \frac{Q}{LHV_{bm}\eta_{gef}} \tag{6.6}$$

The LHV may be related to the higher heating value (HHV) and its hydrogen and moisture contents (Quaak et al., 1999) as

$$LHV_{bm} = HHV_{daf} - 20{,}300 \times H_{daf} - 2{,}260 \times M_{daf} \tag{6.7}$$

Here, H_{daf} is the hydrogen mass fraction in the fuel, M_{daf} is the moisture mass fraction, and HHV_{daf} is the HHV in kJ/kg on a moisture-ash-free basis. By using the definition of these one can relate the HHV on moisture-ash-free basis to that on only dry-basis value as

$$HHV_{daf} = HHV_d\left(\frac{1-M}{1-ASH-M}\right) \tag{6.8}$$

where the subscripts d and daf refer to dry and moisture-ash-free basis, respectively; M is the moisture fraction; and ASH is the ash fraction in fuel on a raw-fuel basis.

On a dry basis, the HHV, HHV_d, is typically in the range 18 to 21 MJ/kg (Van Loo and Koppejan, 2003, p. 48). It may be calculated from the ultimate analysis for the biomass using the following equation (Van Loo and Koppejan, 2003, p. 29):

$$\begin{aligned} HHV_d = {} & 0.3491\,C + 1.1783\,H + 0.1005\,S - 0.0151\,N \\ & - 0.1034\,O - 0.0211\,ASH \end{aligned} \tag{6.9}$$

where C, H, S, N, O, and ASH are the mass fraction of carbon, hydrogen, sulfur, nitrogen, oxygen, and ash in the fuel on a dry basis.

Flow Rate of Gasifiying Medium

The amount of gasification medium has a major influence on yield and composition of the product gas. This section discusses methods for choosing that amount.

Air

The theoretical air requirement for complete combustion of a unit mass of a fuel, m_{th}, is an important parameter. It is known as the *stoichiometric air* requirement. Its calculation is shown in Eq. 2.34. For an air-blown gasifier operating, the amount of air required, M_a, for gasification of unit mass of biomass is found by multiplying it by another parameter ER:

$$M_a = m_{th} ER \tag{6.10}$$

Here, *ER* is the equivalence ratio.

For a fuel feed rate of M_f, the air requirement of the gasifier, M_{fa}, is

$$M_{fa} = m_{th} ER \cdot M_f \tag{6.11}$$

For a biomass gasifier, 0.25 may be taken as a first-guess value for the equivalence ratio, *ER*. A more detailed discussion of this is presented next.

Equivalence Ratio The equivalence ratio is an important gasifier design parameter. It is the ratio of the actual air–fuel ratio to the stoichiometric air–fuel ratio. This term is generally used for air-deficient situations, such as those found in a gasifier.

$$ER\,(<1.0)_{gasification} = \frac{\text{Actual air}}{\text{Stoichiometric air}} = EA\,(>1.0)_{combustion} \tag{6.12}$$

where *EA* is the excess air coefficient.

In a combustor, the amount of air supplied is determined by the stoichiometric (or theoretical) amount of air and its *excess air coefficient*. In a gasifier, the air supply is only a fraction of the stoichiometric amount. The stoichiometric amount of air may be calculated based on the ultimate analysis of the fuel.

The equivalence ratio, *ER*, dictates the performance of the gasifier. For example, pyrolysis takes place in the absence of air and hence the *ER* is zero; for gasification of biomass, it lies between 0.2 and 0.3.

Downdraft gasifiers give the best yield for *ER,* 0.25 (Reed and Das, 1988, p. 25). With a lower *ER* value, the char is not fully converted into gases. Some units deliberately operate with a low *ER* to maximize their charcoal production. A lower *ER* gives rise to higher tar production, however, so updraft gasifiers, which typically operate with an *ER* of less than 0.25, have higher tar content. With an *ER* above 0.25, some product gases are also burnt, increasing the temperature.

The quality of gas obtained from a gasifier strongly depends on the *ER* value, which must be significantly below 1.0 to ensure that the fuel is gasified rather than combusted. However, an excessively low *ER* value (<0.2) results in several problems, including incomplete gasification, excessive char formation, and a low heating value of the product gas. On the other hand, too high an *ER* (>0.4) results in excessive formation of products of complete combustion,

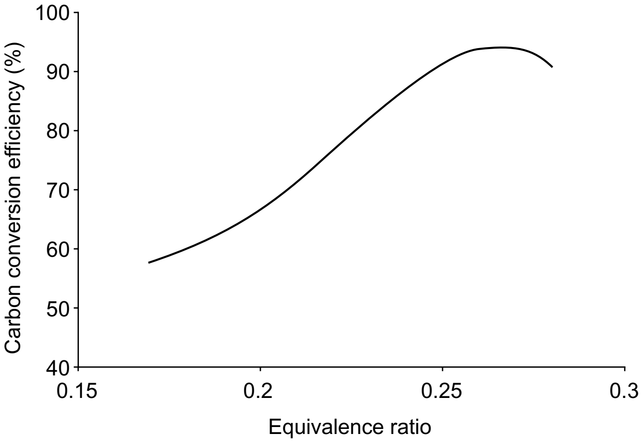

FIGURE 6.20 Effect of equivalence ratio on carbon conversion in a fluidized-bed gasifier.

such as CO_2 and H_2O, at the expense of desirable products, such as CO and H_2. This causes a decrease in the heating value of the gas. In practical gasification systems, the *ER* value is normally maintained within the range of 0.20 to 0.30. Figure 6.20 shows the variation in carbon conversion efficiency of a circulating fluidized-bed gasifier for wood dust against the equivalence ratio. The efficiency increases with ER and then it starts declining. The optimum value here is 0.26, but it may change depending on many factors.

The bed temperature of a fluidized-bed gasifier increases with the *ER* because the higher the amount of air, the greater the extent of the combustion reaction and the higher the amount of heat released (Figure 6.21). Example 6.1 illustrates the calculation procedure for *ER*.

Oxygen

Oxygen is used primarily to provide the thermal energy needed for the endothermic gasification reactions. The bulk of this heat is generated through the following partial and/or complete oxidation reactions of carbon:

$$C + 0.5\ O_2 \rightarrow CO - 111\ \text{kJ/mol} \tag{6.13}$$

$$C + O_2 \rightarrow CO_2 - 394\ \text{kJ/mol} \tag{6.14}$$

It can be seen that for the oxidation of 1 mol of carbon to CO_2, the oxygen requirement is (2 × 16)/12 = 2.66 mol, while that for carbon to CO is (16/12) = 1.33 mol. Thus, the reaction in Eq. (6.13) is more likely to take place in oxygen-deficient regions.

Besides supplying the energy for the endothermic gasification reactions, the gasifier must provide energy to raise the feed and gasification medium to the reaction temperature, as well as to compensate for the heat lost to the reactor walls. For a self-sustained gasifier, part of the chemical energy in the biomass

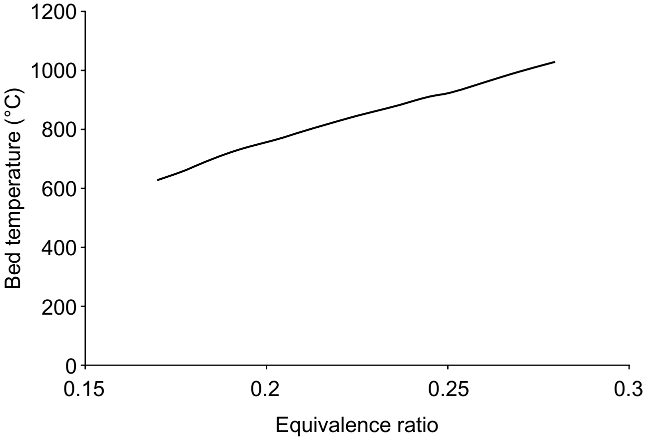

FIGURE 6.21 Gasifier temperature in a CFB riser increases with equivalence ratio.

provides the heat required. The total heat necessary comes from the oxidation reactions. The energy balance of the gasifier is thus the main consideration in determining the oxygen-to-carbon (O/C) ratio.

Equilibrium calculations can show that as the ratio of oxygen to carbon in the feed increases, CH_4, CO, and hydrogen in the product decreases but CO_2 and H_2O in the product increases. Beyond a ratio of 1.0, hardly any CH_4 is produced.

When air is the gasification medium, as is the case for 70% of all gasifiers (Ciferno and Marano, 2002), the nitrogen in it dilutes the product gas. The heating value of the gas is therefore relatively low (4–6 MJ/m^3). When pure oxygen from an air-separation unit is used, the heating value is higher, in the range 10 to 15 MJ/m^3, but a large amount of energy (~2.18 MJ/kg O_2) is spent in separating the oxygen from the air (Grezin and Zakharov, 1988).

The oxygen requirement of a gasifier can be met by either air supply or an air-separation unit that extracts oxygen from air.

Steam

Superheated steam as a gasification medium is used either alone, with air, or with oxygen. It contributes to the generation of hydrogen.

$$C + H_2O \rightarrow CO + H_2 \tag{6.15}$$

The quantity of steam, M_{fh}, is known from the steam-to-carbon (*S/C*) molar ratio.

$$\text{Steam flow rate, } M_{fh} = 18\frac{M_f C}{12}(S/C) \text{ kg steam/kg fuel} \tag{6.16}$$

where M_f is the fuel feed rate, and C is the carbon fraction in the fuel.

The *S/C* mole ratio has an important influence on product composition, as the *ER* has. Both hydrogen and CO increase with an increasing *S/C* ratio for a given temperature and oxygen-to-carbon molar ratio. The production of these two gases increases with decreasing pressure, decreasing oxygen, and decreasing *S/C* ratio. However, there is only a marginal gain in increasing the *S/C* molar ratio above 2 to 3, as the excess steam simply leaves the gasifier unreacted (Probstein and Hicks, 2006, p. 119). So a value in the range of 2.0 to 2.5 can give a reasonable starting value.

Example 6.1

A moving-bed gasifier 4 m in diameter operates at 25 bars of pressure and consumes 750 kg/min (dry-ash–free basis) of bituminous coal, 1930 kg/min of steam, and 280 Nm^3/min of oxygen to produce a product gas that contains 1000 Nm^3 of syngas (a mixture of H_2 and CO). The mean gasifier temperature is 1000 °C. The volumetric composition of the product gas is

CO_2–32%
H_2S–0.4%
CO–15.2%
H_2–42.3%
CH_4–8.6%
C_2H_4–0.8%
N_2–0.7%

The ultimate analysis of the coal on a moisture-ash-free basis is

C–77.3%
H–5.9%
S–4.3%
N–1.4%
O–11.1%

Find

- The *S/C* molar ratio
- The *O/C* molar ratio
- The *ER*
- The hearth load in energy produced per unit of grate area and space velocity

The heating values of the product gas constituents may be taken from Table C.2 in Appendix C.

Solution

From the feed rate of coal, steam, and oxygen, we can find the molar feed rate by dividing the mass rate by the molecular weight as here:

Carbon moles: 750 × 0.773/12 = 48.31 kmol/min
Steam moles: 1980/18 = 107.22 kmol/min
Oxygen moles: 280/22.4 = 12.5 kmol/min

From these we can calculate the following:

S/C molar ratio = 107.22/48.31 = **2.22**
O/C molar ratio = 12.5/48.31 = **0.26**

To find the stoichiometric oxygen requirement, the oxygen required to oxidize carbon to CO_2, hydrogen to H_2O, and sulfur to SO_2 has be to calculated.

- Twelve kg of carbon (1 mol) react with 32 kg of oxygen (1 mol) to produce 1 mol of CO_2:

$$\mathrm{C} + \mathrm{O_2} = \mathrm{CO_2}$$

Therefore, the oxygen required for 1 kg of carbon is 32/12.

- Thirty-two kg of sulfur (1 mol) react with 32 kg of oxygen (1 mol) to produce 1 mol of SO_2:

$$\mathrm{S} + \mathrm{O_2} = \mathrm{SO_2}$$

Therefore, the oxygen required for 1 kg of sulfur is 32/32 = 1.

- Similarly, 4 kg of hydrogen react with 32 kg of oxygen to produce H_2O:

$$2\mathrm{H_2} + \mathrm{O_2} = 2\mathrm{H_2O}$$

Therefore, the oxygen required for 1 kg of hydrogen is 32/4 = 8.

$$\text{Stoichiometric oxygen requirement} = \frac{32C}{12} + 8H + S - O = \frac{32 \times 0.773}{12} + 8 \times 0.059 + 0.043 - 0.111 = 2.465\ \text{kg of O}_2/\text{kg of fuel}$$

The total O_2 required is

$$750 \times 2.465 = 1848.75\ \text{kg of O}_2/\text{min}$$

The O_2 supplied is

$$\text{moles of O}_2 \times 32 = 12.5 \times 32 = 400\ \text{kg of O}_2/\text{min}$$

From this we can calculate

$$ER = 400/1848.75 = \mathbf{0.22}$$

The syngas constituents in the total product gas are CO (15.2%) and H_2 (42.3%). So, to produce 1000 Nm^3/min of syngas, the amount of product gas, Q_{pr}, is

$$Q_{pr} = 1000/(0.152 + 0.423) = \mathbf{1739\ Nm^3/min}$$

The cross-sectional area of the gasifier reactor, *A*, is

$$A = \pi\, 4^2/4 = 12.56\ \text{m}^2$$

Assuming the operating temperature to be 1000 °C and the pressure to be 25 bars, the volumetric flow rate of product gas is

$$Q'_{pr} = Q_{pr}\left(\frac{1}{25}\right)\left(\frac{1273}{273}\right) = 324\ \text{m}^3/\text{min}$$

The space velocity of the gas flow *Vg* is Q'_{pr}/A = 324/ (12.56 × 60) = **0.43 m/s**.

The energy produced per Nm^3 of product gas is found by multiplying the volume fraction by the heating value of each constituent, which is taken from Table C.2 in Appendix C. Adding together the contribution of all product gas constituents gives the total heating value, *HHV*, as

$$HHV = 0.004 \times 25.1 + 0.152 \times (282.99/22.4) + 0.423 \times (285.84/22.4)$$
$$+ 0.086 \times (890.36/22.4) + 0.008 \times 63.4 = 11.33\ \mathrm{MJ/Nm^3}$$

Thus, the total energy produced, E_{total}, is Q_{pr} x HHV

$$= 1739 \times 11.33/60$$
$$= \mathbf{328.3\ MWth}$$

The hearth load is

$$E_{total}/A = 328.3/12.56 = \mathbf{26.14\ MW/m^2}$$

6.6.3 Energy Balance

Unlike combustion reactions, most gasification reactions are endothermic. Thus, heat must be supplied to the gasifier for these reactions to take place at the designed temperature. In laboratory units, this is not an issue because the heat is generally supplied externally. In commercial units, it is a major issue, and it must be calculated and provided for. The amount of external heat supplied to the gasifier depends on the heat requirement of the endothermic reactions as well as on the gasification temperature. The latter is a design choice, and it is discussed next.

Gasification Temperature

Because lignin, a refractory component of biomass, does not gasify well at lower temperatures, thermal gasification of ligno-cellulosic biomass prefers a minimum gasification temperature in the range 800 to 900 °C. For biomass, an entrained-flow gasifier typically maintains a gasification temperature well exceeding 900 °C. For coal, the minimum is 900 °C for most gasifier types (Higman and van der Burgt, 2008, p. 163).

A higher peak gasification temperature is chosen for an entrained-flow gasifier. The higher the ash-melting temperature, the higher the design value of the gasifier temperature. This temperature is raised through the gasifier's exothermic oxidation reactions, so a high reaction temperature also means a high oxygen demand.

In entrained-flow gasifiers, the peak gasification temperature is typically in the range 1400 to 1700 °C, as it is necessary to melt the ash; however, the exit gas temperature is much lower. The peak temperature of a fluidized-bed gasifier is in the range of 700 to 900 °C to avoid softening of bed materials. It is about the same as the gas exit temperature in a fluidized-bed gasifier. In a crossdraft gasifier the gasification temperature is about 1250 °C, whereas the peak gasification temperature is about 1500 °C. The exit-gas temperature of a downdraft gasifier is about 700 °C, but its peak gasifier temperature at the throat is 1000 °C. The updraft gasifier has the lowest gas-exit temperature (200–400 °C), while its gasification temperature may be up to 900 °C (Knoef, 2005). Once the gasification temperature is known, the designer can turn to the heat balance on this basis.

Heat of Reaction

Heat of reaction is the heat gained or lost in a chemical reaction. To calculate it for gasification, we consider an overall gasification reaction where 1 mol of biomass ($C_aH_bO_c$) is gasified in α mols of steam and β mols of oxygen. The overall equation is

$$\begin{aligned} C_aH_bO_c + \alpha H_2O + \beta O_2 &= A' \cdot C + B' \cdot CO_2 + C' \cdot CO + D' \cdot CH_4 \\ &\quad + E' \cdot H_2O + F' \cdot H_2 + Q \end{aligned} \tag{6.17}$$

The equilibrium analysis of Section 5.5.2 gives the mole fraction A', B', C', D', E', and F' in the flue gas for given values of α and β. The chosen S/B ratio defines α, while the ER defines β. The heat of reaction, Q, for the overall gasification reaction (Eq. 6.22) may be found from the heat of formation of the products and reactants:

$$\begin{aligned} \text{Heat of reaction} &= \text{Heat of formation of product} \\ &\quad - \text{heat of formation of reactant} \\ &= \text{heat of formation of } [\mathrm{A} \cdot \mathrm{C} + \mathrm{B} \cdot \mathrm{CO_2} + \mathrm{C} \cdot \mathrm{CO} \\ &\quad + \mathrm{D} \cdot \mathrm{CH_4} + \mathrm{E} \cdot \mathrm{H_2O} + \mathrm{F} \cdot \mathrm{H_2}] \\ &\quad - \text{heat of formation of } [\alpha \cdot \mathrm{H_2O} \\ &\quad + \beta \cdot \mathrm{O_2} + \text{biomass}] \end{aligned} \tag{6.18}$$

The heat of formation at 25 °C, or 298 K, is available in Table C.6 (Appendix C). The heat of formation at any other temperature, T K, can be found from the relation:

$$\Delta H_T^0 = \Delta H_{298}^0 + \sum \left(\int_{298}^{T} A' C_{p,i} dT \right)_{product} - \sum \left(\int_{298}^{T} \alpha C_{p,i} dT \right)_{reactants} \tag{6.19}$$

where $C_{p,i}$ is the specific heat of a substance i at temperature T K, and A', … , β are the stoichiometric coefficients of the products and reactants, respectively. The specific heat of gases as a function of temperature is given in Table C.4 (Appendix C).

The net heat, $Q_{gasification}$, to be supplied to the reactor is thus

$$Q_{gasification} = \Delta H_T \quad \text{kJ/mol} \tag{6.20}$$

This expression takes into account both exothermic combustion and endothermic gasification reactions. If the value of $Q_{gasification}$ works out to be negative, the overall process is exothermic, and so no net heat for the reactions is required.

Example 6.2

Find the heat of reaction for the following reaction at 1000 K:

$$\mathrm{C + H_2O\,(gas) = \frac{1}{2}CH_4 + \frac{1}{2}CO_2}$$

Solution

Taking values at the reference temperature, 298 K, we have

$$\text{Heat of formation at 298 K for C} = 0;\ \mathrm{H_2O\,(g)} = -241.8\ \text{kJ/mol};$$
$$\text{for } \mathrm{CH_4} = -74.8\ \text{kJ/mol};\ \mathrm{CO_2} = -393.5\ \text{kJ/mol}$$

$$\text{Total } \Delta H^0_{298} = \text{product} - \text{reactant} = \left[\frac{1}{2}(-74.8 - 393.5) - (-241.8)\right] = 7.65\ \text{kJ/mol}$$

Now, to find the value at 1100 K, we use Eq. (6.21):

$$\Delta H^0_{1100} = \Delta H^0_{298} + \sum\left(\int_{298}^{1100} (C_{p,CH_4} + C_{p,CO_2})\,dT\right)_{product} - \sum\left(\int_{298}^{1100} C_{p,H_2O}\,dT\right)_{reactants} \tag{6.21}$$

The specific heats of gases are taken from Table C.4 (Appendix C) as

C_{p,CH_4} = 22.35 + 0.0481 T kJ/kmol
C_{p,CO_2} = 43.28 + 0.0114 T − 818363/ T^2 kJ/kmol
C_{p,H_2O} = 34.4 + 0.000628 T + 0.0000056 T^2 kJ/kmol

Substituting these values and integrating the above expression, we get

$$\Delta H^\circ_{1100} = 7.65 + 104.58 - 33.578 = 78.65\ \text{kJ/mol}$$

Thus, this reaction is written as

$$\mathrm{C + H_2O\,(gas) \rightarrow \frac{1}{2}CH_4 + \frac{1}{2}CO_2} + 78.65\ \text{kJ/mol}$$

The reaction is endothermic.

Figure 6.22 shows the energy flow in and out of a gasifier. Biomass enters with its chemical energy and sensible heat. The gasifying agents enter with sensible heat at the reference temperature. External heat is added for heating the feeds to the gasification temperature, for meeting any shortfall in the reaction heat requirement, and for wall losses from the reactor. The product gas, with its chemical energy, leaves at the gasifier temperature. Unburnt char leaves with a potential energy in it. The unutilized steam and other gases also leave at the gasification temperature.

The overall energy balance may be written as

Energy input: Enthalpy of (biomass + steam + oxygen) at reference temperature + heating value of biomass + external heat

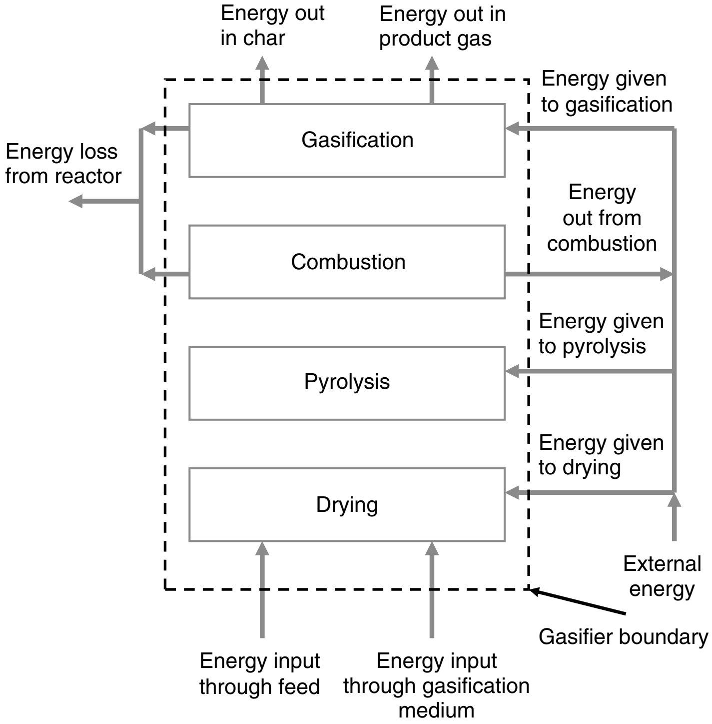

FIGURE 6.22 Energy flow in and out of a gasifier.

Energy output: Enthalpy of product gas at gasifier temperature + heating value of product gas + heat in unconverted char + heat loss from the reactor

If A is the amount of air needed and W is the total steam (from moisture or otherwise) needed to gasify F kg of fuel to produce 1 Nm^3 of product gas, we can write the energy balance of the gasifier taking 0 °C as the reference:

$$\begin{aligned} ACp_aT_0 + FCp_fT_0 + WH_0 + F \times HHV + Q_{ext} &= (C_{CO}V_{CO} + C_{CO_2}V_{CO_2} \\ &+ C_{CH_4}V_{CH_4} + C_{H_2}V_{H_2} + C_{O_2}V_{O_2} + C_{N_2}V_{N_2})T_g + (1 - X_g)WH_g \\ &+ P_cq_c + Q_{gasification} + Q_{loss} + Q_{product} \end{aligned} \tag{6.22}$$

where H_0 and H_g are the enthalpies of steam at the reference temperature and the gasifier exit temperature; C_i and V_i are the volumetric specific heat and the volume of the gas species, i, at temperature T_g leaving the gasifier; $(1 - X_g)W$ is the net amount of steam remaining in the product gas of the gasification reaction; P_c is the amount of char produced; and q_c is the heating value of the char. Here, Q_{loss} is the total heat loss through the wall, radiation from the bed surface, ash drain, and entrained solids, corresponding to 1 Nm^3 of gas generation. This allows computation of external heat addition, Q_{ext} kJ/Nm^3 of product gas to the system. $Q_{product}$ is the amount of energy in the product gas, and $Q_{gasification}$ is the net heat of reaction.

6.7 PRODUCT GAS PREDICTION

A typical gasifier design starts with a desired composition of the product gas. Equilibrium and other calculations are carried out to check how closely that targeted composition is made through a choice of design parameters.

The product of combustion reactions is predominantly made up of carbon dioxide and steam, the percentages of which can be estimated with a fair degree of accuracy from simple stoichiometric calculations. For gasification reactions, this calculation is not straightforward; the fraction of the fuel gasified and the composition of the product gas needs to be estimated carefully. Unlike combustion reactions, gasification reactions do not always reach equilibrium, so only a rough estimate is possible through an equilibrium calculation. Still, this can be a reasonable start for the design until detailed kinetic modeling is carried out in the design optimization stage.

6.7.1 Equilibrium Approach

An equilibrium calculation ideally predicts the product of gasification if the reactants are allowed to react in a fully mixed condition for an infinite period of time. There are two types of equilibrium model. The first one is based on equilibrium constants (stoichiometric model). The specific chemical reactions used for the calculations have to be defined, so this model is not suitable for complex reactions where the chemical formulae of the compounds, the reaction path, or the reaction equations are not known. This requires the second model type, which involves minimization of the Gibbs free energy (nonstoichiometric model). This process is more complex but it is advantageous because the chemical reactions are not needed.

Stoichiometric Model

The stoichiometric model requires a selection of appropriate chemical reactions and information concerning the values of the equilibrium constants. Chapter 5 explains the calculation procedure, so it is not repeated here.

Nonstoichiometric Model

The nonstoichiometric model is based on the premise that at an equilibrium stage the total Gibbs free energy has to be minimized. It is described briefly in Chapter 5. Using Eq. (5.77) we can write the Gibbs free minimization equation for five gas species as follows:

$$\mathrm{CH_4}: \frac{\left(\Delta \overline{G^o}_{CH_4}\right)}{RT} + \ln\left(\frac{n_{CH_4}}{n_{total}}\right) + \frac{1}{RT}\lambda_C + \frac{4}{RT}\lambda_H = 0 \tag{6.23}$$

$$\mathrm{CO_2}: \frac{\left(\Delta \overline{G^o}_{CO_2}\right)}{RT} + \ln\left(\frac{n_{CO_2}}{n_{total}}\right) + \frac{1}{RT}\lambda_C + \frac{2}{RT}\lambda_O = 0 \tag{6.24}$$

$$\mathrm{CO}: \frac{\left(\Delta \overline{G^o}_{CO}\right)}{RT} + \ln\left(\frac{n_{CO}}{n_{total}}\right) + \frac{1}{RT}\lambda_C + \frac{1}{RT}\lambda_O = 0 \tag{6.25}$$

$$\mathrm{H_2}: \frac{\left(\Delta \overline{G^o}_{H_2}\right)}{RT} + \log\left(\frac{n_{H_2}}{n_{total}}\right) + \frac{2}{RT}\lambda_H = 0 \tag{6.26}$$

$$\mathrm{H_2O}: \frac{\left(\Delta \overline{G^o}_{H_2O}\right)}{RT} + \ln\left(\frac{n_{H_2O}}{n_{total}}\right) + \frac{2}{RT}\lambda_H + \frac{1}{RT}\lambda_O = 0 \tag{6.27}$$

The five molar fractions of gases, such as ($n\mathrm{CH_4}/n_{total}$), and the three Lagrangian constants, λ_H, λ_O, and λ_C, can be solved from the five equations and the three mass balance equations for C, H, and O derived from Eq. (5.74). Thus, for given feed and gasification medium and temperature, we can obtain the composition of the product gas.

Equilibrium models have limitations. The effect of tar is not considered here, even though tar can be a major problem in the gasification process and can affect plant operation. An equilibrium model may, for example, result in overestimation of the hydrogen produced. Kinetics, heat, and mass transfer determine the extent of chemicals participation in the chemical equilibrium in a given time or space domain (Florin and Harris, 2008). Furthermore, the equilibrium model assumes infinite speed of reaction and that all reactions will complete; these assumptions are not valid for most practical gasifiers. Nevertheless, equilibrium calculations give a good starting point, providing basic process parameters.

6.8 GASIFIER SIZING

The process design described in the previous section determines such operating parameters as gasification temperature, feed rates of fuel, and gasification medium. Now we can move to the next step, which involves the choice of gasifier configuration and type. Section 6.1.1 discusses the choice of gasifier. Table 6.5 compares the choices by their strength and weaknesses. By carefully examining these along with the type of plant to be designed, we can make a rational choice of gasifier type.

Once the gasifier type has been chosen, the designer can then proceed with the geometric design, where the basic sizes (the geometric dimensions of

TABLE 6.5 Comparison of Strength and Weaknesses of Different Gasifiers

Class	Types	Strength/Weakness	Power Production
Fixed bed	Downdraft	Low heating value, moderate dust, low tar	Small to medium scale
	Updraft	Higher heating value, moderate dust, high tar	
	Crossdraft	Low heating value, moderate dust	
Fluidized bed	Bubbling	Higher than fixed bed throughput, improved mass and heat transfer from fuel, higher heating value, higher efficiency	Medium scale

critical components) of the reactor are determined. At this stage, the designer decides on the geometric configuration of the reactor and its preliminary size. Both configuration and size depend on the reactor technology used.

6.8.1 Moving-Bed Gasifiers

A moving-bed gasifier may be designed on the basis of characteristic design parameters such as specific grate gasification rate, hearth load, and space velocity.

Specific grate gasification rate is the mass of fuel gasified per unit of cross-section area in unit time. The hearth load of a gasifier may be expressed in terms of the fuel gasified, the volume of gas that is produced, or the energy throughput.

$$\text{Hearth load}\left(\text{kg/s}\cdot\text{m}^2\right) = \frac{\text{Mass of fuel gasified}}{\text{Hearth cross-sectional area}}$$

$$\text{Hearth load}\left(\text{Nm}^3/\text{s}\cdot\text{m}^2\right) = \frac{\text{Volumetric gas production rate}}{\text{Hearth cross-sectional area}}$$

or

$$\text{Hearth load}\left(\text{MW/m}^2\right) = \frac{\text{Energy throughput in product gas}}{\text{Hearth cross-sectional area}} \tag{6.28}$$

The hearth load in volume flow rate of gas per unit of cross-section area is also known as *superficial gas velocity* or *space velocity*, as it has the unit of velocity (at reference temperature and pressure).

The following section discusses type-specific design considerations.

Updraft Gasifier

Updraft gasifiers are one of the simplest and most common types of gasifier for biomass. The maximum temperature increases when the feed of air or oxygen increases. Thus, the amount of oxygen feed for the combustion reaction is carefully controlled such that the temperature of the combustion zone does not reach the slagging temperature of the ash, causing operational problems. The gasification temperature may be controlled by mixing steam and/or flue gas with the gasification medium.

The hearth load of an updraft gasifier is generally limited to 2.8 MW/m^2 or 150 $kg/m^2/h$ for biomass (Overend, 2004). For coal it might be higher. In an oxygen-based coal gasifier, for example, the hearth load of a moving bed can be greater than 10 MW/m^2. A higher hearth load increases the space velocity of gas through the hearth, fluidizing finer particles in the bed. Probstein and Hicks (2006) quote space velocities for coal on the order of 0.5 m/h for steam–air gasification and 5.0 m/h for steam–oxygen gasification. Excessive heat generation in such a tightly designed gasifier may cause slagging. Based on the characteristics of some commercial updraft coal gasifiers, Rao et al. (2004) suggest a specific grate gasification rate as 100 to 200 kg $fuel/m^2h$ for RDF pellets, with the gas-to-fuel ratio in the range 2.5 to 3.0. Carlos (2005) obtained a rate of 745 to 916 kg/m^2h with air–steam and air preheat at temperatures of 350 and 830 °C, respectively.

For an updraft gasifier, the height of the moving bed is generally greater than its diameter. Usually, the height-to-diameter ratio is more than 3 : 1 (Chakraverty et al., 2003). If the diameter of a moving bed is too large, there may be a material flow problem, so it should be limited to 3 to 4 m in diameter (Overend, 2004).

Downdraft Gasifier

As we saw in Figures 6.4 and 6.6, the cross-sectional area of a downdraft gasifier may be nonuniform; it is narrowest at the throat. The hearth load is, therefore, based on the cross-sectional area of the throat for a throated gasifier, and for a throatless or stratified downdraft gasifier, it is based on the gasifier cross-sectional area. The actual velocity of gas is, however, significantly higher than the designed space velocity because much of the flow passage is occupied by fuel particles. The velocity is higher in the throat also because of the higher temperature there. Table 6.6 gives some characteristic values of these parameters.

In a downdraft gasifer, the gasification air is injected by a number of nozzles from the periphery (refer to Figure 6.6). The total nozzle area is typically 7 to 4% of the throat area. The number of nozzles should be an odd number so that the jet from one nozzle does not hit a jet from the opposite side, leaving a dead space in between. To ensure adequate penetration of nozzle air into the hearth, the diameter of a downdraft gasifier is generally limited to 1.5 m. This naturally restricts the size and capacity of a downdraft gasifier.

TABLE 6.6 Hearth Load for Downdraft Gasifiers Maximum Values Based on Throat Area

Plant	Gasifier Type	Medium	D_{throat} (m)	$D_{air\ entry}$ (m)	Superficial Velocity at Throat (m/s)	Hearth Load* (MW/m^2)
Gengas	Imbert	Air	0.15	0.3	2.5	15
Biomass Corp.	Imbert	Air	0.3	0.61	0.95	5.7
SERI	Throatless	Air	0.15		0.28	1.67
Buck Rogers	Throatless	Air	0.61		0.23	1.35
Buck Rogers	Throatless	Air	0.61		0.13	0.788
Syngas	Throatless	Air	0.76		1.71	10.28
Syngas	Throatless	Oxygen	0.76		1.07	12.84
SERI	Throatless	Oxygen	0.15		0.24	1.42

*Based on throat area.
Source: Data compiled from Reed and Das, 1988, p. 36.

Table 6.7 lists typical sizes for the Imbert-type downdraft gasifier and shows the relation between throat size and air nozzle diameter.

6.8.2 Fluidized-Bed Gasifiers

No established design method for sizing a fluidized-bed gasifier is available in the literature because, though nearly a century old, this type is still evolving. This section presents a tentative method for determining size based on available information.

Cross-Sectional Area

The inside cross-sectional area of the fluidized-bed gasifier, A_b, is found by dividing the volumetric flow rate of the product gas flow, V_g, by the chosen superficial gas or fluidization velocity through it, U_g, at the operating temperature and pressure.

$$A_b = \frac{V_g}{U_g} \tag{6.29}$$

The volume of gas at the operating temperature and pressure, V_g, is estimated from the mass of air (or other medium), M_{fa}, required for gasification (Eq. 6.29) as well as for fluidization. Thus, V_g is necessarily the gas passing through the grate and the bed.

TABLE 6.7 Sizes of Imbert-Type Gasifiers

d_r/d_h (–)	d_h (mm)	d_r (mm)	$d_{r'}$ (mm)	h (mm)	H (mm)	R (mm)	A (no.)	d_m (mm)	Range of Gas Output (Nm^3/h)	Maximum Wood Consumption (kg/h)	Air Blast Velocity (m/s)
268/60	60	268	150	80	256	100	5	7.5	4–30	14	22.4
300/100	100	300	208	100	275	115	5	10.5	10–77	36	29.4
400/130	130	400	258	110	370	155	7	10.5	17–120	57	32.6
400/150	135	400	258	120	370	155	7	12	21–150	71	32.6
400/175	175	400	308	130	370	155	7	13.5	26–190	90	31.4
400/200	200	400	318	145	370	153	7	16	33–230	110	31.2

Variables not defined in the figure are defined as follows:

d_m = inner diameter of the tuyere
A = number of tuyeres

Source: Data compiled from Reed and Das, 1988.

In some designs, part of the gasifying medium is injected above the distributor grid. In that case, V_g is only the amount that passes through the grid. We can use the mass of gasification medium, M_{fa}, required for gasification for the computation of V_g:

$$V_g = \frac{M_{fa}}{\rho_g} \tag{6.30}$$

where ρ_g is the density of the medium at the gasifier's operating temperature and pressure.

Equation (6.29) requires choosing an appropriate value for the superficial gas (fluidizing) velocity, U_g, through the gasification zone. This is critical as it must be within acceptable limits for the selected particle size to ensure satisfactory fluidization and to avoid excessive entrainment.

Fluidization Velocity

The range of fluidizing velocity, U_g, in a bubbling bed depends on the mean particle size of the bed materials. The choice is made in the same way as for a fluidized-bed combustor. The range should be within the minimum fluidization and terminal velocities of the mean bed particles. The particle size may be within group B or group D of Geladart's powder classification (see Basu, 2006, Appendix I). The typical fluidization velocity for silica sand of about 1 mm mean diameter may, for example, vary between 1.0 and 2.0 m/s.

If the gasifier reactor is a circulating fluidized-bed type, the fluidization velocity in its riser (Figure 6.12) must be within the limits of fast fluidization, which favors groups A or group B particles. Typical fluidization velocity for particle size in the range 150 to 350 microns is 3.5 to 5.0 m/s in a CFB. This type of bed has another important operating condition to be satisfied for operation in the CFB regime. Solids, captured in the gas–solid separator at the gasifier exit, must be recycled back to the gasifier at a rate sufficiently high to create a "fast-fluidized" bed condition in the riser. Additional details about this are available in Basu (2006) or Kunii and Levenspiel (1991).

Gasifier Height

Since gasification involves only partial oxidation of the fuel, the heat released inside a gasifier is only a fraction of the fuel's heating value, and part of it is absorbed by the gasifier's endothermic reactions. Thus, it is undesirable to extract any further heat from the main gasifier column. For this reason, the height of a fluidized-bed gasifier is not determined by heat-transfer considerations as for fluidized-bed boilers. Instead, gas and solid residence times are major considerations.

The total height of the gasifier is made up of the height of the fluidized bed and that of the freeboard above it:

$$\text{Total gasifier height} = \text{bubbling bed height (depth)} + \text{freeboard height} \tag{6.31}$$

Fluidized-Bed Height

The bed height (or depth) of a bubbling fluidized-bed gasifier is an important design parameter. Gas–solid gasification reactions are slower than combustion reactions, so a bubbling-bed gasifier is necessarily deeper than a bubbling-bed combustor, which is typically 1.0- to 1.5-m deep for units larger than 1 m in diameter. Besides pilot plant data or design experience, there is presently no simple means of deciding the bed depth. A deeper bed allows longer gas residence time, but the depth should not be so great compared to its diameter as to cause slugging. The selection of bed height depends on economics. A higher bed height means a higher pressure drop and also a taller reactor. It also should provide a longer residence time for better carbon conversion.

The gasification agent, CO_2 or H_2O, entering the grid takes a finite time to react with char particles to produce the gas. The bulk of the gasifying agent travels up through the bubbles but very little reaction takes place in the bubble phase. Rather, the reaction takes place mostly in the emulsion phase. The extent to which oxygen or steam is converted into fuel gases thus depends on the gas exchange rate between the bubble and emulsion phases as well as on the char-gas reaction rate in the emulsion phase. This is best computed through a kinetic model of the gasifier as illustrated in Section 5.6.2. An alternative is to use an approach based on residence time, as described next.

Residence Time Design Approach A bubbling fluidized bed must be sufficiently deep to provide reactants the time to complete the gasification reactions. This is why residence time is an important consideration for determination of bed height. An approach based on residence time, developed primarily for coal gasification, can be used for biomass char gasification, which gives at least a first estimate of the bed height for a biomass-fueled bubbling fluidized-bed gasifier.

The residence time approach is based on the assumption that the conversion of char into gases is the slowest of all gasifier processes, so the reactor should provide adequate residence time for the char to complete its conversion to the desired level. Here is a simplified method.

Given the following assumption:

- The reactivity factor is f_o, = 1 (which lies between $0 < f_o \le 1$).
- The solid is in a perfectly mixed condition (i.e., continuous stirred-tank reactor).

Then, the volume of the fluidized bed, V, is calculated using the equation

$$V = \frac{W_{out}\theta}{\rho_b} \tag{6.32}$$

where W_{out} is the char moving out; kg/s = $(1-X)\ W_{in}$; X is the fraction of the char in the converted feed; ρ_b is the bed density, which can be estimated theoretically from fluidization hydrodynamics and regime (kg/m^3); and θ is the residence time of the char in the bed, or reaction time (s).

The residence time approach assumes that the water–gas reaction, ($C + H_2O \rightarrow CO + H_2$), as written in Eq. (6.33) is the main gasification reaction, where the char is consumed primarily by the steam gasification reaction for *n*th-order kinetics:

$$\frac{1}{m}\frac{dC}{dt} = k[H_2O]^n \tag{6.33}$$

where m is the initial mass of the biomass and C is the total amount of carbon gasified in time, t. Taking a logarithm of this,

$$\ln\left(\frac{1}{m}\frac{dC}{dt}\right) = \ln(k) + n\ln[H_2O] \tag{6.34}$$

experiments can be carried out taking a known weight of the biomass and measuring the change in carbon conversion at different time intervals for a given temperature, steam flow, and pressure. Using these data, graphs are plotted between $\ln\left(\frac{1}{m}\frac{dC}{dt}\right)$ and $\ln[H_2O]$. The y-intercept in this graph will give the value of k, and the slope will give the value of n. An example of such a plot is shown in Figure 6.23.

The experiment is carried out for different operating temperatures such as 700 °C, 800 °C, and 900 °C, so, for each temperature, one k value is obtained.

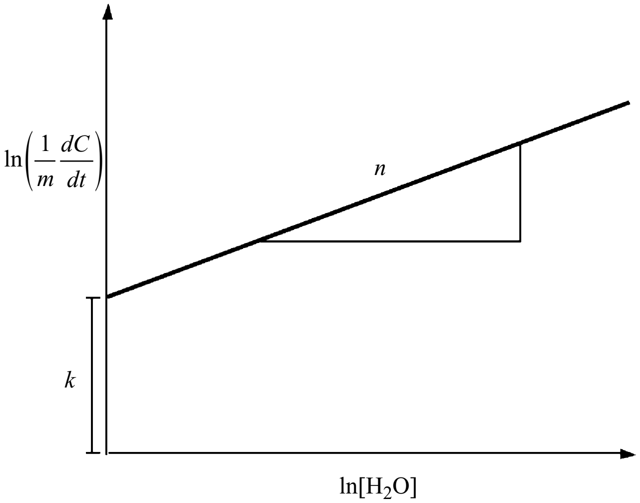

FIGURE 6.23 Plot of Eq. (6.34) for determination of residence time.

Now k can be expressed as

$$k = k_0 \exp\left(-\frac{E_a}{RT}\right)$$

$$\ln k = \ln k_0 - \frac{E_a}{RT} \tag{6.35}$$

This shows that if we plot a graph between ln k and $1/T$, the y-intercept will give the value of k_0 and the slope will give the value of $(-E_a/R)$.

The reaction rate for the steam gasification of biomass is given by

$$\frac{dC}{dt} = k_0 m \exp\left(-\frac{E_a}{RT}\right)[\mathrm{H_2O}]^n \tag{6.36}$$

This gives the generalized reaction rate that shows the dependence of the gasification rate on temperature, mass of carbon or char, and concentration of steam/air/oxygen.

From a knowledge of the reaction rate, the residence time, θ, can be calculated as

$$\theta = C_0 \frac{X}{r} \tag{6.37}$$

where C_0 is the initial carbon in the biomass particle, kg; X is the required carbon conversion (–); and r is the steam gasification reaction rate (kg/s). We can avoid such experiments if there is a suitable expression for the rate of steam gasification of the designed biomass char (Sun et al., 2007).

From knowledge of the required solid residence time, θ, then, the bed volume, V_{bed}, is

$$V_{bed} = \frac{F[C]\theta}{(1-\varepsilon)\rho_s x_{char}} \tag{6.38}$$

where $F[C]$ is the char feed rate into the gasifier and ρ_s is the density of the bed solids. In a typical bubbling bed, the bed voidage is ~0.7. The bed generally contains 5 to 8% (by weight) of reacting char (x_{char}); the remaining solids are inert bed materials.

The bed height, H_{bed}, is known by dividing bed volumer by the bed area, A_b, which is known from chosen superficial velocity

$$H_{bed} = \frac{V_{bed}}{A_b} \tag{6.39}$$

Design charts for residence time, θ, of test coals for different feed conversions and S/C or O/C ratios are given in the *Coal Conversion Systems Technical Data Book* (U.S. DOE, 1978). The residence time may be adjusted for the reactivity of the char in question and for the reactivity of its partial gasification before it enters the gasifier.

Other Considerations

Although virgin biomass contains little or no sulfur, some waste biomass fuels do. For these, limestone is fed into the fluidized-bed gasifier for in-bed sulfur removal. The height of the gasifier (freeboard and bed) should be adequate to allow the residence time needed for the desired sulfur capture.

The tar produced should be thermally cracked inside the gasifier as far as possible. Therefore, the depth of the gasifier should be such that the gas residence time is adequate for the desired tar conversion/cracking.

The deeper the bed, the higher the pressure drop across it and the higher the pumping cost of air. Because bubble size increases with bed height, a deeper bed gives larger bubbles with reduced gas–solid mixing. Furthermore, if the bubble size becomes comparable to the smallest dimension of the bed cross-section, a highly undesirable slugging condition is reached. This imposes another limit on how deep the dense section of a fluidized bed can be.

Some biomass char, like that from wood, is fine and easily undergoes attrition in a fluidized bed. In such cases a deeper bed may not guarantee a longer residence time (Barea, 2009). Here, special attention must be paid to capturing the char and either combusting it in a separate chamber to provide heat required by the gasifier, or reinjecting it at an appropriate point in the bed where solids are descending.

A kinetic model (*n*th-order, shrinking particle, and shrinking core) may also be used to determine the residence time, the net solid holdup, and therefore the height of the dense bed.

Freeboard Height

Entrainment of unconverted fine char particles from the bubbling bed is a major source of carbon loss. The empty space above the bed, the *freeboard*, allows entrained particles to drop back into it. A bubbling, turbulent, or spouted fluidized bed must have such a freeboard section to help avoid excessive loss of bed materials through entrainment and to provide room for conversion of finer entrained char particles. The freeboard height must be sufficient to provide the required residence time for char conversion. It can be determined from experience or through kinetic modeling.

A larger cross-sectional area and a taller freeboard increase the residence time of gas/char and reduce entrainment. From an entrainment standpoint, the freeboard height need not exceed the transport disengaging height (TDH) of a bed because no further reduction in entrainment is achieved beyond this.

6.9 ENTRAINED-FLOW GASIFIER DESIGN

Because the gas residence time in an entrained-flow reactor is very short—on the order of a few seconds—to complete the reactions, the biomass particles must be ground to extremely fine sizes (less than 1 mm). The residence time requirement for the char is thus on the order of seconds. Section 6.9.1 describes some important considerations for entrained-flow gasifier design.

Although an entrained-flow gasifier is ideally a plug-flow reactor, in practice this is not necessarily so. The side-fed entrained-flow gasifier, for example, behaves more like a continuous stirred-tank reactor (CSTR). As we saw in Figure 6.15, at a certain distance from the entry point, fuel particles may have different residence times depending on the path they took to arrive at that section. Some may have traveled a longer path and so have a longer residence time. For this reason, a plug-flow assumption may not give a good estimate of the residence time of char.

6.9.1 Gasifier Chamber

Most commercial entrained-flow gasifiers operate under pressure and therefore are compact in size. Table 6.8 gives data on some of these operating in the United States and China.

A typical downflow entrained-flow gasifier is a cylindrical pressure vessel with an opening at the top for feed and another at the bottom for discharge of ash and product gas. The walls are generally lined with refractory and insulating materials, which serve three purposes: (1) they reduce heat loss through the wall, (2) they act as thermal storage to help ignition of fresh feed, and (3) they prevent the metal enclosure from corrosion.

TABLE 6.8 Characteristic Sizes of Some Entrained-Flow Gasifiers

Gasifier	Volume (m^3)	Reactor External Diameter (m)	Reactor Internal Diameter (m)	Reactor Height (m)
Tennessee Eastman	12.7	2.79	1.67	4.87
Cool water	17	3.17	2.13	3.73
Cool water	25.5	3.17	2.13	6
Cool water	12.7	2.79	1.67	4.62
Shandong fertilizer	12.7	2.79	1.67	4.87
Shanghai Chemical	12.7	2.79	1.67	4.87
Harbei fertilizer	12.7	2.79	1.67	4.87

Source: Data compiled from Zen, 2005.

The thickness of the refractory and insulation used is to be chosen with care. For example, biomass ash melts at a lower temperature and is more corrosive than most coal ash, so special care needs to be taken in designing the gasifier vessel for biomass feedstock.

The construction of a side-fed gasifier is more complex than that of a top-fed gasifier, as the reactor vessel is not entirely cylindrical and requires numerous openings. The bottom opening is for the ash drain, the top opening is for the product gas, and the side ports are for the feed. Additional openings may also be required depending on the design. Because of the complexity in the design of a pressure vessel operating at 30 to 70 bars and temperatures exceeding 1000 °C, any additional openings or added complexity in the reactor configuration must be weighed carefully against perceived benefits and manufacturing difficulties.

6.9.2 Auxiliary Items

The following subsections discuss the design of auxiliary systems in fluidized-bed gasifiers.

Position of Biomass Feeding Position

The feed points for the biomass should be such that entrainment of any particles in the product gas is avoided. This can happen when the feed points are located too close to the expanded bed surface of a bubbling fluidized bed. If they are in close proximity to the distributor plate, excessive combustion of the volatiles in the fluidizing air produced can occur. To avoid this, they should be some distance further above the grate.

Nascent tar is released close to the feed point, so tar cracking can be important for some designs. If tar is a major concern, the feed port should be close to the bottom of the gasifier so that the tar has adequate residence time to crack (Barea, 2009).

Distributor Plate

The distributor plate of a fluidized bed supports the bed materials. It is no different from that used for a fluidized combustor or boiler. The ratio of pressure drop across the bed and that across the distributor plate must be estimated to arrive at the plate design. More details are available in books on distributor plate design, including Basu (2006, Chapter 11). The typical open area in the air distributor grate is only a few percentage points.

Bed Materials

For the process design of a fluidized-bed gasifier, the choice of bed materials is crucial. These comprise mostly granular inorganic solids and some (<10%) fuel particles. For biomass, sand or other materials are used (as explained next);

coal gasification requires granular ash produced from the gasification process. Sometimes limestone is added with coal particles to remove sulfur. At different stages of calcination and sulfurization, the limestone can also form a part of the bed material.

Biomass has very little ash (less than 1% for wood), so silica sand is normally used as the inert bed material. This is a natural choice because silica is inexpensive and the most readily available granular solid. One major problem with silica sand is that it can react with the potassium and sodium components of the biomass to form eutectic mixtures having low melting points, thereby causing severe agglomeration. To avoid this, the following alternative materials can be used:

- Alumina (Al_2O_3)
- Magnesite ($MgCO_3$)
- Feldspar (a major component of Earth's crust)
- Dolomite ($CaCO_3 \cdot MgCO_3$)
- Ferric oxide (Fe_2O_3)
- Limestone ($CaCO_3$)

Magnesite (MgO) was successfully used in the first biomass-based IGCC plant in Värnamo, Sweden (Ståhl et al., 2001).

Tar is a mixture of higher-molecular-weight (higher than benzene) chemical compounds that condenses on downstream metal surfaces at lower temperatures. It can plug the passage and/or make the gas unsuitable for use. The bed materials, besides serving as a heat carrier, can catalyze the gasification reaction by increasing the gas yield and reducing the tar. Bed materials that act as a catalyst for tar reduction are an attractive option. Some are listed here (Pfeifer et al., 2005; Ross et al., 2005):

- Olivine
- Activated clay (commercial)
- Acidified bentonite
- Raw bentonite
- House brick clay

Common house brick clay can be effectively used in a CFB gasifier to reduce tar emission and enhance hydrogen production. The alkalis deposited on the bed materials from biomass may potentially behave as catalysts if their agglomerating effect can be managed (Ross et al., 2005).

Tar production can be reduced using olivine. The Fe content of olivine is catalytically active, and that helps with tar reforming (Hofbauer, 2002). Nickel-impregnated olivine gives even better tar reduction as nickel is active for steam tar reforming (Pfeifer et al., 2005).

Bingyan et al. (1994) reported using ash from the fuel itself (sawmill dust) as the bed material in a CFB gasifier. This riser is reportedly operated at a very low velocity of 1.4 m/s, which is 3.5 times the terminal velocity of the biomass particles. Chen et al. (2005) tried to operate a 1-MWe CFB gasifier with rice husk alone, but the system had difficulty with fluidization in the loop seal because of the low sphericity of the husk ash; however, the main riser reportedly operated in the fast bed regime without major difficulty.

6.10 DESIGN OPTIMIZATION

Design optimization generally starts after the preliminary design is complete and actual project execution is set to begin. It has two aspects: (1) process and (2) engineering.

Process optimization tells the designer if the preliminary design will give the best performance in terms of efficiency and gas yield, and how this is related to the operation and design parameters. Commercial simulation programs (mathematical models) or computational fluid dynamics codes are the most effective tools for this purpose. Engineering optimization involves optimizing the reactor configuration to enhance its operability, maintainability, and cost reduction.

6.10.1 Process Optimization

Process optimization enhances gasifier performance in terms of the following indicators:

- Cold- and hot-gas efficiency
- Unconverted carbon and tar concentration in the product gas
- Composition and heating value of the product gas

One can approach optimization either through experiments or through kinetic modeling.

Experiments are the best and most reliable means of optimizing process parameters, as they are based on the actual or prototype gasifier. However, they have several limitations, and are expensive. Furthermore, practical difficulties may not allow all operational parameters to be explored. An alternative is to conduct tests in a controlled laboratory-scale unit and to calibrate the resulting data to the full-scale unit. This allows the scale-up of data from the laboratory to the full-scale unit with a reasonable degree of confidence.

Optimization through Kinetic Modeling

With a kinetic model we can predict the performance of a gasifier already designed because it utilizes both configuration and dimensions of the reactor. Kinetic modeling can help optimize or fine-tune the operating parameters for best performance in a given situation. Section 5.6 described a kinetic model for gasifiers.

6.11 PERFORMANCE AND OPERATING ISSUES

Gasifier performance is measured in terms of both quality and quantity of gas produced. The amount of biomass converted into gas is expressed by gasification efficiency. The product quality is measured in terms of heating value as well as amount of desired product gas.

6.11.1 Gasification Efficiency

The efficiency of gasification is expressed as cold-gas efficiency, hot-gas efficiency, or net gasification efficiency. These are described in the following subsections.

Cold-Gas Efficiency

Cold-gas efficiency is the energy input over the potential energy output. If M_f kg of solid fuel is gasified to produce M_g kg of product gas with an LHV of Q_g, the efficiency is expressed as

$$\eta_{cg} = \frac{Q_g M_g}{LHV_f \, M_f} \tag{6.40}$$

where LHV_f is the lower heating value (LHV) of the solid fuel.

Example 6.3

Air–steam gasifier data include the mass composition of the feedstock:

C–66.5%
O–7%
H–5.5%
N–1%
Moisture–7.3%
Ash–12.7%
LHV–28.4 MJ/kg

and the volume composition of the product gas:

CO–27.5%
CO_2–3.5%
CH_4–2.5%,
H_2–15%
N_2–51.5%

The dry air supply rate is 2.76 kg/kg of feed; the steam supply rate is 0.117 kg/kg of feed; the moisture content is 0.01 kg of H_2O per kg of dry air; and the ambient temperature is 20 °C.

Find:

- The amount of gas produced per kg of feed
- The amount of moisture in the product gas

- The carbon conversion efficiency
- The cold-gas efficiency

Solution

Table C.3 (Appendix C) shows the mass fraction of N_2 and O_2 in air as 0.755 and 0.232, respectively. The nitrogen supply from air is

$$0.755 \times 2.76 = 2.08\ \text{kg}\ N_2/\text{kg feed}$$

The total nitrogen supplied by the feed air and the fuel feed, which carry 1% nitrogen, is

$$2.08 + 0.01 = 2.09\ \text{kg}\ N_2/\text{kg feed} = (2.09/28) = 0.0747\ \text{kmol}\ N_2/\text{kg feed}$$

noting that volume percent equals molar percent in a gas mixture.

Since the product gas contains 51.5% by volume of nitrogen, the amount of the product gas per kg of feed is

$$0.0747/0.515 = \mathbf{0.145\ kmol\ gas/kg\ feed}$$

Similarly, the oxygen from the air flow to the gasifier is

$$0.232 \times 2.76 = 0.640\ \text{kg/kg feed}$$

The steam supplied per kg of fuel is 0.117 kg, so the oxygen associated with the steam supply is

$$0.117 \times (8/9) = 0.104\ \text{kg/kg feed}$$

Oxygen also enters through the 7.3% moisture in the fuel and the 1% moisture in the air feed. The total oxygen from moisture is

$$0.073 \times (8/9) + 0.01 \times 2.76 \times (8/9) = 0.065 + 0.0245 = 0.0895\ \text{kg/kg feed}$$

The total oxygen flow to the gasifier, including the 7% that comes with the fuel, is

$$0.640 + 0.104 + 0.0895 + 0.07 = 0.9035\ \text{kg}\ O_2/\text{kg feed}$$

Hydrogen Balance

The total hydrogen inflow to the gasifier with fuel, steam, and moisture in the fuel and moisture in the air is

$$0.055 + 0.117/9 + 0.073/9 + 0.0276/9 = \mathbf{0.0792\ kg/kg\ feed}$$

The hydrogen leaving with H_2 and CH_4 in dry gas, noting that 1 mole of CH_4 contributes 2 mols of H_2, is

$$(0.15 + 2 \times 0.025) \times 0.145 = 0.029\ \text{kmol/kg feed} = 0.029 \times 2$$
$$= \mathbf{0.058\ kg\ hydrogen/kg\ feed}$$

To find the moisture in the product gas, we deduct the hydrogen in the dry gas from the total hydrogen inflow obtained earlier, using the hydrogen balance:

$$\text{Hydrogen inflow} - \text{hydrogen out through dry product gas}$$
$$= 0.0792 - 0.058 = 0.0212\ \text{kg/kg feed}$$

The steam or moisture associated with this hydrogen in the gas is

$$0.0212 \times (18/2) = \mathbf{0.1908\ kg/kg\ feed}$$

Carbon Balance

The carbon-bearing gases—CO, CO_2, and CH_4—in the dry gas each contain 1 mol of carbon. So the total carbon in 0.145 kmol/kg of fuel product gas is

$$(0.275 + 0.035 + 0.025) \times 0.145 = 0.0485\ \text{kg mol/kg feed} = 0.0485 \times 12 = 0.583\ \text{kg/kg feed}$$

The carbon input, as found from the composition of the feed, is 0.665 kg/kg feed. The carbon conversion efficiency is found by dividing the carbon in the product gas by that in the fuel:

$$= (0.583/0.665) \times 100 = \mathbf{87.6\%}$$

Energy Balance

The heats of combustion for different gas constituents are taken from Table C.2 (Appendix C). They are:

CO–12.63 MJ/nm^3
Hydrogen–12.74 MJ/nm^3
Methane–39.82 MJ/nm^3

We note that 1 kg of feed produces 0.145 kmol of gas, the volumetric composition of which is

CO–27.5%
CO_2–3.5%
CH_4–2.5%
H_2–15%
N_2–51.5%

By multiplying the heating value of the appropriate constituents of the product gas, we can find the total heating value of the product gas (the volume of 1 kmol of any gas is 22.4 nm^3):

$$(12.63 \times 0.275 + 12.74 \times 0.15 + 39.82 \times 0.025)\,\text{MJ/nm}^3 \times 0.145\ \text{kmol/kg feed} \times 22.4\ \text{nm}^3/\text{kmol} = \mathbf{20.6\ MJ/kg\ feed}$$

The total energy input is equal to the heating value of the feed, which is 28.4 MJ/kg.

From Eq. (6.40), the cold-gas efficiency is

$$(20.6/28.4) \times 100 = \mathbf{72.5\%}$$

Hot-Gas Efficiency

Sometimes gas is burned in a furnace or boiler without being cooled, creating a greater utilization of the energy. Therefore, by taking the sensible heat of the hot gas into account, the hot-gas efficiency, η_{hg}, can be defined as

$$\eta_{hg} = \frac{Q_g M_g + M_g C_p (T_f - T_0)}{LHV_f M_f} \tag{6.41}$$

where T_f is the gas temperature at the gasifier exit or at the burner's entrance, and T_0 is the temperature of the fuel entering the gasifier. The hot-gas efficiency assumes the heating of the unconverted char to be a loss.

Example 6.4

The gas produced by the gasifier in Example 6.3 is supplied directly to a burner at the gasifier exit temperature, 900 °C, to be burnt for co-firing in a boiler. Find the hot-gas efficiency of the gasifier.

Solution

The product gas enters the burner at 900 °C (1173 K). To find the enthalpy of the product gas, we add the enthalpies of its different components. Specific heats of individual components are calculated using the relations from Table C.4 (Appendix C). For example, the specific heat of CO at 1173 K is

$$27.62 + 0.005 \times 1173 = 33.48 \text{ kJ/kmol.K}$$

From Example 6.3, the amount of product gas is 0.145 kmol/kg fuel. The enthalpy of CO in the product gas that contains 27.5% CO above the ambient temperature, 25 °C or 298 K, is

$$(0.145 \times 0.275)\text{kmol/kg feed} \times 33.48 \text{ kJ/kmol.K} \times (1173 - 298)\text{K} \times 10^{-3} \text{ MJ/kJ} = 1.168 \text{ MJ/kg feed}$$

Similarly enthalpy of other products,

CO_2: (0.145 × 0.035) × 56.06 × (1173–298) × 10^{-3} = 0.249 MJ/kg feed
H_2: (0.145 × 0.15) × 31.69 × (1173–298) × 10^{-3} = 0.0603 MJ/kg feed
N_2: (0.145 × 0.515) × 32.13 × (1173–298) × 10^{-3} = 0.21 MJ/kg feed
CH_4: (0.145 × 0.025) × 78.65 × (1173–298) × 10^{-3} = 0.249 MJ/kg feed

The amount of steam in the flue gas was calculated as 0.1908 kg/kg of feed. To find the enthalpy of this steam above 298 K, we take values of the steam enthalpy at 1 bar of pressure at 1173 K and 298 K. The values are 4398.05 and 104.93 kJ/kg, respectively, so the enthalpy in water is

$$H_2O: 0.1908x\ (4398.05 - 104.93) \times 10^{-3} = 0.819 \text{ MJ/kg feed}$$

Adding these we get the total enthalpy of the product gas at 900 C.

$$1.168 + 0.249 + 0.060 + 0.21 + 0.249 + 0.819 = 2.76 \text{ MJ/kg feed}$$

The total thermal energy is

$$\text{Heating value} + \text{enthalpy} = 20.6 + 2.76 = 23.34 \text{ MJ/kg coal}$$

The total gasifier efficiency is

$$\text{Total thermal energy/heat in feedstock} = \frac{23.34}{28.4} \times 100\% = \mathbf{82.2\%}$$

Net Gasification Efficiency

The enthalpy or energy content of the gasification medium can be substantial, and so, for a rigorous analysis, these inputs should be taken into consideration. At the same time, part of the input energy is returned (energy credit) by the tar or oil produced as well as by any recovery of the heat of vaporization in the product gas. A more rigorous energy balance may thus be written as

- Total gross energy input = fuel energy content + heat in gasifying mediums
- Net energy input = total energy input – energy recovered through burning tar, oil, and condensation of steam in the gas

The net gasification efficiency can be written as

$$\eta_{net} = \frac{\text{Net energy in the product gas}}{(\text{Total energy input to the gasifer} - \text{credits})} \tag{6.42}$$

Example 6.5

In most steam-fed gasifiers, a large amount of steam remains unutilized. For the given problem, find the amount of unutilized steam. Also find the cold-gas and net gasification efficiency of a fixed-bed gasifier that uses steam and oxygen to gasify grape wastes (HHV = 21,800 kJ/kg). The product gas composition (mass basis) is

CO–31.8%
H_2–3.1%
CO_2–38.2%
CH_4–1.2%
C_3H_8–0.9%
N_2–1%
H_2O–44.8%

The HHV of the product gas is 8.78 MJ/kg.

The ultimate and proximate analyses of the biomass are as given in Table 6.9. The total fuel feed rate is 25 kg/s; the oxygen feed rate is 5.3 kg/s. The steam is fed into the gasifier at a rate of 27 kg/s at 180 °C and 5 bars of pressure. The product contains dry gas, condensable moisture, and tar. The tar production rate is 1.3 kg/s and is analyzed to contain 85% carbon and 15% hydrogen by weight. The heating value of the tar is 42,000 kJ/kg. The oxygen is produced from air using an oxygen-separation unit (OSU) that consumes 4000 kJ of energy/kg of the oxygen produced (assume full conversion of char).

Find the amount of product gas produced and the fraction of steam that remains unutilized.

Solution

Hydrocarbon hydrogen from the ultimate analysis is $5.83 \times (1 - 0.04) = 5.6\%$. Additional hydrogen also in the moisture is $0.04 \times (2/18) = 0.44\%$. Thus, the total

TABLE 6.9 Analyses of Ultimate and Proximate Biomass

	Proximate Analysis in Mass (%)		Ultimate Analysis (dry basis) in Mass (%)
Ash	4.2	Carbon	55.59
Volatile matter	70.4	Hydrogen	5.83
Fixed carbon	21.4	Nitrogen	2.09
Moisture	4.0	Sulfur	0.21
		Oxygen	32.08
Total	100.0	Ash	4.2

hydrogen, on an as-received basis, is 5.6 + 0.40 = 6.0%. The feed rate of the total hydrogen through the fuel is 25 × 6.0/100 = 1.5 kg/s.

A mass balance between input and output helps determine the production rate of the gas. Output equals input, so

Product + ash + tar and oil = fuel + oxygen + steam
Product + (25 × 0.042) + 1.3 = 25 + 5.3 + 27
Product = 54.95 kg/s

The product contains gas, the composition of which was given previously (M_{gas}), as well as the condensate, M_{cond}. To find the gas we carry out a carbon balance from its measured composition.

Part (a) Carbon Balance

The total carbon in the gas (%) is

$$\text{Molecular weight of carbon} \times (\%\text{ mass of compound/molecular weight of the compound}) = 12\,(31.8/28 + 38.2/44 + 1.2/16 + 0.9/44) = 33.29\%$$

The carbon balance gives

$$\text{Carbon in gas} + \text{carbon in tar and oil} = \text{carbon in fuel}$$

$$M_{gas} \times (33.29.9/100) + 1.3 \times 0.85 = 25 \times 0.5559$$

$$M_{gas} = 38.4\,\text{kg/s}$$

$$\text{Total product} = 54.95 = 38.4 + M_{cond}$$

$$M_{cond} = 16.55\,\text{kg/s}$$

Part (b) Water Balance

Water enters the gasifier through the steam as well as through the moisture in the fuel, so

$$\text{Water in steam} + \text{water in fuel} = \text{water used in gasification} + \text{water leaving as waste steam/water}$$

The water used in gasification is

$$27 + 25 \times 0.04 - 38.4 \times 0.448 = 9.8\ \text{kg/s}$$

Therefore, the percent of steam not utilized is

$$1 - 9.8/27 = 63.7\%$$

6.11.2 Operational Considerations

A large number of operational issues confront a biomass gasifier. Universal to all gasifier types are problems related to biomass handling and feeding. Bridging of biomass over the exit of a hopper is common for plants that use low-shape-factor (flaky) biomass such as leaves and rice husk. This problem is discussed in more detail in Chapter 8.

Fixed-Bed Gasifier

Charcoal particles become porous and finer during their time in the gasification zone. Thus, in a downdraft gasifier, when fine charcoal drops into the ash pit, the product gas can easily carry the particles as dust. Escaping particles can be a source of carbon loss, and they often plug downstream equipment.

The movement of solids in any layer of a moving-bed gasifier should be equal to the feed rate of the fuel at the top. Even with that balance, if the fuel is dry, the pyrolysis zone may, in an updraft gasifier, travel upward faster, thus consuming the layer of fresh fuel above and leading to premature pyrolysis. The gas lost in this way may result in lower gasification efficiency.

On the other hand, if the fuel is moist, its pyrolysis may be delayed. This may move the pyrolysis zone downward. In the extreme case, the cooler pyrolysis zone may sink sufficiently to extinguish the gasification and combustion reaction. Clearly, a proper balance of rates of fuel flow and air flow is required for stabilization of each of these zones in respective places.

Fluidized-Bed Gasifier

The startup of a fluidized-bed gasifier is similar to the startup of a fluidized bed combustor. The inert bed materials are preheated either by an overbed burner or by burning gas in the bed. Once the bed reaches the ignition temperature of the fuel, the feed is started. Combustion is allowed to raise the temperature. After that, the air/oxidizer-to-fuel ratio is slowly adjusted to switch to gasification mode.

One major problem with fluidized-bed gasifiers is the entrainment (escape) of fine char with the product gas. The superficial velocity in a fluidized bed is often sufficiently high to transport small and light char particles, contributing to major carbon loss. A tall freeboard can reduce the problem, but that has a cost penalty. Instead, most fluidized-bed gasifiers use a cyclone and a recycle system to return the entrained char particles back to the gasifier.

Entrained-Flow Gasifier

The startup procedure for an entrained-flow gasifier takes a long time because a startup burner must heat up the reactor vessel wall. During this time, the reactor vessel is not pressurized. Once oil or gas flame heats up the thick refractory wall to ~1100 °C, the startup burner is withdrawn and the fuel is injected along with the oxidizer (Weigner et al., 2002). The hot reactor wall serves as an igniter for the fuel, which once ignited continues to burn in the combustion zone, consuming the oxygen. For this reason, the fuel injector in an entrained-flow reactor is also called the burner. The reactor is pressurized slowly once the main fuel is ignited.

The gasifying medium is rarely premixed with the fuel. The fuel and the medium are often injected coaxially, as in a pulverized-coal (PC) burner in a boiler or furnace. They immediately mix on entering the reactor. The operation of a gasifier "burner" is similar to that of conventional burners, so design methods for PC or oil burners can be used for a rough and an initial sizing. The use of a separate startup burner involves replacing it with a fuel injector. This is especially difficult for water-cooled walls because their lower thermal inertia cannot hold the wall temperature long enough. Integration of the startup burner in the existing fuel injector is the best option.

Tar Cracking

Several options for tar control and destruction are available; these were discussed in Chapter 4. In fixed-bed gasifiers, thermal cracking or burning has been used with success. In one such design, as shown in Figure 6.24, the air entering the gasifier passes through an aspirator that entrains the tar vapor. The mixture is then burnt in the combustion zone. The aspirator can be outside or inside the gasifier.

Symbols and Nomenclature

A_b = cross-sectional area of the fluidized bed (m^2)
ASH = fractional of ash in the fuel in dry basis (–)
C = fractional of carbon in the fuel in dry basis (–)
C_i = volumetric specific heat of gas i ($kJ/nm^3.K$)
C_o = initial carbon in the biomass (kg)
C_p = specific heat of the gas (kJ/kg.C)
E_a = activation energy (kJ/mol)
EA = excess air coefficient (–)
ER = equivalence ratio (–)
F = amount of dry fuel required to obtain 1 Nm^3 of product gas (kg/nm^3)
$F[C]$ = char feed rate into the gasifier (kg/s)

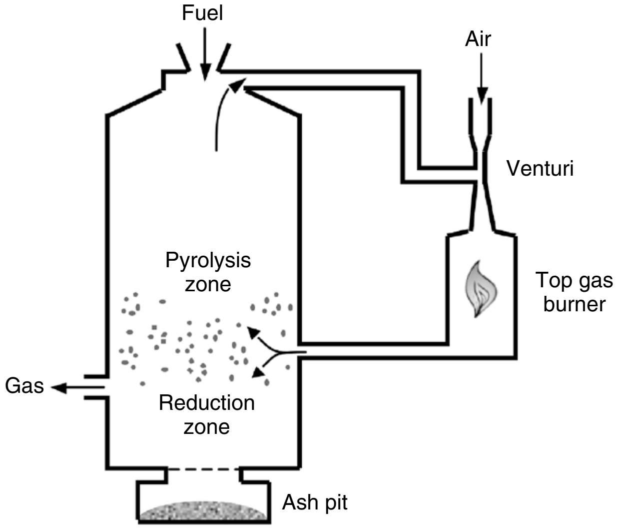

FIGURE 6.24 Gasifier with an aspirator for cracking tar. Fresh air picks up the tar from the gasifier and injects it into the high-temperature combustion zone.

H = fractional of hydrogen in the fuel in dry basis (–)
HHV = higher heating value (kJ/kg)
HHV_d = higher heating value of biomass on dry basis (MJ/kg)
HHV_{daf} = higher heating value of biomass on dry ash–free basis (MJ/kg)
H_{bed} = height of the bed (m)
H_g = enthalpy of steam at gasification temperature (kJ/kg)
H_{in} = heat of the input gas (kJ)
$[H_2O]$ = concentration of steam (–)
k = rate constant (s^{-1})
k_0 = pre-exponential constant in the Arrhenius equation (s^{-1})
LHV_{bm} = lower heating value of the biomass (MJ/kg)
LHV_{daf} = lower heating value of biomass on dry ash-free basis (MJ/kg)
LHV_f = lower heating value of the solid fuel (MJ/Nm^3)
LHV_g = lower heating value of the produced gas (MJ/Nm^3)
m = mass-flow rate of carbon or char (kg/s)
m_{th} = theoretical air requirement for complete combustion of a unit of biomass (kg/kg)
M_a = amount of air required for gasification of unit mass of biomass (kg/kg)
M = fractional of moisture in the fuel (–)
M_{daf} = moisture based on dry ash-free basis
M_f = fuel flow rate (kg/s)
M_{fh} = quantity of steam (kg/s)
M_g = gas produced (kg/s)
n = order of reaction (–)
n_i = number of moles of species i (–)
N = fractional of nitrogen in the fuel in dry basis (–)
n_{total} = total number of moles

O = fractional of oxygen in the fuel in dry basis (–)
P_c = amount of char produced per nm^3 of product gas (kg/nm^3)
q_c = heating value of char (kJ/kg)
Q = power output of the gasifier (MWth)
Q_{ext} = external heat addition to the system (kJ/Nm^3)
Q_g = Lower heating value of the product gas from gasification (MJ/Nm^3)
$Q_{gasification}$ = heat supplied to gasify 1 mol of biomass (kJ/mol)
Q_{loss} = heat loss from the gasifier (kJ/Nm^3)
r = steam gasification reaction rate (kg/s)
R = universal gas constant (0.008314 kJ/mol.K)
S = fractional of sulfur in the fuel in dry basis (–)
SC = steam to carbon molar ratio (–)
t = time (s)
T = temperature (K)
T_f = gas temperature at the exit (°C)
T_g = gas temperature (°C)
T_0 = gas temperature at the entrance (°C)
U_g = fluidizing velocity (m/s)
V = volume of the fluidized bed (m^3)
V_{bed} = volume of the bed (m^3)
V_{daf} = volatile based on dry mass-free basis
V_g = gas generation rate (m^3/s)
V_g = volumetric flow rate of product gas (Nm^3/s)
V_i = volumetric fraction of gas species i (–)
W = total steam needed in Eq. 6.22 (kg/s)
W_{in} = rate of the char moving in (kg/s)
W_{out} = rate of the char moving out (kg/s)
x_{char} = weight of the reacting char (kg)
X = fraction of char in the feed converted (–)
X_c = fixed carbon fraction in the fuel (kg carbon/kg dry fuel)
X_{char} = char fraction in bed (–)
X_g = fraction of steam used up in gasification
ε = voidage of the bed (–)
λ_I = Lagrangian multiplier for species i (–)
ρ_g = density of air at the opening temperature and pressure of the gasifier (kg/m^3)
θ = residence time of char in bed or reactor (s)
ρ_b = bed density (kg/m^3)
ρ_s = density of bed solids (kg/m^3)
η_{gef} = gasifier efficiency (–)
η_{ceff} = cold gas efficiency (–)
η_{cg} = cold gas efficiency of the gasifier (–)
η_{hg} = hot gas efficiency of the gasifier (–)
η_{net} = net gasification efficiency of the gasifier (–)
ΔH_T = heat of formation at temperature T (kJ/mol)