Chapter 11

Direct Reduction and Smelting Processes

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11.1 Introduction

During the past century, many efforts were made to develop processes for producing iron for steelmaking that could serve as alternatives and/or supplements to the conventional blast furnace. Many of these projects were stimulated by a desire or necessity to use lower grade ores and available fuels that are unsuitable for the blast furnace. Processes that produce iron by reduction of iron ore below the melting point of the iron produced are generally classified as direct reduction processes and the products referred to as direct reduced iron (DRI). The processes that produce a molten product, similar to blast furnace hot metal, directly from ore are classified as direct smelting processes. In some of the more ambitious projects, the objective is to produce liquid steel directly from ore and these processes are classified as direct steelmaking processes. These broad categories are clearly distinguished by the characteristics of their respective products, although all of these products may be further treated to produce special grades of steel in the same refining or steelmaking process.

While the blast furnace is expected to remain the world's chief source of iron units for steelmaking as long as adequate supplies of suitable coking coals remain available at competitive cost, DRI contributes 3–4% of the world's total ironmaking capacity with approximately 42 million annual tonnes (Mt/y) of DRI capacity having been installed in the free world as of 1997. As of the end of 1996 there were 15.5 Mt/y of direct reduction capacity under construction, which is expected to increase world DRI production to 55 Mt/y by the year 2000 and 70 Mt/y by the year 2005.

The major part of DRI production is used as a substitute for scrap in the electric arc steelmaking furnace (EAF). DRI derived from virgin iron units is a relatively pure material which dilutes contaminants in the scrap and improves the steel quality. The availability of low cost scrap and the high cost of energy restrict the use of DRI in most highly industrialized countries. Direct reduction processes are favored in those locations with abundant reserves of inexpensive natural gas, non-coking coals and/or hydroelectric power, and that have access to suitable iron ores or agglomerates. This chapter presents a review of the alternative processes that have achieved some measure of pilot or commercial success, with emphasis on those now in commercial operation.

11.2 Historical Development and Background

Sponge iron provided the main source of iron and steel for many centuries before the blast furnace was developed around 1300 A.D. As described in Chapter 1, sponge iron was produced in relatively shallow hearths or shaft furnaces, using charcoal as both fuel and reductant. The product of

Table 11.1 Classification of Processes

Direct Reduction Processes—Reducing Gas Generated Externally from the Reduction Furnace (Section 11.3)	Direct Reduction Processes–Reducing Gas Generated from Hydrocarbons in the Reduction Furnace (Section 11.4)
Gas-based Processes	**Coal-based Processes**

Shaft Processes
    Midrex
    HYL Process
    Purofer
	Rotary Kiln Processes
	    Krupp-CODIR
	    SL/RN
	    DRC
	    ACCAR/OSIL

Fluidized Bed Processes
    Fior/FINMET
    Iron Carbide
    Circored
	Shaft and Hearth Processes
	    Kinglor-Metor
	    Fastmet
	    INMETCO

Reduction Smelting Processes (Section 11.5)	**Inactive Processes of Historical and Developmental Interest (Section 11.6)**
COREX Direct Iron Ore Smelting (DIOS) AISI Direct Steelmaking HIsmelt Romelt Cyclone Converter Furnace (CCF)	Höganäs Process[a] Wiberg-Soderfors Process Sumitomo Dust Reduction Process (SDR) Sumitomo Pre-reduction Method Process (SPM) Kawasaki Process Armco Process Nippon Steel Process High Iron Briquette Process (HIB) (a) Still operating for metal powder.

these early smelting processes was a spongy mass of coalesced granules of nearly pure iron inter-mixed with considerable slag. Usable articles of wrought iron were produced by hammering the spongy mass, while still hot, to expel most of the slag and compact the mass. By repeated heating and hammering, the iron was further freed of slag and forged into the desired shape.

All of the methods whereby low carbon wrought iron can be produced directly from the ore are referred to as direct processes. After the development of the blast furnace, which made large quanti-ties of iron having a high carbon content available, low carbon wrought iron was produced by refining this high carbon material. Because two or more steps were involved, the processes came to be known as indirect processes. Direct methods are still in use and have never been wholly abandoned even by the most advanced nations. The ease with which iron ores are reduced makes the direct processes appear simple and logical, primarily because the reduction takes place at relatively low temperatures.

In modern times, sponge iron has found increasing use in various industrial processes other than in the manufacture of wrought iron. The iron produced in sponge form has a very high surface area and is used in the chemical industry as a strong reducing agent. It is chemically much more active than steel or iron in the form of millings, borings, turnings or wire. Sponge iron may be produced as a granular material or as a sintered mass, depending upon the methods of manufacture. In the purified granular form, commonly known as powdered iron, it is used in the manufacture of many useful articles by the techniques of powder metallurgy. The iron powders are compacted by pres-sure into the approximate shape of the finished article, then sintered at 950–1095°C (1740–2000°F) in furnaces provided with a protective atmosphere to prevent oxidation and finally pressed or machined to their final shape. Iron powders are produced not only by direct reduction

of iron ores or oxides using solid carbonaceous reducing agents and gaseous reducing agents such as carbon monoxide and hydrogen, but also by electrolytic processes and by thermal decomposition of iron carbonyl, $Fe(CO)_2$.

Today sponge iron is referred to as direct reduced iron (DRI). Most DRI production is melted in electric arc furnaces for steelmaking. Minor amounts may be charged to the ironmaking blast furnace. The attempts to develop large scale direct processes have embraced practically every known type of apparatus suitable for the purpose including pot furnaces, reverberatory furnaces, regenerative furnaces, shaft furnaces, rotary and stationary kilns, retort furnaces, rotary hearth furnaces, electric furnaces, various combination furnaces, fluidized bed reactors and plasma reactors. Many reducing agents including coal, coke, graphite, char, distillation residues, fuel oil, tar, producer gas, coal gas, water gas and hydrogen have also been tried. To facilitate classification and discussion of the different processes, broad categories are shown in Table 11.1. Processes employing reducing gases generated externally from the reduction furnace are described in Section 11.3. Processes in which the reducing gas is generated from hydrocarbons in the reduction furnaces are described in Section 11.4. Direct smelting processes that were developed are also listed in Table 11.1 and are reviewed in Section 11.5. Section 11.6 covers inactive processes of historical and developmental interest.

11.2.1 Definition and Terms

11.2.1.1 Chemical Reactions

The reduction of iron ore by direct processes is accomplished by the same chemical reactions that occur in the blast furnace:

$$3Fe_2O_3 + H_2 = 2Fe_3O_4 + H_2O \qquad (11.2.1)$$

$$3Fe_2O_3 + CO = 2Fe_3O_4 + CO_2 \qquad (11.2.2)$$

$$Fe_3O_4 + H_2 = 3FeO + H_2O \qquad (11.2.3)$$

$$Fe_3O_4 + CO = 3FeO + CO_2 \qquad (11.2.4)$$

$$FeO + H_2 = Fe + H_2O \qquad (11.2.5)$$

$$FeO + CO = Fe + CO_2 \qquad (11.2.6)$$

$$3Fe + CO + H_2 = Fe_3C + H_2O \qquad (11.2.7)$$

$$3Fe + 2CO = Fe_3C + CO_2 \qquad (11.2.8)$$

$$CO_2 + C = 2CO \text{ (Boudouard Reaction)} \qquad (11.2.9)$$

$$H_2O + C = CO + H_2 \qquad (11.2.10)$$

$$FeO + C = Fe + CO \qquad (11.2.11)$$

$$3Fe + C = Fe_3C \qquad (11.2.12)$$

Where the reduction is carried out below approximately 1000°C (1832°F) the reducing agents are CO and H_2 and the DRI produced will be porous and have roughly the same size and shape as the original iron ore particle or agglomerate. The metallic iron also absorbs carbon according to reactions 11.2.7 and 11.2.8. DRI from the gas-based processes typically contains 1–2.5% C as cementite.

Above approximately 1000°C (1832°F) gaseous reduction of solid iron oxides continues, however, carbon now reacts with CO_2 and H_2O according to reactions 11.2.9 and 11.2.10 to form CO and H_2 and renew the reducing potential of the gas. The resultant net reaction 11.2.11 is an important mechanism for processes that produce DRI directly from coal without prior gasification of its fixed carbon.

At approximately 1200°C (2192°F), considered to represent close to the upper limit for the direct reduction process, a pasty, porous mass forms. Above this temperature the metallic iron formed will absorb any carbon present with resultant fusing or melting of the solid even though the melting

point of pure iron is 1530°C (2786°F). The direct smelting processes all operate with product temperatures higher than 1300°C (2372°F) because carbon is absorbed rapidly and a liquid hot metal forms that can be handled effectively during subsequent processing.

11.2.1.2 Effectiveness of Reduction Reactions for DRI

The terms associated with measuring the effectiveness of the reduction reactions for DRI are defined as follows:

1. Percent total iron (Fe_T) is the total iron in a sample, including metallic iron and iron chemically combined with other elements such as oxygen.

$$Fe_T = \frac{\text{weight of iron}}{\text{total weight of sample}} \times 100$$

2. Percent metallic iron ($Fe°$) is iron chemically uncombined and as cementite (Fe_3C).

$$Fe° = \frac{\text{weight of metallic iron}}{\text{total weight of sample}} \times 100$$

3. Percent metallization refers to that portion of the total iron present as metallic iron.

$$\text{Metallization} = \frac{\text{total weight of metallic iron}}{\text{total weight of iron}} \times 100 = \frac{Fe°}{Fe_T} \times 100$$

4. Percent reduction refers to the oxygen that has been removed in producing the DRI in reference to initial Fe as Fe_2O_3.

$$\text{Reduction} = \frac{\left(\dfrac{\%O}{\%Fe_T}\right)_{Fe_2O_3} - \left(\dfrac{\%O}{\%Fe_T}\right)_{DRI}}{\left(\dfrac{\%O}{\%Fe_T}\right)_{Fe_2O_3}} \times 100$$

$$\text{Reduction} = \left[1 - 2.327 \times \left(\frac{\%O}{\%Fe_T}\right)_{DRI}\right] \times 100$$

This definition gives the state of the iron contained in the DRI. Confusion can arise as some investigators define percent reduction in reference to the initial oxide state of the ore which may include Fe_3O_4 and FeO. In that case, percent reduction defines the amount of reduction performed in the DRI process.

5. Percent gangue refers to the non-iron compounds (SiO_2, Al_2O_3, CaO, MgO, etc.) in the DRI.

$$\text{Gangue} = 100 - \%O \text{ with Fe} - \%Fe° - \%C$$

2 DRI Quality

ortions of iron oxide and gangue in DRI above certain minimums increase the power require-
 in the EAF compared to an equivalent quantity of scrap. A portion of the iron oxide reacts
 carbon in the EAF to produce metallic iron and carbon monoxide according to reaction

11.2.11, which is endothermic. Furthermore, the gangue and the associated flux require energy for melting. However, a metallization that is too high decreases both the fuel efficiency and the productivity of the direct reduction process, and some of the gangue in DRI substitutes for slag building agents normally present in the EAF charge.

The metallization of DRI normally ranges between 90–95% depending on the process and on the reducibility of the original iron oxide. Based on the level of metallization, the carbon content of the DRI is controlled in some direct reduction processes to be between 1–2.5% to facilitate reduction of FeO by carbon during melting. This increases iron recovery and the CO generated promotes foamy slag practice in the EAF. The carbon as Fe_3C is more desirable than loose carbon fines or soot which may not be useful to the process. Any iron oxide from DRI that is not reduced in the EAF enters the slag.

The gangue in DRI that substitutes for normal slag building agents does not penalize the EAF operation and is usually between 2–4%. However, the actual amount depends on the proportion of silica (SiO_2) in the gangue which must be fluxed and on the percentage of iron units in the charge that is derived from DRI.

Another impact on the operation of the EAF pertains to fines in the DRI which affect iron recovery, increase the dust loading and contaminate electrical parts. The allowable fines is usually approximately 5% at −5 mm.

A major advantage of DRI in EAF operations is the absence of contaminating residuals such as copper and tin. Phosphorus, manganese and vanadium contained in the ore remain in the gangue in the gas-based processes and are usually not a factor in DRI–EAF steelmaking operations. The sulfur content of DRI is also relatively low, depending on the amount of sulfur in the fuels and reductants used. Sulfur contents of DRI vary from less than 0.005% in the direct reduction processes employing sulfur-free gas to approximately 0.02% in direct reduction processes employing sulfur-bearing coal and limestone together with iron oxide in the charge mix. In the latter case, coals containing more than approximately 1.5% sulfur probably necessitate desulfurization during steelmaking.

11.2.3 Special Precautions

DRI, whether in the form of particles, pellets, or molded briquettes, tends to revert to the oxide state when exposed to natural environments. The large pore surface area of DRI makes it susceptible to spontaneous reoxidation. Spontaneous reoxidation is undesirable because of the possible temperature rise during storing or shipment. Even if ignition can be prevented, DRI will gradually lose metallization because of reoxidation due to weathering. When the oxygen in the gas in contact with the DRI becomes depleted, the DRI also reacts with moisture to produce hydrogen gas. This could produce a flammability or explosion hazard in confined spaces. Because of the potential hazards, the loss in value, and the possible harmful effects during subsequent melting, producers and users usually protect DRI from contact with water.

The stability and behavior characteristics of DRI depend on a number of variables affecting the pore surface area. These include the mineralogy of the oxide ore or agglomerate, reduction temperature, time at the reduction temperature, gas composition, age and previous history of exposure and handling. As one example, the DRI stability increases with reduction temperature because the pore surface area decreases as the temperature is raised. Other methods of improving the stability involve making the iron in the pore surfaces unavailable for reoxidation or decreasing the pore surface area mechanically.

To prevent the deterioration of DRI, the industry has studied the techniques used for the protection of structural metals from corrosion. Impervious coatings have been developed utilizing oxide passivation, chemical solutions, oily organic rust inhibitors, paints, etc. Midrex has patented the Chemaire® process, a sodium silicate air passivation treatment.

While some coatings potentially contaminate steel, passivation produces an iron oxide coating consistent with the steelmaking process. Passivation involves treatment with a mildly oxidizing gas at

temperatures below 540°C (1000°F) for the formation of a tightly bonded magnetite layer that is essential for the protective process to work. Passivation to increase stability from spontaneous combustion in a dry atmosphere can be achieved by oxidation without measurable loss of metallization. Given sufficient time, the internal surface area of porous DRI pellets or lumps also acquires the oxide coating which tends to increase the stability of fines generated in subsequent handling. Some autogenous passivation occurs during aging of DRI in a protected ambient atmosphere. Passivation is usually performed between 90–200°C (200–400°F) with various gases, including combustion gases and reducing gas containing less than approximately 2% oxygen.

At the current state of the art, passivation or economical coatings acceptable in the subsequent melting process are not completely reliable. Incidents of heating of DRI in the holds of ships have been reported. The problem may be intensified by contamination with seawater or by lack of uniformity in the DRI as produced or treated. To avert possible catastrophes, an emergency supply of inert gas should be available for ship holds or enclosed storage bins. Such storage areas could be flooded with the inert gas should the temperature rise above a safe minimum or should an explosion hazard arise. The inert medium should probably be nitrogen rather than flue or exhaust gas, which contains carbon dioxide. It is possible that carbon dioxide can be a source of oxygen for the oxidation of DRI, however, the generation of poisonous carbon monoxide is the more important concern.

While coating procedures attempt to protect the iron in the pores from reoxidation, hot briquetting decreases the pore surface area of DRI. In the FIOR process, the –2 mm product is briquetted while the material is hot and malleable. The individual briquettes attain densities greater than 5.0 g/cm³. This DRI product, now referred to as hot briquetted iron (HBI), can be stored in the open prior to loading, can tolerate a fine spray of fresh water to control dust accumulation and can be unloaded under less restricted weather conditions. Still, HBI requires adequate surface ventilation and should not be loaded if the briquette temperature exceeds safety standards. Ship holds or bilges that are used to store the briquettes should be dry and weatherproof. Also, because the briquettes may deplete oxygen in the storage compartments, caution should be exercised when entering.[1] The hot molded DRI concept has also been extended to processes producing DRI in the form of larger coarse particles or pellets.

11.3 Direct Reduction Processes with Reducing Gas Generated Externally from the Reduction Furnace

In this section, representative process descriptions and simplified process flowsheets are given for each major direct reduction process to illustrate the types of equipment used and to describe the flow of materials through the plant.

In most direct reduction processes described in this section, natural gas is reformed in a catalyst bed with steam or gaseous reduction products from the reduction reactor. Partial oxidation processes that gasify liquid hydrocarbons, heavy residuals and coal are also discussed. The reformer and partial oxidation gasifier are interchangeable for several of the direct reduction processes. A summary of the gas generation schemes facilitates the process descriptions.

11.3.1 Methods to Produce Reducing Gas

### 11.3.1.1 Reformed Natural Gas[2]

Reducing gas rich in carbon monoxide and hydrogen is produced by reforming natural gas with steam in catalyst-filled reformer tubes according to the reaction (for saturated hydrocarbons):

$$C_nH_{(2n+2)} + nH_2O = nCO + (2n + 1)H_2 \qquad (11.3.1)$$

High steam/carbon ratios were once required to prevent carbon formation in the reformer tubes and concomitant catalyst deterioration. The resultant gas had to be cooled to condense excess water vapor and reheated in a second step before introduction to the reducer. Modern catalysts now permit

operation with steam/carbon ratios approaching stoichiometric to produce gas with 95% carbon monoxide and hydrogen. Lower steam usage and elimination of the reheating step saves energy with this operation, sometimes referred to as one-step reforming.

In the natural gas based reducing processes that recycle part of the reducer offgas through the reformer, additional carbon monoxide and hydrogen is generated according to the reaction:

$$C_nH_{(2n+2)} + nCO_2 = 2nCO + (n+1)H_2 \qquad (11.3.2)$$
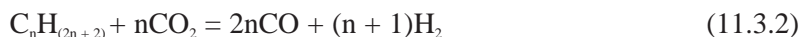
The conversion efficiency of reformers is very high as measured by the approach to equilibrium of the gaseous products. Methane breakthrough increases with pressure and decreases with increasing temperature and increasing steam/carbon ratio. The efficiency of conversion in a reformer is contingent on the use of a sulfur-free feedstock to prevent poisoning and deactivation of the catalyst. Reformers generally employ sulfur guards such as activated carbon absorbent. Reducer offgas used in reforming may contain sulfur from the iron oxide feed. This sulfur may be removed by first using the gas to cool the product which absorbs the sulfur.

Regarding the reformer energy balance, reactions 11.3.1 and 11.3.2 are endothermic and take place at 850–1000°C (1560–1830°F). The required energy is supplied by recuperation from the flue gas in a convection section of the reformer and by radiant burners which heat the reformer tubes. For one-step steam reforming the natural gas is proportioned approximately 40% for fuel and 60% for process gas. Reducer offgas, which contains carbon monoxide and hydrogen, may substitute for a portion of the fuel required.

11.3.1.2 Partial Oxidation for Gasification of Hydrocarbons[3]

The partial oxidation of fuels, especially coal or oil, is effected in gasifiers by reaction with a gas containing a high percentage of oxygen:

$$C_nH_m + (n/2)O_2 = nCO + (m/2)H_2 \qquad (11.3.3)$$

Compared to the reformer reactions, partial oxidation produces less gas per unit of hydrocarbon. For example, a unit of methane gives four units of gas in reaction 11.3.1 compared to three units of gas in reaction 11.3.3. However, additional fuel is needed to sustain the reformer reactions 11.3.1 and 11.3.2; reaction 11.3.3 is exothermic. Excess oxygen is required to retard soot formation and the resulting formation of carbon dioxide and steam provides even more energy to the system. The high energy level of partial combustion processes allows the use of steam or recycled offgas from the reducer to replace a part of the gaseous oxygen to increase gas production and moderate the temperature of the gas produced. The gas may be enriched by passing it through a bed of hot coke or by mixing it with processed offgas from the reducer. In the latter case, equipment is used to condense moisture and remove CO_2 from the reducer offgas. Because of the smaller amount of hydrogen derived from steam, partial oxidation processes produce a gas with a lower hydrogen to carbon monoxide ratio than does steam reforming of natural gas. The process employs fuels of higher molecular weight containing more carbon than methane. Example ratios of hydrogen to carbon monoxide in partial oxidation gasifier gas vary from approximately 0.4 to 1.2 in fuels ranging from bituminous coal to light oil.

The various gasifiers employ alternative concepts comprising entrained beds, fluidized beds, packed beds, partial combustion burners and plasma generators. Various equipment members are protected by a water cooling system incorporated into the steam plant to recover the energy. The units may produce a liquid slag or a dry ash dust which must be separated from the gas in cyclones. Limestone may be mixed with sulfur-bearing fuel to effect desulfurization in the gasifier. In another method of desulfurization, the product gas may be passed through a bed of dolomite. It is desirable to perform the gas cleaning steps while the gasifier gas is hot. This saves fuel associated with reheating the gas and precludes carbon deposition which may occur at intermediate temperatures in the gas reheater. In an alternative to gas desulfurization, the DRI absorbs the sulfur in the reducer and desulfurization is performed in the subsequent steelmaking operation.

Table 11.2 Midrex DRI and HBI Installations

Name of Plant	Capacity (10³ tonnes/yr)	No. of modules	Start-up year	Product
Middle East/North Africa				
QASCO, Umm Said, Qatar	400	1	1978	DRI
Hadeed I, Al-Jubail, Saudi Arabia	800	2	1982–83	DRI
Khuzestan Steel, Ahwaz, Iran	1200	3	1985–92	DRI
ANDSK I, El Dikheila, Egypt	716	1	1987	DRI
LISCO, Misurata, Libya	1100	2	1989–90	DRI
Hadeed II, Al-Jubail, Saudi Arabia	650	1	1992	DRI
NISCO, Mobarekeh, Iran	3200	5	1992–94	DRI
ANDSK II, El Dikheila, Egypt	800	1	1997	DRI
Sub-total	**8866**	**16**		
Asia/Oceania				
Amsteel, Labuan Island, Malaysia	650	1	1984	HBI
Essar I and II, Hazira, India	880	2	1990	HBI
Essar III, Hazira India	440	1	1992	HBI
Ispat Industries, Dolvi, India	1000	1	1994	DRI
Hanbo, Asan Bay, South Korea	800	1	1997	DRI
Sub-total	**3770**	**6**		
Latin America				
Siderca, Campana, Argentina	400	1	1976	DRI
SIDOR I, Matanzas, Venezuela	350	1	1977	DRI
Acindar, Villa Constitucion. Argentina	600	1	1978	DRI
SIDOR II, Matanzas, Venezuela	1275	3	1979	DRI
Caribbean Ispat 1 and 2, Point Lisas, Trinidad	840	2	1980–82	DRI
Caribbean Ispat 3, Point Lisas, Trinidad	1360	1	1999	DRI
OPCO, Puerto Ordaz, Venezuela	830	1	1990	HBI
Venprecar, Matanzas	660	1	1990	HBI
Comsiqua, Matanzas, Venezuela	1000	1	1998	HBI
Sub-total	**7315**	**12**		
North America				
Georgetown Steel, South Carolina, USA	400	1	1971	DRI
Sidbec-Dosco Ispat I, Contreceur, Canada	400	1	1973	DRI
Sidbec-Dosco Ispat II, Contreceur, Canada	600	1	1977	DRI
IMEXSA, Lazaro Cardenas, Mexico	1200	1	1996	DRI
Tuscaloosa, Mobile, Alabama, USA	800	2	1997	DRI
AIR, Convent, Louisiana, USA	1200	1	1997	DRI
Sub-total	**4600**	**7**		
Europe				
Ispat Hamburger Stahlwerke, Germany	400	1	1971	DRI
OEMK, Stary Oskol, Russia	1667	4	1983–88	DRI
Sub-total	**2067**	**5**		
Africa				
Delta Steel, Aladja-Ovwian, Nigeria	1020	2	1982	DRI
Saldanha Steel, Saldanha Bay, South Africa	804	1	1998	DRI
Sub-total	**1824**	**3**		
Grand Total	**28,442**	**49**		

11.3.2 Shaft Furnace Processes

11.3.2.1 Midrex Process[4,5]

The Midrex process was developed by the Surface Combustion Division of Midland-Ross Corporation in the mid-1960s. The Midrex Division became a subsidiary of Korf Industries in 1974. Midrex was subsequently acquired by Kobe Steel, Ltd. in 1983.

The first commercial Midrex plant was installed near Portland, Oregon and started production in 1969. The plant included two shaft reduction furnaces of 3.4 m (11.2 ft) inside diameter and had a total capacity of 300,000 tonnes per year. The average energy consumption of this early plant was approximately 15 GJ/tonne (12.9 MBtu/ton) of DRI. Many difficult engineering and operating problems were solved during the first several years of operation of this plant that contributed significantly to the design, construction and operation of larger Midrex plants throughout the world.

By the end of 1999, 49 Midrex modules having a total capacity of 28.4 Mt/y are expected to be in operation, Table 11.2. The largest module has a capacity of 1.2 Mt/y; the reduction shaft in this plant has an inside diameter of 6.65 m (21.8 ft).

The Midrex direct reduction flowsheet is shown in Fig. 11.1. The main components of the process are the shaft furnace, the gas reformer and the cooling gas system. The temperature and composition of the gas to the shaft furnace is controlled to maintain optimum bed temperature for reduction, degree of metallization, carbonization level and efficient utilization of the reducing gas.

The direct reduction furnace is a steel vessel with an internal refractory lining in the reduction zone. The solids flow continuously into the top of the furnace through seal legs. The reduction furnace is designed for uniform movement of the burden through the furnace. The cooled DRI is continuously discharged through seal legs at the bottom of the furnace. Inert gas is injected into the seal legs to prevent escape of process gases. The DRI is screened to remove fines that are usually briquetted to provide a usable DRI product.

Reducing gas, approximately 95% H_2 and CO, enters the furnace through a bustle pipe and ports located at the bottom of the reduction zone at 760–927°C (1400–1700°F) and flows countercurrent
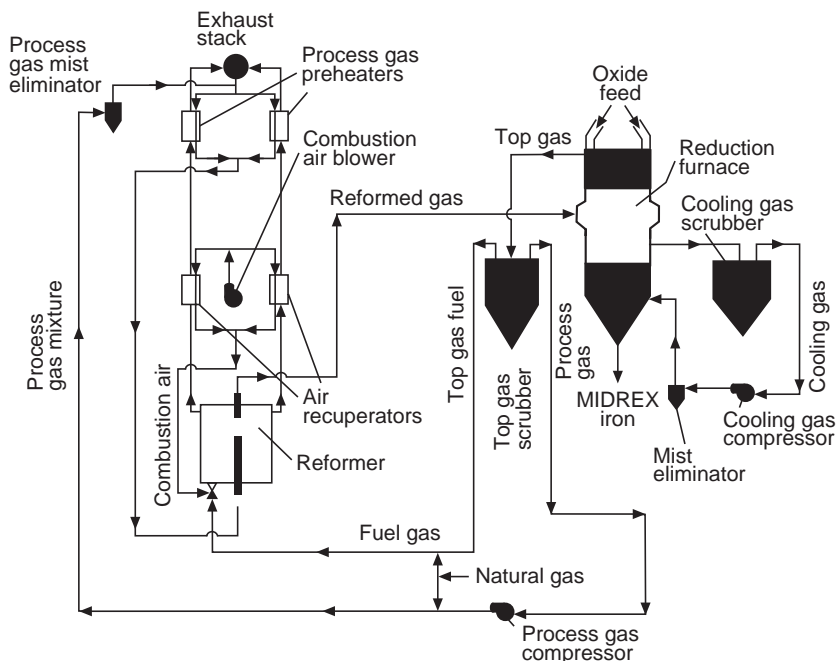

Fig. 11.1 Midrex standard direct reduction process flowsheet.

to the descending solids. The shaft offgas containing approximately 70% CO plus H_2 flows from the shaft outlet to a scrubber where it is cooled and scrubbed to remove dust. Most of the top gas is recompressed, enriched with natural gas, preheated to approximately 400°C (750°F) and piped into the reformer tubes where it is reformed to carbon monoxide and hydrogen according to reactions 11.3.1 and 11.3.2. The hot reformed gas containing approximately 95% CO plus H_2 is then recycled to the shaft.

The excess top gas provides fuel for the reformer burners. Hot flue gas from the reformer is used in the heat recuperators to reheat combustion air for the reformer burners and to preheat the process gas before reforming. The addition of heat recuperators to these gas streams has enhanced process efficiency, helping to decrease annual fuel usage to a reported low figure of 11.4–11.6 GJ/tonne (9.8–10.0 MBtu/ton) of DRI.

Cooling gases flow countercurrent to the burden in the cooling zone of the shaft. The gas then leaves at the top of the cooling zone, flows through the cooling gas scrubber and is compressed, passed through a demister and recycled.

In an alternative flowsheet, shaft offgas is used for cooling prior to introduction into the reformer. The DRI absorbs sulfur in the top gas and prevents sulfur poisoning of the catalyst.

11.3.2.2 HYL Process[6,7,8,9]

In the HYL process developed by Hojalata y Lamina S.A. (Hylsa) of Monterrey, Mexico, lump ore and fired pellets are reduced in fixed-bed retorts by reformed natural gas. The first commercial HYL plant was installed at Monterrey and started production late in 1957. This plant has a capacity of 200 tonnes per day of DRI and the reactors are approximately 2.5 m (8.2 ft) in diameter and hold approximately 15 tonnes of ore in a 1.5 m (4.9 ft) deep bed. The reactors in the more recent plants are 5.4 m (17.7 ft) in diameter and 15 m (49.2 ft) high. Design capacity is approximately 1900 tonnes per day of DRI having an average reduction of approximately 90%. The energy consumption in the most recent plants is 14.9 GJ/tonne (12.8 MBtu/ton) of 90% reduced DRI. In the more recent plants of the HYL II design, high temperature alloy tubes in the reducing gas reheating
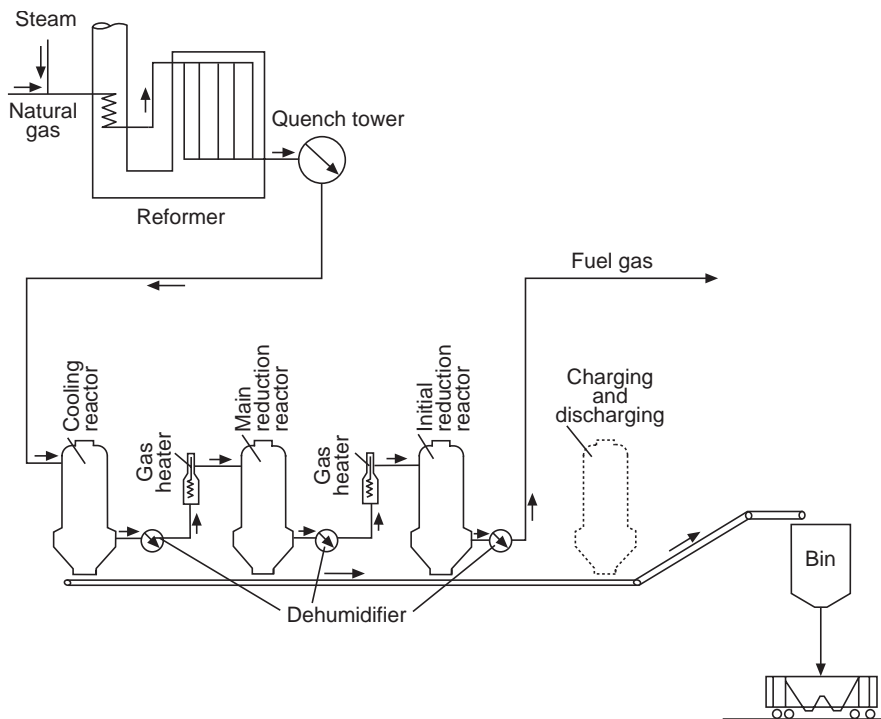

Fig. 11.2 HYL process flowsheet.

furnaces are used, which permits heating the gas to high temperatures, and the number of heating furnaces was reduced to two units from the original four units.

In the HYL II process, reducing gas is generated by reforming natural gas with excess steam, Fig. 11.2, to prevent carbon formation and to promote long catalyst life. The reducing section consists of a set of four reactors, three of which are in operation, while the fourth is engaged in discharging and charging operations. The HYL process is a cyclical batch operation, and the three on-line reactors operate in series. Reduction is performed in two stages, an initial reduction stage and a main reduction stage. Cooling, carburization (Fe_3C) and the final adjustment of metallization are performed in the third stage. Each stage takes roughly three hours. An intricate system of valves permits the reactors to be connected in any desired order so that any one reactor can be connected in its correct process stage.

In 1979, a 9,000 tonnes/yr pilot plant investigated moving-bed technology as an alternative to fixed-bed technology. The first moving bed plant was a conversion of the second plant at Monterrey. The four fixed-bed reactors were replaced by the new shaft furnace and the process, then called HYL-III, was started up in Monterrey. The compatibility of the HYL and HYL-III processes made it possible to convert and modernize HYL plants to the new technology. Hylsa 3M in Monterrey and 2P in Puebla, Usiba in Brazil and P.T. Krakatau Steel in Indonesia were converted to HYL-III. Commercial DRI and HBI installations based on HYL technology currently in operation or under construction in Latin America, the Middle East, Asia and Europe are listed in Table 11.3. The flexibility of the process has allowed for numerous changes and upgrades, such that the generational label is no longer accurate. HYL has, since 1997, referred to its technology as simply the HYL process.

Fig. 11.3 shows a general diagram of the HYL process, which has the flexibility to produce three different product forms, depending on the specific requirements of the user. Cold DRI is normally used in adjacent meltshops close to the direct reduction facilities. This product is also shipped and exported safely on a routine basis, following approved guidelines to avoid reoxidation. Hot briquetted iron can be produced from hot discharged DRI. HBI is generally exported. HBI production can also be used to process low quality, friable iron ores that cannot be used to produce cold DRI. Hot discharged DRI can be pneumatically transported directly to an EAF as HYTEMP® iron.

Reducing gases are generated by self-reforming in the reduction reactor, feeding natural gas as make-up to the reducing gas circuit and injecting oxygen at the inlet of the reactor. The partial oxidation of natural gas with oxygen in a combustion chamber generates reducing gases and energy
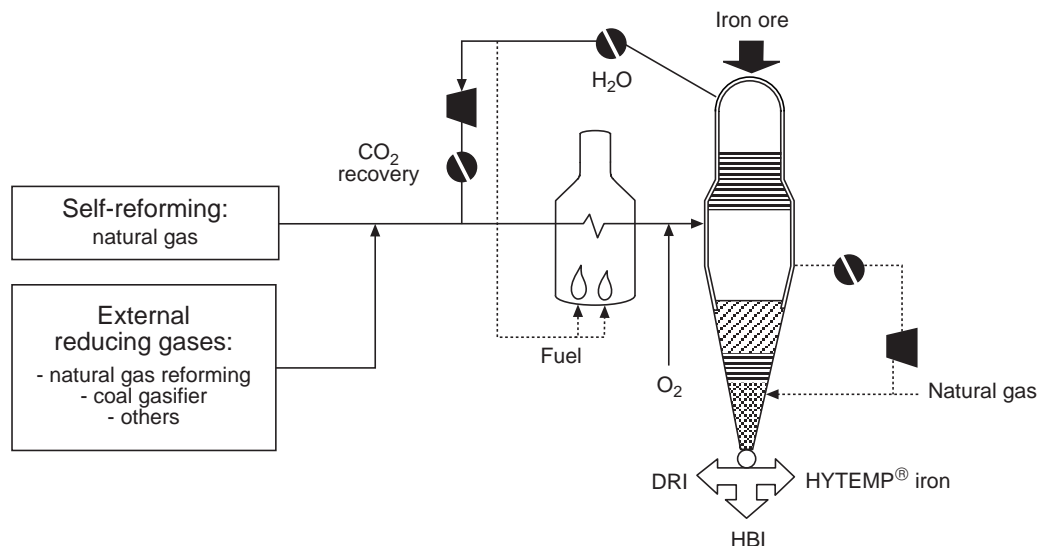

Fig. 11.3 HYL process diagram.

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Table 11.3 HYL Installations

Name of Plant	Capacity (10^3 tonnes/yr)	No. of modules	Start-up year	Product
Middle East				
SEIS I and II, Khor-Al Zubair, Iraq	540	2	1979	DRI
SEIS III and IV, Khor-Al-Zubair, Iraq	940	2	1988	DRI
ASCO I, Ahwaz, Iran	340	1	1993	DRI
ASCO II and III, Ahwaz, Iron	680	2	1994	DRI
Hadeed III, Al-Jubail, Saudi Arabia	1100	1	1994	DRI
Sub-total	**3600**	**8**		
Asia				
PTKS I and II, Kota Baja, Indonesia	1120	2	1978	DRI
PSSB I and II, Kemaman, Malaysia	1200	2	1993	DRI
Grasim Industries, Raigad, India (HYL-III)	750	1	1993	HBI
PTKS IA and IB, Kota Baja, Indonesia	1360	2	1994	DRI
Sub-total	**4430**	**7**		
Latin America				
SIDOR HI, Matanzas, Venezuela	360	1	1976	DRI
SIDOR HIIA and HIIB, Matanzas, Venezuela	1140	2	1980	DRI
Hysla 2M5, Monterrey, Mexico (HYL-III)	250	1	1980	DRI
SIDOR HIIC, Matanzas, Venezuela	570	1	1981	DRI
Hylsa 3M5, Monterrey, Mexico (HYL-III)	500	1	1983	DRI
IMEXSA 1A and 1B. Lazaro Cardenas, Mexico (HYL-III)	1000	2	1988	DRI
IMEXSA 2A and 2B, Lazaro Cardenas, Mexico (HYL-III)	1000	2	1990	DRI
USIBA II, Salvador, Bahia, Brazil (HYL-III)	320	1	1994	DRI
Hylsa 2P5, Puebla, Mexico (HYL-III)	630	1	1994	DRI
Posven, Matanzass, Venezuela (HYL-III)	1500	2	1998	DRI
Hylsa 4M, Monterrey, Mexico (HYL-III HYTEMP)	750	1	1998	DRI
Sub-total	**8020**	**15**		
Europe				
Lebedinsky, GOK. Russia	1000	1	1998	HBI
Sub-total	**1000**	**1**		
Grand Total	**17,050**	**31**		

required for the reduction. Further cracking and reforming reactions are carried out in the shaft due to the catalytic effect of metallic iron. In addition to the self-reforming of reducing gases within the reduction reactor, the HYL process can use externally supplied sources of reducing gas. Some of the alternative gas sources include gases from natural gas steam reformers and from coal gasification units. Other reducing gas sources may be hydrogen, coke oven gas, gases from smelters-gasifiers, and tail gases from other direct reduction processes.

The components in the reducing gas circuit are the reactor reduction zone, the top gas quenching/scrubbing system, the reducing gas recycle compressor, the CO_2 removal system and the process gas heater. The natural gas stream or reducing gas make-up is mixed with the reducing gas recycle stream from the CO_2 removal system. The reducing gas stream is heated to 933°C in the gas heater. Hot reducing gas is fed to the reduction zone at 5.5 kg/cm^3 and flows counter to the solids.

Offgas from the reactor at about 400°C passes through a quenching/scrubbing system where water produced during reduction is condensed and removed along with most of the dust in the gas. Scrubbed gas is fed to the CO_2 removal system by means of the reducing gas recycle compressor. Decarbonated reducing gas is mixed with the natural gas stream or reducing gas make-up to close the reducing gas circuit.

A belt conveyor transports iron ore pellets and/or lump ores to the top of the reduction tower where an automated system of bins and pressure locks receives the ore at atmospheric pressure in an open bin, pressurizes it in intermediate bins and charges it continuously to the reactor. Oxygen is removed from the iron ore in the upper section of the reactor by the reducing gases and the product is then carburized. A rotary valve at the outlet of the reactor regulates the continuous flow of the charge through the reactor. DRI is removed by automated pressurized bins and pressure locks.

In the production of cold DRI, the cooling gas circuit comprises the reactor cooling zone, the cooling gas quenching/scrubbing system and the cooling gas recycle compressor. The cooling gas is fed at the lower conical part of the reactor, at roughly 40°C, flowing counter to the descending DRI. The gas distribution is uniform throughout, with good contact between the gas and the solid material. The cooling gas leaves the reactor at the upper conical part at about 460°C and is then quenched/scrubbed by means of cooling water. Scrubbed cooling gas passes through the cooling gas recycle compressor for recycle to the reactor after being made up with natural gas injected to the cooling gas circuit for optimum efficiency and control of the cooling and carburization processes.

In the HYTEMP® iron process, hot DRI, at 700°C, is discharged continuously from the reactor to the pneumatic transport system. The product is transported by means of a reducing carrier gas to surge bins located above the meltshop for controlled feeding to the electric arc furnace.

For the production of HBI, hot DRI, at 700°C, is discharged continuously from the reactor to the hot briquetting machines arranged below. The HBI is cooled to about 80° in vibrating cooling conveyors using cooling water, and then discharged to the HBI transport conveyor.

The expected specific consumption figures in the HYL process for two typical production options are presented in Table 11.4.

The self-reforming configuration has been in operation since April 1998 at Hylsa Flat Products Division in Monterrey, Mexico. The plant has a nominal capacity of 700,000 tonnes/yr of DRI. Production during the first six months of operation has been at capacity, with metallization levels of 95% and carbon of 4% as iron carbide (cementite). Carbon levels up to 5.3% have been reached with no loss in metallization or productivity. Use of the product in a 150 tonne DC EAF has reduced overall cost/tonne of liquid steel by eliminating carbon additions, reducing electrical energy due to the faster melting reactions and improved slag foaming.

Table 11.4 HYL Process Consumption Figures

Item	Unit (/tonne DRI)	Self-Reforming 94% Metallization 4%C	Natural Gas-Steam Reformer 94% Metallization 2.2%C
Iron Ore	tonne	1.45	1.45
Natural Gas	Gcal	2.26	2.42
Oxygen	Nm³	36.0	0
Electricity	kWh	100	60
Water Make-up	m³	1.00	1.40
Labor	manhour	0.15	0.20
Supplies and Maintenance	US$	3.15	3.50

11.3.2.3 Purofer Process[10,11,12]

In the late 1950s, Huttenwerk Oberhausen AG of Thyssen, Germany, experimented with the reduction of iron ore on a small pilot travelling grate using reducing gas made from regenerative reforming of natural gas. In 1961, the travelling grate was replaced with a small pilot shaft furnace, and in 1970, a semi-commercial shaft furnace plant with a capacity of 500 tonnes per day was set up at Oberhausen. This plant was originally equipped to generate reducing gas by reforming natural gas with air, but was later changed to reform natural gas with process offgas.

In 1976 and 1977, two Purofer plants were put into operation in Iran and Brazil, each with a capacity of about 350,000 tonnes per year. In Iran, reducing gas is generated by catalytically reforming natural gas with process offgas. In Brazil, the Texaco partial oxidation process generates reducing gas from heavy fuel oil. This plant has been shut down since 1979.

In the Purofer process, the reduction furnace is a refractory-lined shaft with a rectangular cross-section and rounded corners. A double-bell charging system at the top of the shaft feeds pellets and/or lump ore to the system. The shaft tapers outward from top to bottom. Two scraper bars at the shaft base remove the hot reduced product from both sides of the shaft to a collection hopper, which has a gas-tight connection with the furnace bottom. A double lock hopper at the bottom of the collection hopper is used to extract the product into sealed steel containers without exposing the product to air. In Iran, Purofer DRI is transported by containers to a briquetting press. The plant in Brazil was provided for transport of the hot DRI directly to the steelmaking furnace.

The Iran plant has two regenerative catalytic reformers, arranged so that when one is used for reforming, the other is being reheated on a cyclic basis, Fig. 11.4. The reformer feed is a mixture of natural gas and recycled, cleaned furnace offgas. Excess offgas (about one-third of the total) together with natural gas is used for heating the reformers. Heat-up burners are located between the two chambers of each reformer. Part of the combustion air is conducted through a heat exchanger and the products of combustion are finally discharged from the stack.

In the Iran plant, the offgas is cleaned and cooled to remove water vapor, and then mixed with natural gas in stoichiometric proportions. This mixture is fed into one of the catalytic reformers at 1400°C, where it is reformed to 95% H_2 and CO with a mole ratio of H_2/CO of 1.4. The reformer temperature drops as the reforming reaction is endothermic. At 1200°C the mixture of natural gas
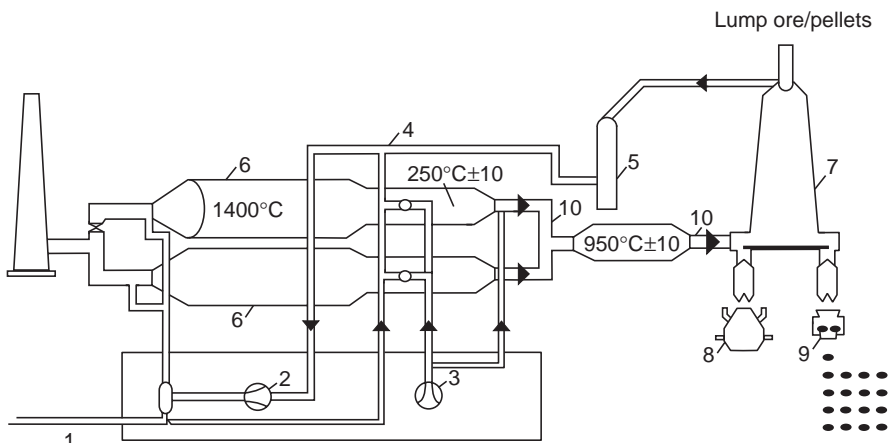

Fig. 11.4 Schematic of the Purofer process based on natural gas reforming.

and process offgas is switched to the other reformer, which has been reheated from 1200°C to 1400°C by combustion of a portion of the offgas during the concurrent heating cycle.

During reforming, sulfur from the natural gas is deposited on the catalyst, but during the reheating the sulfur is oxidized and is removed with the offgas. Each cycle lasts about 40 minutes. The reducing gas leaving the reformer is cooled to 950 ± 10°C.

11.3.3 Fluidized Bed Processes

11.3.3.1 FIOR Process[13,14]

The FIOR (Fluid Iron Ore Reduction) process is a continuous direct reduction process developed by Esso Research and Engineering Company (ERE), a subsidiary of the Standard Oil Company of New Jersey (renamed Exxon Corporation in 1973). Development started in 1955 in cooperation with Arthur D. Little, Inc. and in the early stages it was called the Esso-Little process. After 1960, ERE continued development on their own and pilot plants were subsequently built at Baton Rouge, Louisiana (1961) and Dartmouth, Nova Scotia (1965).

The FIOR flowsheet is shown in Fig. 11.5. The process comprises a series of four fluid bed reactors in series, a reformer to produce fresh reducing gas and a briquetting section to compact the hot product. Dry, sized ore is raised to the top of the system by a pneumatic lift or a skip hoist. The ore is fed continuously into the preheater where it is heated to approximately 800°C (1470°F) and combined moisture is removed. The ore then flows into the reduction reactors in sequence where it is reduced to approximately 92% metallization at 700–750°C (1290–1380°F) and 10 atm. pressure.

Reducing gas enters the final bed and flows counter to the descending ore. Fresh reducing gas is produced by reforming natural gas with steam in a catalytic reformer. Reactor offgas leaving the first reducer is scrubbed and cooled to remove dust and most of the water produced during reduction. Some of this gas is used as fuel and the remainder is compressed and blended with fresh gas from the reformer before the combined stream enters the reducing gas furnace.

Hot product is discharged into the briquetting feed bin and fed to the briquetting press. Pillow-shaped briquettes are formed by compacting the hot fines between pockets in the press rolls. The briquettes are discharged from the press in a continuous sheet, which is broken into individual briquettes in a trommel screen and then cooled by air in a rotary cooler where a thin film of iron oxide forms on the surface of the briquettes rendering them inert.
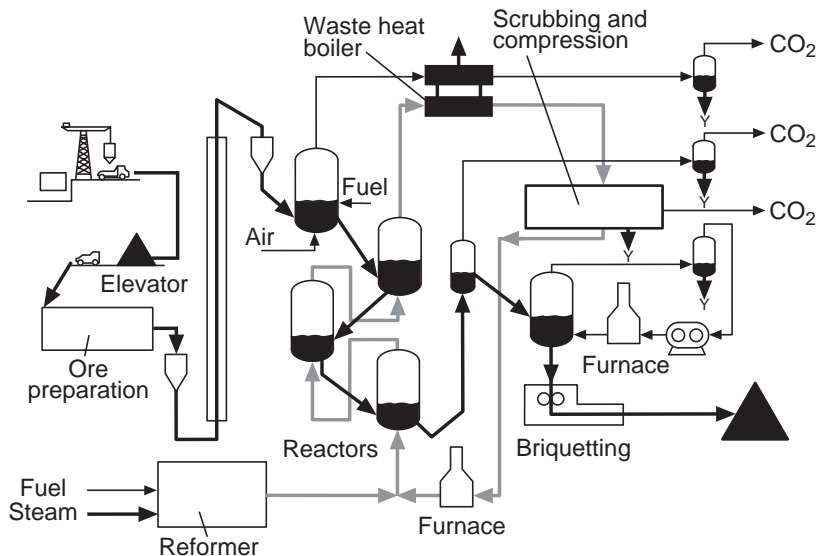

Fig. 11.5 FIOR fluidized bed direct reduction process flowsheet.

The first FIOR plant with a capacity of 400,000 tonnes per year of briquettes was installed at Puerto Ordaz, Venezuela and started operations in 1976 and achieved operation at design capacity in 1984.

11.3.3.2 FINMET Process[15]

The FINMET process, a joint development of Fior de Venezuela and Voest-Alpine, Austria, is based on the FIOR process, described in the previous section. The primary aim of the FINMET development is to improve the energy efficiency of FIOR by using top gas for ore heating and by removing CO_2 from reformed gas as well as recycle gas. Laboratory and pilot plant tests were conducted at Voest-Alpine, Linz in 1993. Operating parameters were established and new equipment tested at the FIOR plant in Venezuela. A 1.0 million tonnes/yr FINMET module based on two reactors and common reformer was engineered.

Two FINMET plants, each to provide 2.0 millions tonnrs/yr, are under construction by Orinoco Iron C.A. in Puerto Ordaz, Venezuela and by BHP DRI in Port Hedland, northwestern Australia. Each plant will have four reactor modules. The Venezuela plant is located adjacent to the FIOR unit. Ore feed for the FIOR and FINMET plants in Venezuela will be San Isidro Fures. The BHP plant will use concentrate from Mt. Newman mines, Pilbara.

The FINMET plant essentially consists of an ore drying section, reducing reactors, steam reformer, recycle gas system and product handling, Fig. 11.6. The ore feed will be dried in fluid-bed driers and charged at the reactor top by buckets conveyors and lockhoppers. Four 4.5 m (14.8 ft) diameter reduction reactors are provided in series. Ore fines flow by gravity countercurrent to the reducing gas. The product is briquetted in double-roll briquetting machines and the briquettes are cooled in a forced-air cooler.

Internal cyclones in the reactor clean the offgas and retain the dust. Recycle gas from the top reactor is quenched and scrubbed in a wet scrubber for use as fuel in the reformer and for recycle, after compression. Make-up gas is supplied from a conventional steam reformer. Two reactor trains will use a common reformer. Reformed gas and recycle gas are passed through a CO_2 removal system, then preheated and fed to the reducers.

The reduction temperature increases from 550°C in the first reactor to 780–800°C in the lowest reactor. Pressure in the reactors is between 11 and 13 bars gauge (150–190 psig). Hot reduced fines
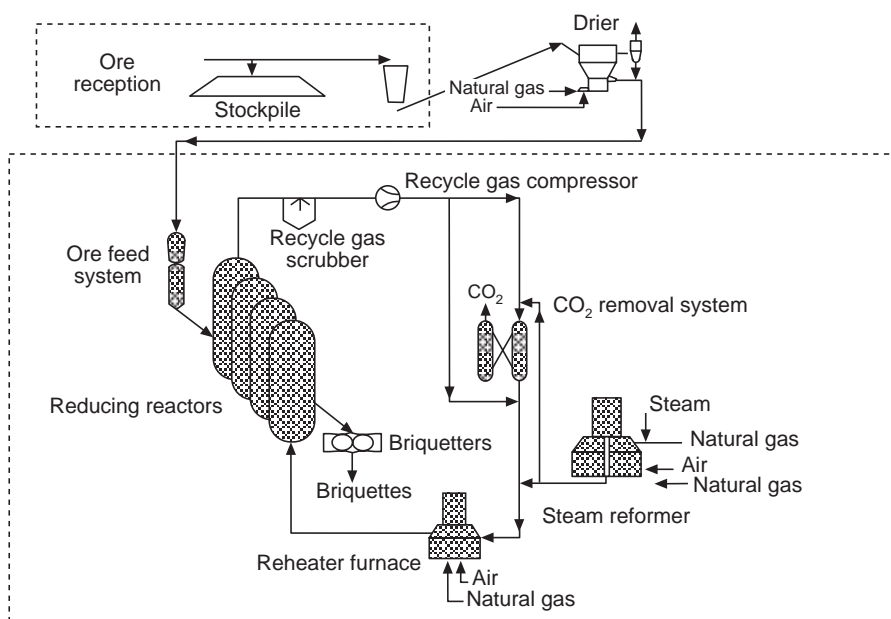

Fig. 11.6 Flowsheet of the FINMET process.

are briquetted to a density over 5 g/cm^3 (0.18 lb/in^3). The briquette strips are separated in revolving drums or trommels and fines generated are recycled to the briquetting machine. Briquettes are cooled on a forced air cooler.

The ore fines for FINMET should contain at least 67% Fe and less than 2% gangue. Fines below 13 mm, with at most 15% through 150 microns, are preferred. Moisture content less than 2% in the ore feed to the reducers is desired.

11.3.3.3 Iron Carbide Process[16,17]

Iron carbide is a high melting point, non-pyrophoric, strongly magnetic synthetic compound obtained in granular form. It contains about 90% total iron and about 7% total carbon. The primary use of the product is as a metallic charge for substitution of hot metal, DRI or steel scrap. The iron carbide process involves reduction of preheated fine iron ore particles (0.1–1.0 mm) in a closed-circuit fluidized-bed reactor by preheated process gas comprising natural gas and cleaned reactor top gas. The reduction takes place at 550–600°C and about 1.8 atmospheres.

The iron carbide process was originally developed at Hazen Research Institute in Golden, Colorado during the 1970s. Iron ore samples from several countries were tested at Hazen. A 50 mm diameter batch reactor was used for the laboratory tests, followed by continuous tests on a 600 mm (24 in.) diameter reactor. The product was successfully converted to steel by MEFOS in Sweden in a basic oxygen furnace in 1979. Additional development work was subsequently carried out in a pilot plant constructed from an idle vanadium plant located in Wundowie, Western Australia, in 1989. The Wundowie reactor was 1.83 m (6 ft) in diameter, operated with a bed depth of about 3.7 m (12.1 ft), and produced about 25 tonnes per day for steelmaking trials worldwide. Nucor Corporation converted iron carbide from the pilot plant at their mill in Darlington, South Carolina using a 32-ton electric arc furnace. An existing gunite tank was used to inject the iron carbide at rates up to 90 kg/min.
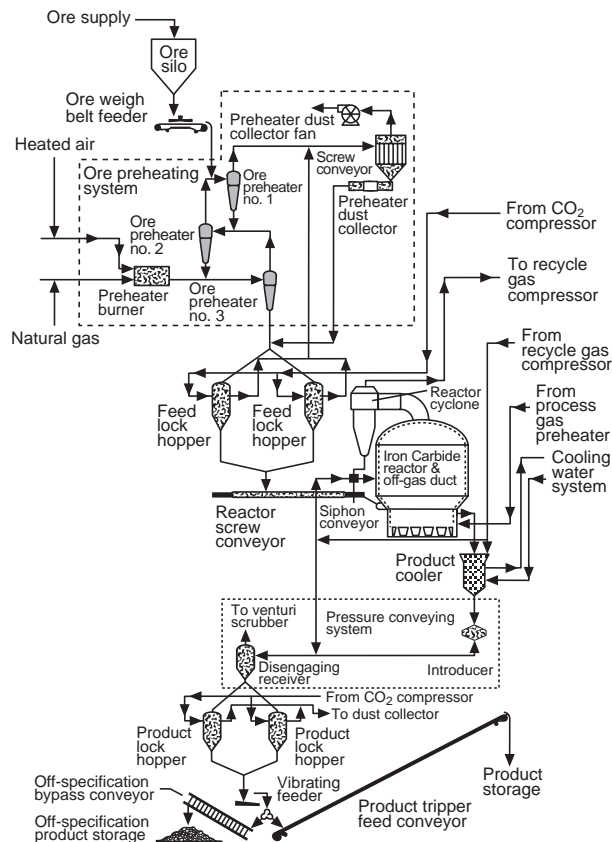

Fig. 11.7 Flowsheet of the iron carbide process.

A schematic flow sheet of the iron carbide process is given in Fig. 11.7. The ore fines are fed from a day bin to a natural gas-fired suspension-type cyclone preheater where they are preheated to about 700°C. The hot ore is fed into the fluidized-bed reactor by screw conveyor. The reduction reactor has an inside diameter of 14.2 m (46.6 ft) and a height of 25 m (82 ft), and is fitted with six baffles that prevent bypassing feed to the discharge. The reactor is designed to contact 550,000 Nm^3 per hour of process gas and 53 tons per hour of ore for production of 900 tons per day of iron carbide.

The offgas is cleaned in a reactor cyclone. A recycle gas heat exchanger recovers sensible heat from the offgas and preheats the process gas. The offgas is further cooled and cleaned in a venturi scrubber and packed towers to 29°C prior to compression and recycle to the reactor. The water slurry leaving the venturi is fed to a clarifier for solids/water separation. A cylindrical, bottom-fired hydrogen plant (reformer) equipped with an MEA desulfurization unit produces desulfurized natural gas and synthetic gas. Process gas from the hydrogen plant mixed with reactor offgas, compressed

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Table 11.5 Iron Carbide Process Inputs

Item of Input	Per ton of Iron Carbide
A. Raw Materials	
Ore Fines	1350 kg
B. Operating Inputs	
Electric Power	255 kWh
Natural Gas	12.6 GJ
Labor	60 Nos.

and preheated to 550°C is fed to the reactor. The inlet gas contains 60% CH_2, 34% H_2, 2% CO and 1% H_2O. The process involves reduction of Fe_2O_3 to Fe_3O_4 to Fe with CO and H_2, followed by carburization of Fe to Fe_3C with carbon from CH_4 and CO.

$$3Fe_2O_3 + 5H_2 + 2\,CH_4 = 2Fe_3C + 9H_2O \qquad (11.3.4)$$

Product typically containing 90% Fe as Fe_3C, 8% Fe as FeO and 6.2% C, is discharged from the reactor bottom by a screw feeder at 400°C and fed through lock hoppers to a cylindrical cooler where it is cooled by water to 70°C.

Nucor Corporation constructed a plant in Point Lisas, Trinidad, West Indies in 1994. Based on the recent performance of the Nucor plant, the major operating parameters for the process are presented in Table 11.5. Sustained operations at the design capacity and product quality were never attained by Nucor Corp. In early 1999 the plant was permanently closed.

11.3.3.4 Circored Process[18]

Circored is a natural gas based process developed by Lurgi Metallurgie GmbH, Germany for producing DRI from iron ore fines. The process uses fluidized beds on a scale adopted by Lurgi for over 20 years for other applications. Development of the process was initiated in the late 1970s with the pilot plant tests conducted at the ELRED plant of ASEA in Sweden. Tests were also carried out in the 3 tonnes/hr circulating fluidized bed (CFB) demonstration unit at Thyssen Stahl in Duisburg, Germany. These tests focused on the treatment of steel plant wastes.
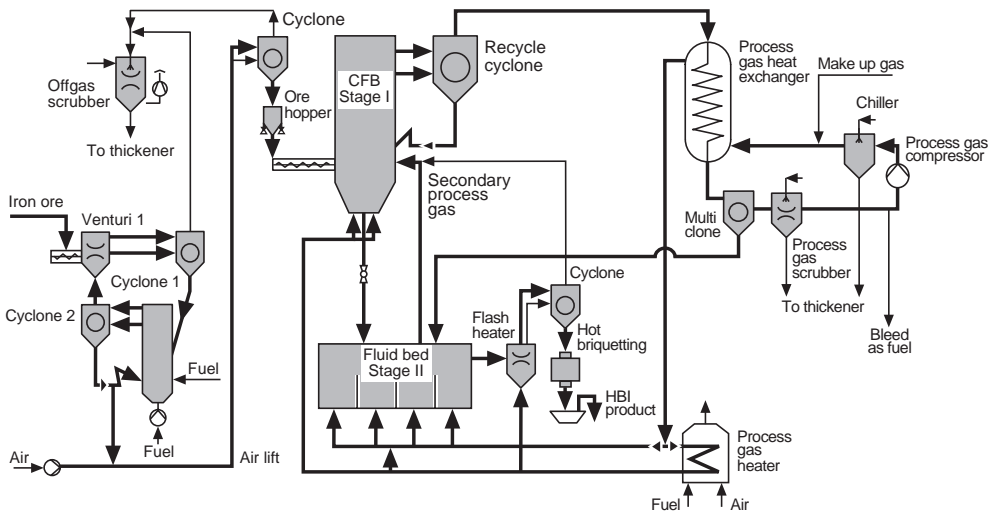
Fig. 11.8 Flowsheet of the Circored process.

Interest in the process was revived in the early 1990s and simulation tests using iron ore fines were conducted in pilot plants at Lurgi's Research Center in Frankfurt. These results and tests on other fluidized bed reactor applications provided scale-up data for the CFB and fluidized bed (FB) reactors for the commercial Circored plant. The process uses a high efficiency CFB reactor for the first reduction stage and a stationary fluidized bed (FB) reactor for final reduction.

Fig. 11.8 is a preliminary flowsheet of the Circored process. Reduction to 60 to 80% is accomplished in the CFB reactor. The CFB provides favorable heat and mass transfer conditions, which results in retention times of 20–30 min.

High gas velocities are used in the CFB, which is designed with a relatively smaller diameter. Good lateral and vertical mixing of solids in the CFB ensures uniform temperature distribution in the reactor and uniform product chemistry. The CFB also enables recycle of the offgas from the final reducer and improves overall gas utilization.

The final reduction takes place in the second stage FB reactor. The reactions here are predominantly diffusion-controlled to produce product with a metallization of 93%. The FB operates with relatively low gas velocities and provides retention times of 2–4 hrs.

Iron ore fines are preheated to about 800°C in a CFB preheater and introduced into the CFB reducer. The pre-reduced material is charged to the compartmentalized FB reducer to complete the reduction. The process operates at a pressure of 4 bars gauge (58 psig) with pure hydrogen as the reductant. Sticking of the fines is avoided by maintaining an operating temperature of about 630°C. The metallized fines are heated to about 680°C in a flash-type heater for briquetting. Hydrogen is also used as the heating medium in the flash heater. Should the Circored unit be installed in a steel plant, direct charging of the metallized fines into the steelmaking facility is an option.

The offgas from the CFB recycle cyclone is passed through a process gas heat exchanger and a multiclone for the recovery of dust for recycle to the last compartment of the FB reducer. The gas from the multiclone is scrubbed and quenched for removal of the remaining dust and condensate produced during reduction. The process gas is then compressed and further dried in a chiller. Fresh make-up hydrogen, produced in a standard steam reformer equipped with a CO_2 removal system, is added after the compression stage. The process gas is preheated to about 800°C before introduction to the reducers. Two-thirds of the fresh gas is used in the secondary FB reactor, and one-third in the CFB reactor.

Typical specific consumption data for the raw materials and other key inputs for the operation of a Circored unit are presented in Table 11.6.

The first commercial Circored unit was built by Cliffs and Associates, Ltd. at Point Lisas Industrial Complex in Trinidad in 1998. The plant, with a capacity of 500,000 tons/yr of HBI, is jointly owned by Cleveland-Cliffs Mining Company, LTV Steel and Lurgie.

Table 11.6 Circored Process Typical Process Inputs

Item of Input	Per ton of HBI
A. Raw Materials and Energy	
Iron Ore Fines (67% Fe)	1470 kg
B. Operating Inputs	
Electric Power	100 kWh
Natural Gas	11.5 GJ
Water	0.6 m^3
Labor	0.23 manhours

11.4 Direct Reduction Processes with Reducing Gas Generated from Hydrocarbons in the Reduction Furnace

11.4.1 Kiln Processes

The rotary kiln is a revolving horizontal cylinder comprising a shell with an internal refractory lining. Seals at each end join the rotating cylinder to the stationary equipment for adding materials and discharging product from the furnace. The furnace is inclined at an angle of 3–4° from the horizontal toward the discharge end so the burden travels through the rotary kiln by rotation and gravity.

Coal, flux and iron oxide are metered into the high end of the kiln and pass through a heating zone where coal is devolatilized, flux is calcined and the charge is heated to the reduction temperature. Iron oxide is reduced in the reduction zone by carbon monoxide by reactions 11.2.2, 11.2.4 and 11.2.6. At the elevated bed temperature, part of the carbon dioxide reacts with the carbon in the coal according to the Boudouard reaction 11.2.9.

A portion of the process heat is usually provided by a burner at the solids discharge end of the kiln. The burner operates with a deficiency of air to maintain a reducing atmosphere in the kiln. Additional process heat is supplied by combustion of coal volatiles and the carbon monoxide from the bed. Combustion air is supplied through ports spaced along the length of the kiln. The airflow is controlled to maintain a relatively uniform temperature profile in the reduction zone and a neutral or slightly reducing atmosphere above the bed. The kiln gas flows countercurrent to the flow of solids.

The iron oxide feed (lump ore or pellets) should fulfill certain requirements regarding chemical composition, size distribution and behavior under reducing conditions in the kiln. The feed should have a high iron content so that costs for further processing in the electric furnace are as low as possible. Sulfur and phosphorus contents should also be low. The minimum size should be controlled at approximately 5 mm. Besides being elutriated from the kiln, fines contribute to accretion buildup (ringing) in the kiln. Reduced fines in the product also reoxidize more rapidly. Ore reducibility, a measure of the time required to achieve a desired degree of metallization under a standard set of conditions, has a strong influence on the capacity of the kiln. The behavior of the ore under reducing conditions is important, especially with regard to swelling and subsequent decrepitation.

Important factors in the selection of coals are reactivity, volatile matter content, sulfur content, ash content and ash softening temperature. Coal reactivity is indicative of the coal's reduction potential. With increased reactivity, the throughput of the rotary kiln can be expected to increase within certain limits. In a simplified explanation, the process seeks a bed temperature where heat transfer to the bed balances the kinetics of reaction 11.2.9. A higher coal reactivity allows the bed to operate at a lower temperature which enhances heat transfer. Consideration should be given to the fact that the volatile content generally increases with reactivity. Because coals with high volatile content generate more gas than can be used in the process for reduction and fuel, the recovery of the sensible and chemical heat contained in the waste gas would have to be considered for overall heat economy. Low-sulfur coals are preferred to minimize sulfur pickup by the DRI. Dolomitic limestone is used as a scavenger for sulfur but increases the heat load. A high ash softening temperature is also desirable to minimize buildup of accretions in the kiln.

The solids discharged from the rotary kiln are cooled, then screened and separated magnetically. DRI fines are briquetted and used for steelmaking. A carbon char is separated and recycled to the kiln to increase fuel efficiency. The offgas passes through a gravity separation chamber that can also serve as an afterburner and is then cooled and cleaned before being released to the atmosphere. Dust from the settling chamber is transported to a waste disposal area.

11.4.1.1 Krupp-CODIR Process[19,20]

The CODIR process of Krupp Industries, West Germany, stems from the original Krupp-Renn process developed in the 1930s to treat high silica ore without the addition of limestone to produce
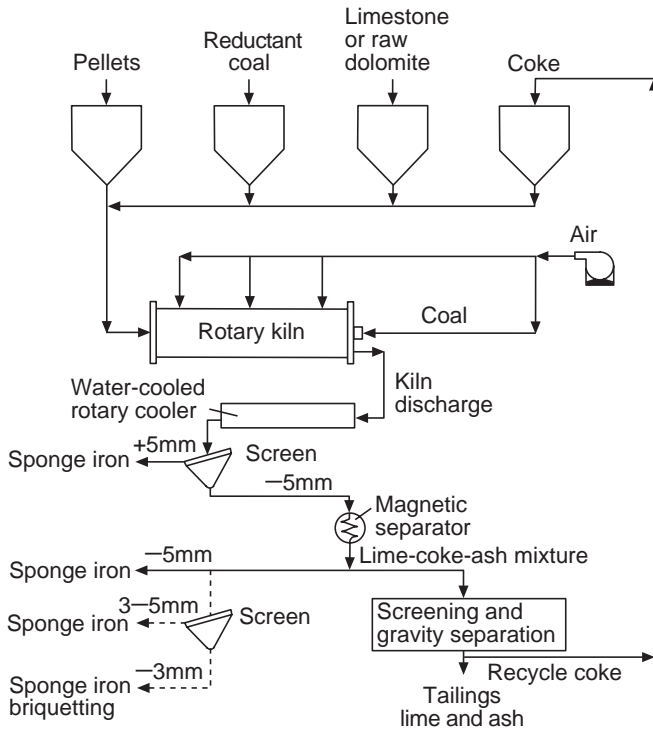
Fig. 11.9 CODIR process flowsheet.

an agglomerated product for subsequent conversion to steel. CODIR operates at a lower temperature than Krupp-Renn and produces a standard DRI product. Limestone or dolomite in the furnace charge absorbs a substantial part of the sulfur introduced with fuel.

A CODIR plant designed to produce 150,000 tonnes per year started operation in 1973 at the Dunswart Iron and Steel Works, Ltd. at Benoni, Republic of South Africa. The reduction kiln in this plant is 4.0 m (14.2 ft) inside diameter and 74 m (243 ft) long. The energy consumption is approximately 15.9 GJ/tonne (13.7 MBtu/ton) of DRI when low-volatile anthracite is used for the reduction coal. The gross energy requirement increases when higher volatile matter coals are used.

A schematic flowsheet of the plant is shown in Fig. 11.9. The reduction kiln is fed continuously at a controlled rate with sized iron ore or pellets, dolomite for sulfur absorption and coal for reduction and energy. As the raw material charge moves to the discharge end it is progressively heated to approximately 1000°C (1830°F) by the hot kiln gas generated by partial combustion of coal with air in a burner at the discharge end of the kiln. The kiln is fitted with six shell mounted fans that supply air to burner tubes that enter through the shell and discharge air along the kiln centerline. This added air burns CO formed during reduction as well as volatiles from the coal and helps maintain the kiln temperature between 905–1050°C over approximately 60% of its length, enabling attainment of a high degree of metallization.

The solids leaving the kiln are cooled to 150°C by water sprays in a rotary cooler. The DRI is then separated from the ash, recycle char and spent dolomite by screening and magnetic separation. The kiln offgas passes through a dust settling chamber and an after burning chamber where the remaining combustibles are burned prior to scrubbing in a conditioning tower, final cleaning in an electrostatic precipitator and exhausting to the atmosphere through an induced-draft fan and clean gas stack. The recovery of energy from the kiln offgas by including a waste heat boiler in the circuit has been adopted in some CODIR plants.

CODIR plants currently in operation use iron ore 5×25 mm in size with an iron content of 66–68%. The reductant coal normally is –10 mm. Coal containing 79% fixed carbon, 10% volatile matter and 10% ash and a calorific value of 29,330 kJ/kg is typical. Pulverized coal is used as the burner fuel, with a typical analysis of 57% fixed carbon, 26.5% volatile matter, and 16.5% ash and a calorific value of 25,140 kJ/kg. Fuel oil and natural gas have also been used as the burner fuel. The desulfurizing agent used is raw limestone or raw dolomite.

The discharge from the cooler is screened into the following four fractions: (1) the +5 mm portion is DRI, which goes directly to product storage; (2) the magnetic portion of the –5 mm, +3 mm fraction is DRI and goes to product storage while the nonmagnetic portion consisting of char is returned to the char bins to be recycled; (3) the magnetic portion of the –3 mm, +1 mm fraction is DRI and goes to product storage while the nonmagnetic portion, a mixture of char and coal ash, is treated by air jigging to separate the char from the ash; and (4) the magnetic portion of the –1 mm fraction is also DRI and is generally briquetted while the nonmagnetic portion is mainly ash and is discarded.

There are four rotary kilns based on the CODIR process that are in commercial operation. The CODIR units at Dunswart Iron and Steelworks in South Africa and Sunflag Iron and Steel Company in India are rated at 150,000 tonnes per year. The two units at Goldstar Steel and Alloys Ltd. in India each have a capacity of 110,000 tonnes per year.

11.4.1.2 SL/RN Process[21,22,23,24]

The forerunner to the SL/RN process, the R-N process (for Republic Steel Company and National Lead Corporation) was developed in Norway, primarily to recover TiO_2 from titaniferous ores for producing pigments. Further development showed that other iron bearing ores could also be treated to produce iron and a pilot

Fig. 11.10 SL/RN direct reduction process flowsheet.

plant was built in the United States. In 1964 Lurgi Chemie acquired the R-N patents and world rights and developed the technology further in cooperation with the Steel Company of Canada, Ltd. (Stelco) to form the SL/RN process.
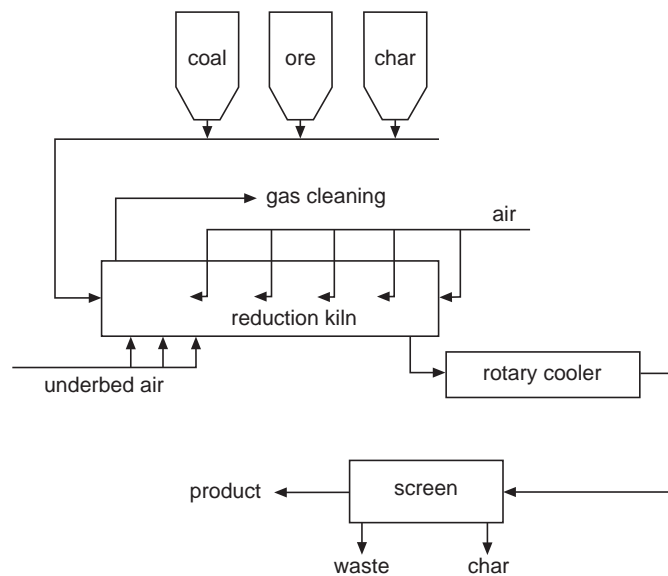
The SL/RN flowsheet is shown in Fig. 11.10. The reduction kiln is very similar to the system used in CODIR and the other coal based plants described later. The major differences between SL/RN and CODIR are that in SL/RN the product is cooled by indirect water sprays on the outside of the cooler, and the kiln offgas is cooled in a waste heat recovery boiler. Indirect product cooling requires a longer cooler than does direct cooling, however, there is no loss of metallization during indirect cooling. In the direct spray cooler, the loss of 0.5–1.0% metallization by reoxidation offsets the economic advantage of a smaller cooler.

By 1998, 15 SL/RN kilns, presented in Table 11.7, were in operation in Brazil, India and South Africa for the production of DRI for steelmaking. Their combined annual capacity is about 1.68 million tonnes.

The SL/RN process is fed with indurated pellets and/or lump iron ore. Iron sands are used at New Zealand Steel, with design modifications to provide efficient operation. A wide range of fuels and reductants including lignite, char, low temperature coke, coke breeze and anthracite coal have been used satisfactorily. Depending on the fuel used, the proportion of the reductant fed through the inlet of the kiln with the oxide feed and the proportion fired through the burner at the kiln exit will be adjusted. With very low-volatile coal, a supplementary fuel such as natural gas or fuel oil is fed through the central burner or through the air tubes to maintain the proper temperature profile in the kiln. Smooth kiln operation is achieved when operating with a relatively high volatile coal charged together with the iron burden through the kiln feed end.

Coals with a high reactivity, low free-swelling index and high ash fusion temperature are preferred. The coal should also be non-coking. A low ash fusion temperature will promote formation of accretions in the kiln. The coal ash composition is also important because a siliceous ash might react with ferrous oxide to form the low melting compound, ferrous silicate, and interfere with the reduction to metallic iron.

The kiln exit gases pass through a settling chamber for coarse dust removal and a two-stage afterburner. Combustible gases are burned in the first stage, followed by fine carbon particles in the second stage. The afterburner chamber temperature is controlled by means of water sprays. In some

Table 11.7 Status of Commercial SL/RN Plants

Name of Plant	Capacity (10³ tonnes/yr)	No. of Modules	Start-up Year
Acos Finos Piratini, Charquedas, Brazil	60	1	1973
Sponge Iron India Ltd., Paloncha, India	60	2	1980 1986
Siderperu, Chimbote, Peru	90	3	1980
ISCOR Vanderbijlpark, South Africa	720	4	1984
Bihar Sponge Iron Ltd., Chandil, India			
– BSIL I	150	1	1989
– BSIL II	150	1	1998
Prakash Industries Ltd., Champa, India			
– PIL I	150	1	1993
– PIL II	150	1	1996
Nova Iron & Steel Company, Bilaspur, India	150	1	1994
Total	**1680**	**15**	

plants, the gas emerging from the afterburner is passed through waste heat boilers, scrubbed and then discharged through the stack.

The DRI is discharged at about 1000°C. An enclosed chute at the discharge removes lumps and transfers the hot DRI to a rotary cooler. The cooler, a horizontal metal cylinder 3.6 m (12 ft) diameter and 40–60 m (131–197 ft) long is immersed in a water trough and has water sprayed on the upper surface. The cooler discharges the DRI into a hopper, which serves as an air lock to prevent air ingress. Trommel screens located at the hopper outlet separate the discharge into fractions which then pass through magnetic separators. Screens downstream separate the nonmagnetic portions.

Recent experiments show that air admitted through ports below the bed in the preheat zone will burn some of the gases that would otherwise leave the kiln unburned. Such a practice would also permit the use of lignite with up to 60% moisture without predrying. This technique would greatly improve the efficiency of the process and arrangements are now being made to install it on several existing SL/RN kilns.

11.4.1.3 The DRC Process[25]

The Davy DRC (Direct Reduction Corporation) process had its origin in the early efforts of Western Titanium Limited of Australia to develop a method for producing synthetic rutile by reducing the iron oxide in ilmenite with coal and leaching the metallized iron from the titanium dioxide. The technology was later used for DRI production. Renamed the Azcon process, a 3.5 m (11.5 ft) diameter, 45 m (148 ft) long kiln based on the process went into operation in 1978 at Rockwood, Tennessee. The plant had a rated capacity of 50,000 tons per year. After the process had been operated successfully for about one year, the name was changed to the DRC process for the newly formed company that took charge of its operation and marketing.
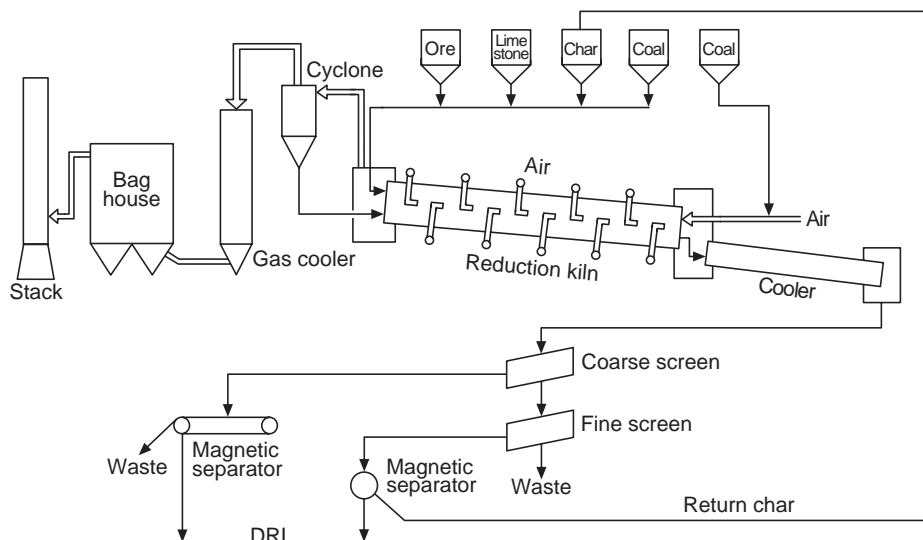
Fig. 11.11 Flowsheet for the DRC process.

The equipment for the DRC process is similar to that of other rotary kiln reduction processes, Fig. 11.11. The iron ore for the DRC process is sized to –20 mm +5 mm. The process requires coal with a high ash fusion temperature and a high char reactivity. Limestone is used as the desulfurizing agent.

In the preheat zone, a mixture of raw materials is heated to the reduction temperature by the controlled combustion of coal volatiles; in the reduction zone, further controlled combustion of coal provides heat for the endothermic reduction reactions. The kinetics of the kiln reactions and, thus, the required kiln residence time to achieve the target metallization are strongly dependent on the temperature. Too much heat supplied at any given location will cause overheating of the bed, resulting in accretion formation. In the DRC process, thermocouple systems are used, one for continuous trend recording, and the other based on quick-response thermocouples to give an accurate reading of both gas and bed temperatures in the kiln. These temperatures are then compared with the desired value and air rates adjusted accordingly. For stable, repeatable control of the air rate each air tube is supplied with its own air fan and venturi. The venturi also contains dampers that give a linear response of the air flow to the damper position over 90% of the damper travel.

The product passes through the rotary cooler and is then screened to separate out residual carbon, ash, and gangue materials. The product is divided into three size fractions, which are treated in a separation train using magnetic separation and air elutriation.

The kiln offgas is at a high temperature and contains residual volatiles which are burned, thus increasing the gas temperatures still further. Typically, the gas is passed through a radiant cooler and cleaned in a bag filter before being released to the atmosphere. Alternately, the sensible heat of the gas can be used to generate steam. If this steam is used for power generation, the process can become a net exporter of electricity. A 150,000 tons per year DRC plant can support a turbo-generator of 11 MW rating, making available approximately 500 kWh of electricity for export per ton of DRI produced after meeting the internal consumption of 100 kWh per ton of DRI.

The first commercial DRC plant was at SCAW Metals, Germiston, South Africa. The SCAW No. 1 kiln went into operation in July 1983. A second kiln was put into operation in November 1989. These kilns are rated at 75,000 tonnes per year. The plant uses the local Sishen lump iron ore and Thabazimbi non-coking coal and is reportedly producing DRI with an average metallization of 93%.

The DRC process was adopted at the first commercial direct reduction installation in China. A 300,000 tonnes per year DRC facility has been installed at the Tianjin plant and consists of two

identical units. The kilns are 5.0 m (16.4 ft) in diameter and 80 m (262 ft) long; the coolers are 3.4 m (11.1 ft) in diameter and 50 m (164 ft) long. The Tianjin plant uses Sishen lump ore from South Africa and Shanxi Shenfu coal. An average product metallization of 92% is obtained. The plant is capable of producing about 72 tonnes per hour of steam.

11.4.1.4 ACCAR/OSIL Process[26]

The Allis-Chalmers Controlled Atmosphere Reactor (ACCAR) produces highly metallized DRI in a ported rotary kiln. Liquid, solid and gaseous fuels, singly or in combination, are used directly in the kiln without an external reformer or gasifying plant. The ACCAR kiln is equipped with an intricate port system and with valving arranged radially around the circumference of the kiln and spaced uniformly along its length for liquid and/or gaseous fuel injection under the bed and for air injection above it. The flow rates through the ports are controlled to maintain optimum temperature profile and gas composition along the length of the kiln. Versatility in the use of fuel permits use of the most economical fuels available.

The ACCAR process was developed by the Allis-Chalmers Corporation, Milwaukee, Wisconsin. Allis-Chalmers conducted direct reduction experiments in 1960 with an unlined metal kiln mounted concentrically inside a larger ceramic lined kiln. Iron oxide pellets were charged into the unlined kiln co-current to a mixture of fuel, steam and air. Pellet reduction and fuel reforming took place in the inner kiln. The gas leaving the inner kiln was mixed with air and burned in the annular space between the two kilns to provide the heat for reduction and reforming. The exhaust gases from the annular space were used to preheat the pellets.

In 1965, Allis-Chalmers developed a laboratory kiln called the controlled atmosphere furnace. Reducing gas from an external source could be introduced through ports in the lower arc of the kiln below the surface of the bed of solids. As the kiln rotated, the reducing gas shut off and air was introduced when the bed reached the upper arc of the kiln. The air burned the excess reducing gas in the freeboard and generated the heat for the process. Experiments showed that hydrocarbons injected below the bed of iron oxide pellets would be reformed into reducing gas within the bed. In 1969, a pilot plant based on the ported kiln concept was constructed in Milwaukee. This kiln, which was 0.6 m (2 ft) in diameter and 7 m (23 ft) in length, was used to obtain operating and design information for scale-up. In 1973, a demonstration plant with an annual capacity of about 35,000 tons of DRI was constructed by modifying an existing kiln at Niagara Metals Limited. In 1976, the SL/RN direct reduction plant at the Falconbridge Nickel Iron Refinery at Sudbury, Ontario was revamped to convert it to the ACCAR System. This plant had an annual capacity of about 200,000 tonnes of DRI. These plants were subsequently dismantled.

The only commercial plant based on the ACCAR system currently in operation is at Orissa Sponge Iron Limited (OSIL) in India. Originally rated for 120,000 tonnes per year, this plant was designed to use both coal and fuel oil. The plant now operates with 100% coal as the reductant and fuel. The process is now called the OSIL process and is marketed by Orissa Sponge Iron Limited. A 150,000 tonnes per year unit based on the OSIL process has been installed by Lloyd's Steel in India and is operational. A second unit of identical capacity is under construction at Lloyd's Steel.

The OSIL process is similar to the other coal-based rotary kiln processes, except that it has a ported rotary kiln, Fig. 11.12. The kiln is refractory lined and has a 2% slope from the charge to the discharge end. The injection ports are located beyond the preheat section of the kiln in rings around the shell. Fuel and air are alternately introduced through the ports by means of valving. A dam at the discharge end maintains the desired depth of material in the kiln. A central burner through the discharge hood which fires axially along the kiln is used for start-up, but can also be used to provide supplementary fuel to the port manifolds. The air for the manifolds is supplied by electrically driven fans mounted on the kiln shell.

Hot DRI is transferred to the cooler through a refractory lined chute. The cooler is an unlined rotating steel cylinder with external water sprays. Following the cooler, there are screens and a magnetic separator. The waste gases from the kiln are cooled and scrubbed before being discharged to the atmosphere through the stack.
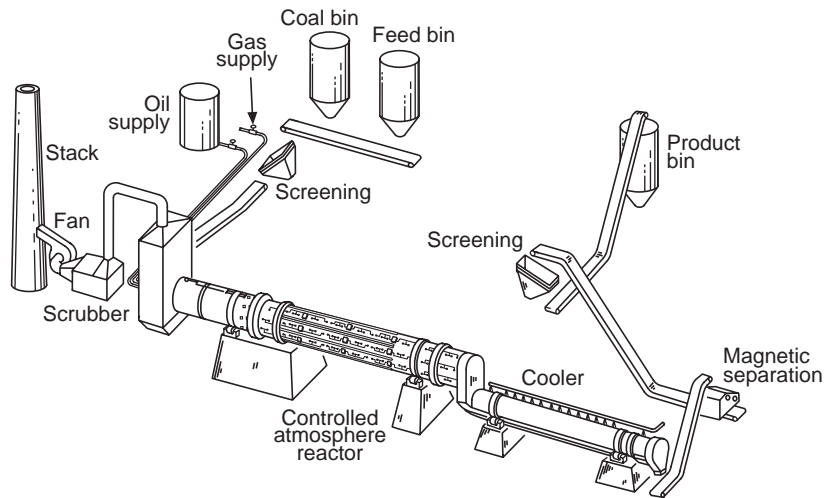
Fig. 11.12 Schematic flowsheet of ACCAR/OSIL process.

The carbon content of the DRI can be controlled at any desired level up to about 2.5% by adjusting the ratio of air-to-fuel along the length of the kiln. The DRI product is cooled to 100°C in the rotary cooler.

11.4.1.5 Other Rotary Kiln Direct Reduction Processes

Besides the processes listed above, a number of other coal based rotary kiln direct reduction processes have been developed and are currently in commercial operation. These processes are similar in equipment, raw materials and operating philosophy. A list of plants based on these processes is given in Table 11.8. The total installed capacity of these plants is 550,000 tonnes per year, primarily in India, with smaller capacities in Peru and South Africa.

11.4.2 Shaft and Hearth Processes

11.4.2.1 Kinglor-Metor Process[27,28]

The Kinglor-Metor process produces iron continuously by heating a mixture of ore and coal in an externally-fired rectangular shaft or retort. Kinglor-Metor shafts are constructed with retort walls of highly conductive silicon carbide. Additional energy is provided by burning some of the carbon monoxide generated during reduction with air in a preheating zone in the upper part of the retort. Fig. 11.13 shows a schematic flowsheet of the process.

A pilot plant comprising two reactors was installed at Buttrio, Italy by Danieli & Cie., SpA and started operations in 1973. The
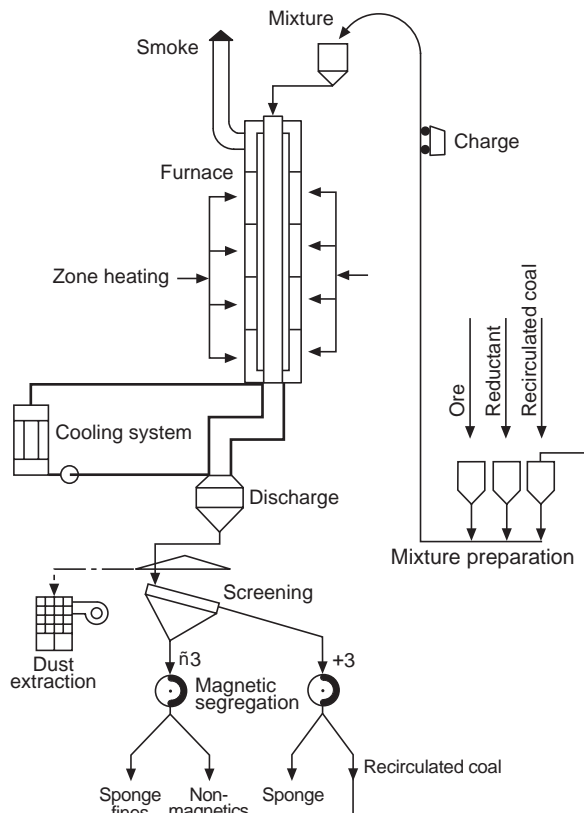

Fig. 11.13 Schematic flowsheet of the Kinglor-Metor process.

Table 11.8 Other Commercial Rotary Kiln Based Direct Reduction Plants

Name of Plant	Capacity (10³ tonnes/yr)	No. of Modules	Start-up Year
SIIL PROCESS			
Bellary Steels Ltd.			
Bellary, India	60	2	1992–93
Hindustan Electro-Graphites Ltd.			
Borai, India	60	2	1992
Kumar Metallurgicals Ltd.			
Nalgonda, India	30	1	1993
Raipur Alloys Ltd.			
Raipur, India	30	1	1993
Tamil Nadu Sponge Ltd.			
Salem, India	30	1	1993
Aceros Arequipa			
Pisco, Peru	60	2	1996
Sub-total	**270**	**9**	
TISCO PROCESS			
Tata Sponge Iron Ltd.			
Ipitata I	120	1	1996
Ipitata II	120	1	1998
Sub-total	**240**	**2**	
DAV PROCESS			
Davsteel			
Cullinan, South Africa	40	1	1985
Sub-total	**40**	**1**	
Grand Total	**550**	**12**	

reactors were vertical shafts of conical shape approximately 11 m (36 ft) high with a top diameter of 0.4 m (1.3 ft) and a bottom diameter of 0.7 m (2.3 ft). The pilot plant operations demonstrated the process to be simple to construct, easy to operate and flexible with respect to feed and reductant requirements.

A commercial plant capable of producing 40,000 tonnes/yr was installed by Ferriere Arvedi & Cie., SpA in Cremona, Italy in 1978. The plant consisted of two identical 20,000 tonnes/yr modules. Each module contained six vertical retorts 13 m (42.7 ft) high, 12.5 m (41 ft) long and 3 m (9.8 ft) wide. Solid fuel requirements of approximately 8.5 kJ/ tonne of DRI and gaseous fuel requirements of approximately 7.9 kJ/tonne were claimed. A plant was also installed in Burma in 1981 and is the only plant still in operation.

11.4.2.2 FASTMET/Inmetco Rotary Hearth Processes[29,30,31]

The FASTMET process, developed during the early 1990s by Midrex Direct Reduction Corporation to provide a coal based process for North American locations, is very similar to the rotary hearth pioneered by Inmetco in Ellwood City, Pennsylvania for treating waste dusts from steel plants and also a proposed process studied by Salem Furnace Company in Carnegie, Pennsylvania.

The nucleus of the FASTMET process originated with the development of the Heat Fast process in the mid-1960s. The Heat Fast process consisted of the following steps: (1) mixing and pelletization of iron ore fines and pulverized coal; (2) drying the green pellets on a grate; (3) pre-reduction on a RHF; and (4) cooling in a shaft furnace.

The next step in the evolution of the FASTMET process was the development of the Inmetco process for reduction of stainless steel mill wastes in 1974. The process consisted of the following steps: (1) mixing and pelletization of mill wastes and pulverized coal; (2) pre-reduction of the green pellets on a RHF; (3) discharge of hot pellets into transfer bins; and (4) charging of hot pellets into a submerged arc furnace.

The process was tested in a pilot plant located at Port Colborne, Ontario, and a commercial unit of 60,000 tons per year capacity was set up in 1978 at Ellwood City, Pennsylvania.

Midrex revived investigations in 1989 and tested a wide range of raw materials in its laboratory in Charlotte, North Carolina. A 2.5 m (8.2 ft) diameter pilot RHF with a capacity of 150 kg/hr of reduced iron was installed in 1991 to simulate the reduction portion of the process. This process simulator was utilized to develop data for the design of an industrial unit.

Based on the successful laboratory and pilot plant tests, Midrex and its parent company Kobe Steel started construction of a FAST-MET demonstration plant in April 1994 at the Kakogawa works of Kobe Steel in Japan. The plant has a production capacity of 2.6 tonnes/hr, and was commissioned in September 1995. The demonstration plant is reportedly operating under stable conditions and several tests have been conducted to develop process parameters for scale-up to a 60 tonne/hr industrial scale unit. Tests have also been made for producing hot briquetted iron (HBI) and for integrating the FASTMET process with DRI melting in an electric arc furnace.

Fig. 11.14 The FASTMET process flowsheet.

K-M Iron, Inc., a joint venture formed by Kobe Steel and Mitsui & Co., have announced plans for a FASTMET facility in Louisiana or Alabama. The plant will consist of two modules, each with a capacity of 450,000 tonnes of DRI per year.

In the FASTMET process, shown schematically in Fig. 11.14, iron ore concentrate, reductant and binder are mixed and formed into green pellets that are dried at 120°C and fed to the rotary hearth furnace. The pellets are placed on a
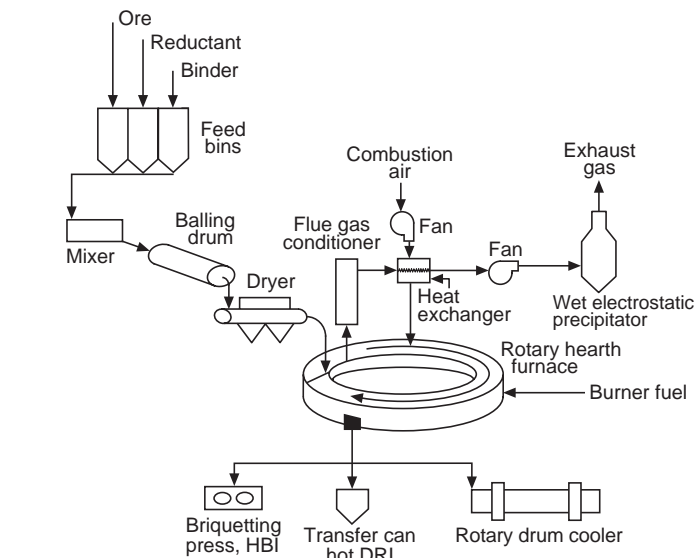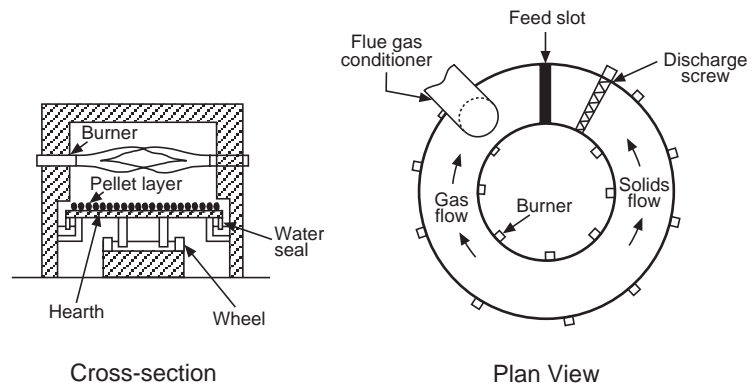

Fig. 11.15 Cross-section and plan view of the rotary hearth furnace.

Table 11.9 FASTMET/Inmetco Typical Process Inputs

Item of Input	Per ton of HBI
A. Raw Materials	
Oxide Feed (68% Fe)	1300 kg
Reductant Coal (75% FC)	380 kg (2.72 Gcal)
Burner Coal (75% FC)	120 kg (1.28 Gcal)
Bentonite (as binder)	21kg
Lime (for flue gas desulfurization)	6 kg
B. Operating Inputs	
Electric Power	60 kWh
Water	1 m^3
Nitrogen	2.5 m^3
Labor	42 Nos.

solid hearth one to two layers deep as shown in Fig. 11.15. As the hearth rotates, the pellets are heated to 1250–1350°C by means of fuel burners firing into the freeboard above the hearth. Reduction to metallic iron is completed in 6–12 min. depending on the materials, temperature and other factors. The DRI is discharged at approximately 1000°C and can be hot charged to an adjacent melter, hot briquetted or cooled indirectly before storage and/or shipment.

The residence time of the pellets in the RHF is 12 min. The hot, reduced iron from the RHF is partially cooled to about 1000°C by a water-cooled discharge screw. The product can be obtained either in the form of cold DRI, hot DRI or HBI. The product can be discharged from the RHF either to a transfer device for conveying the hot DRI directly to an adjacent steelmaking facility, or by gravity to a briquetting system for producing HBI.

The offgas from the RHF flows to a gas handling system, where the SO_2, NO_x and particulates can be reduced to the desired limits. The hot offgas is used for preheating the combustion air for the RHF burners and to supply the heat for drying. The process is designed to recycle all process water. All fines generated in the process are recycled though the feed.

Typical specific consumption data for the raw materials (dry basis) and other key inputs are given in Table 11.9

 A number of other coal based direct reduction processes utilizing the RHF as the reduction reactor have been developed or are under development. The most notable is the Inmetco process which is based on the experience gathered from processing of stainless steel mill wastes at the Inmetco plant in Ellwood City. Mannesmann Demag, Germany currently holds the marketing rights for the Inmetco process.

Iron Dynamics, Inc., a subsidiary of Steel Dynamics, Inc., is establishing a 520,000 tons per year coal-based RHF facility at its mini-mill in Butler, Indiana.[32] This plant is also scheduled for commissioning in 1999.

11.5 Reduction Smelting Processes

In the late 1950s and early 1960s, the concept of producing liquid iron or semi-steel directly from ore fines or concentrates and an inexpensive and plentiful fuel and reductant such as pulverized coal was the subject of considerable pilot plant development activity. These projects, which encountered serious technical, process and/or mechanical problems, were reviewed briefly in the second general reference in the bibliography.

In recent years, interest has been renewed in developing such concepts to commercial viability, stimulated in part by prospects of long term shortages of high quality metallurgical coals, by increasing environmental problems associated with coke ovens, agglomeration plants and other operations ancillary to blast furnace ironmaking and by strong incentives to decrease the high capital requirements associated with blast furnace ironmaking. The most prominent projects representing this renewed interest are the KR process developed by Voest-Alpine AG and their subsidiary Korf Engineering in a pilot plant at Kehl, West Germany (now known commercially as the COREX process), the Direct Iron Ore Smelting (DIOS) process under development in Japan, the AISI direct steelmaking process studied in a pilot plant at Universal, Pennsylvania and the HIsmelt process originally developed in Germany and now being pursued cooperatively by Rio Tinto and Kvaerner Metals.

Although there have been significant worldwide efforts to develop reduction smelting processes based on coal and ore directly during the past decade, no demonstration or commercial plants have been built as of 1998 with the exception of COREX. The basic principles and process parameters have been demonstrated through laboratory and pilot plant studies, however, until larger scale operations that demonstrate the long term performance of processes are established, COREX is likely to remain the only commercial reduction smelting alternative to the blast furnace. The original objectives to go directly to steel or semi-steel in these concepts also remain to be demonstrated even on a pilot scale.

11.5.1 COREX Process[33,34]

The COREX process, shown schematically in Fig. 11.16, is a two-stage operation in which DRI from a shaft furnace like the Midrex and HYL process is charged into a final smelter-gasifier. Reducing gas for the shaft furnace is produced by partial combustion of coal with oxygen in the fluidized bed of the smelter-gasifier. The energy needed to complete the reduction of the DRI and produce the hot metal and slag is provided by the partial combustion. The liquid products are tapped periodically and partially spent offgas from the shaft reducer is exported along with excess gas produced in the smelter-gasifier.

The smelter-gasifier operates at 3–5 bars and comprises an upper fluidized bed zone at approximately 1500°C (2730°F)and a lower melting and liquid collection zone at approximately 1550°C (2820°F). Coal and limestone are injected into the freeboard above the fluidized bed zone where they are heated rapidly to 1000–1200°C (1830–2190°F). The volatile matter is driven off and shattered fixed carbon particles fall into the gasification zone where a gas with high oxygen content is injected through blast furnace-type
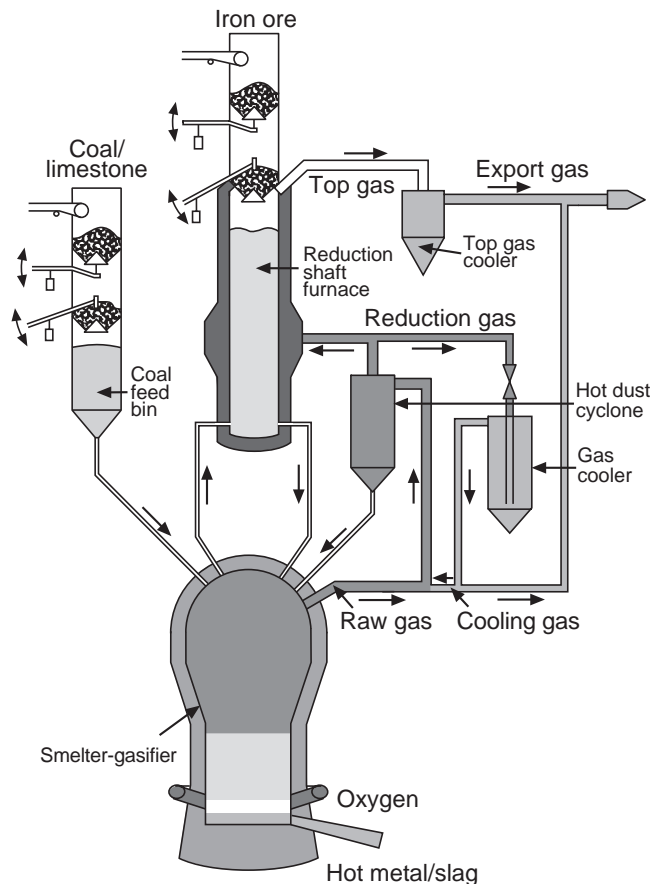

Fig. 11.16 Schematic representation of the COREX process.

tuyeres to burn the carbon to carbon monoxide. The exothermic combustion provides the energy to complete the reduction of the hot DRI and to melt the slag and hot metal. The gas leaving the smelter-gasifier is cooled to 800–900°C (1470–1650°F) and cleaned in a hot cyclone to recycle entrained fines. A portion of the clean gas is then introduced into the shaft furnace as reducing gas containing more than 94% CO plus H_2. The remaining gas is mixed with the cleaned offgas from the shaft furnace and the mixture used as export fuel gas.

COREX uses approximately one tonne of coal per tonne of hot metal, with approximately 45% of the total energy input used in ironmaking and the rest going to export fuel gas. The hot metal produced has carbon and silicon contents similar to blast furnace hot metal; however, the sulfur content is much higher because nearly all of the sulfur in the coal enters the slag and hot metal. In this connection, organic sulfur in the coal gasifies and is absorbed by the DRI and returned to the smelter-gasifier as iron sulfide.

The first commercial scale COREX plant was built for ISCOR in South Africa during 1985 through 1987 by Voest-Alpine Industrialanlagenbau. The plant, rated at 300,000 tonnes per year, was commissioned in November 1989 after testing, modification and operation following the startup in 1987. Since commissioning, the COREX plant at ISCOR has performed very well at the design specifications. A second COREX plant was constructed for POSCO at their Pohang Works in Korea in 1998.

11.5.2 DIOS[35,36]

The DIOS process has been under development in Japan since the 1980s and culminated in a 100 tonnes per day pilot plant at NKK Steel. A schematic flowsheet of the process (pilot plant) is shown in Fig. 11.17. Iron ore is preheated in the first of two fluidized bed reactors in series and pre-
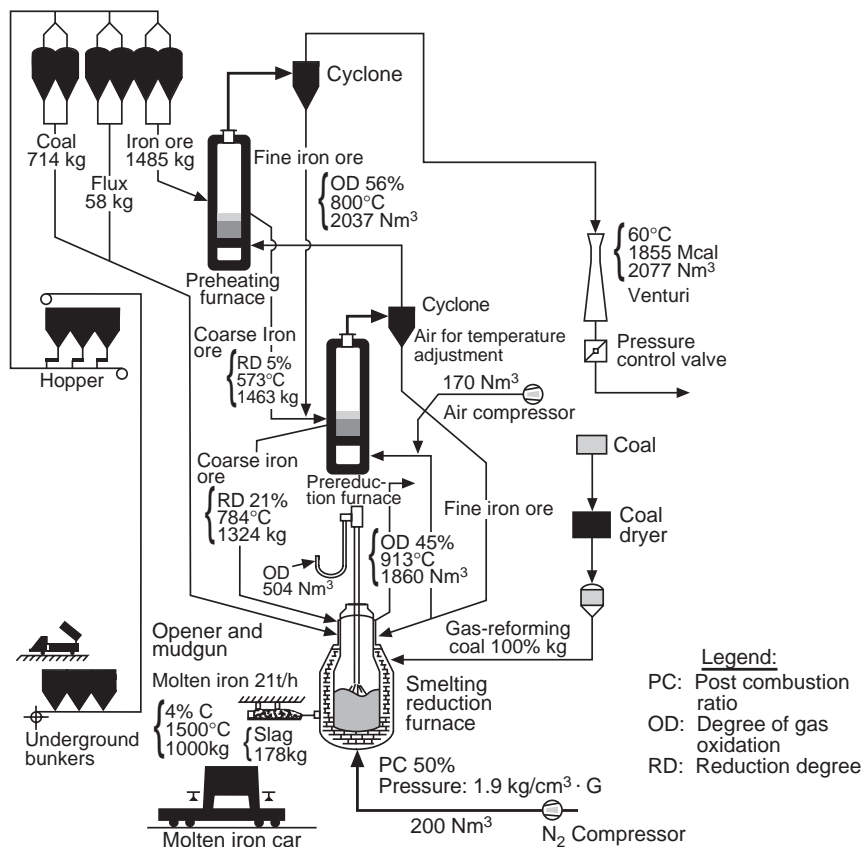

Fig. 11.17 Schematic diagram of the DIOS process.

reduced to 15–25% in the second reactor using cleaned offgas from the smelter. Dust removed from the smelter offgas and fines removed from the gases leaving the fluidized bed reactors are injected back into the smelter. In addition, a small amount of coal fines, on the order of 50 kg per tonne of hot metal production, is injected into the smelter offgas to cool the offgas and provide additional CO and H_2 for pre-reduction. Most of the coal, 600–700 kg per tonne of hot metal, is gravity-fed into the smelter. Oxygen is injected into the smelter for combustion of primary coal and for post-combustion. The oxygen lance is designed to provide both high velocity oxygen for carbon oxidation (hard blow) and lower velocity oxygen for post-combustion in the freeboard (soft blow) simultaneously. The aim post-combustion is approximately 40%, and with pre-reduction to 20%, the coal consumption is roughly 700–800 kg per tonne of hot metal depending on coal type. In general, coal consumption is higher with high volatile coals.

11.5.3 AISI Direct Steelmaking[37]

The AISI process is very similar to DIOS except that AISI is based on iron oxide pellets with pre-reduction to approximately 30% in a shaft furnace similar to that used in the Midrex and HYL process. A schematic flowsheet for AISI is shown in Fig. 11.18 with material balances for typical operations. The AISI process was studied during the 1990s at 5–10 tons per hour to develop design data for a demonstration plant which was never undertaken.

11.5.4 HIsmelt[38]

The cooperative efforts of CRA and Midrex in the early 1990s led to the construction of a 150-tonne per day pilot plant in Australia. The original HIsmelt reactor differed from DIOS and AISI in several major ways, in that it was a horizontal vessel into which coal and pre-reduced ore were injected directly into the metal bath. The final reduction reactions took place in the metal phase; in DIOS and AISI the medium for reduction and heat transfer is a deep slag bath. Also, in the original HIsmelt, devolatilization of the coal in the iron bath caused a relatively large injection of metal
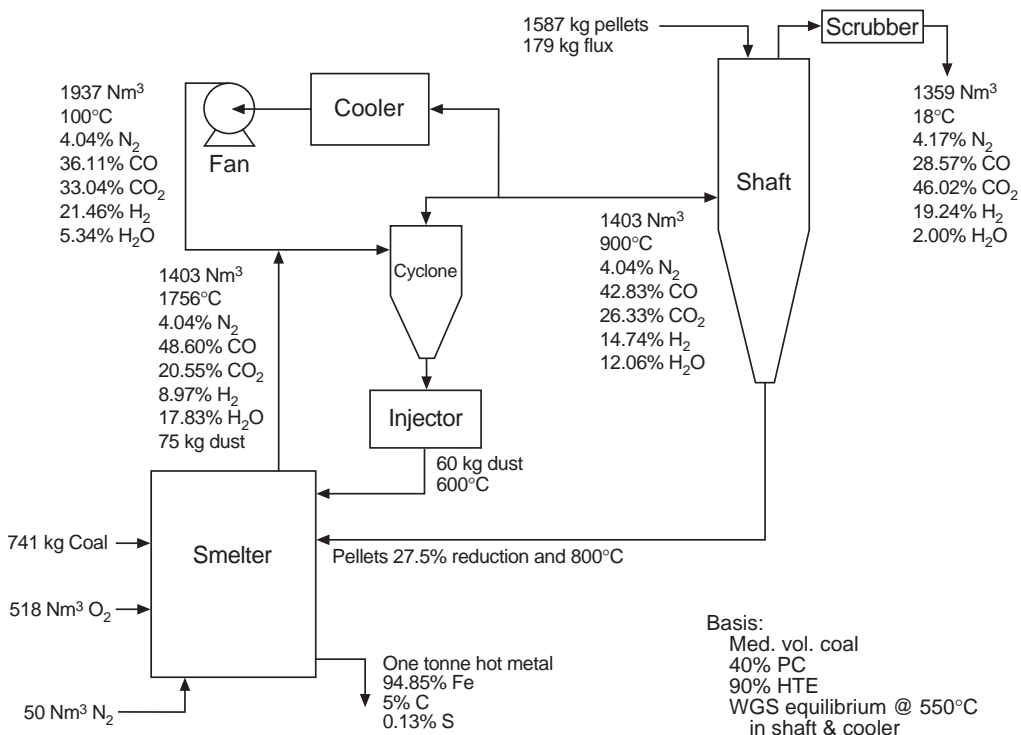

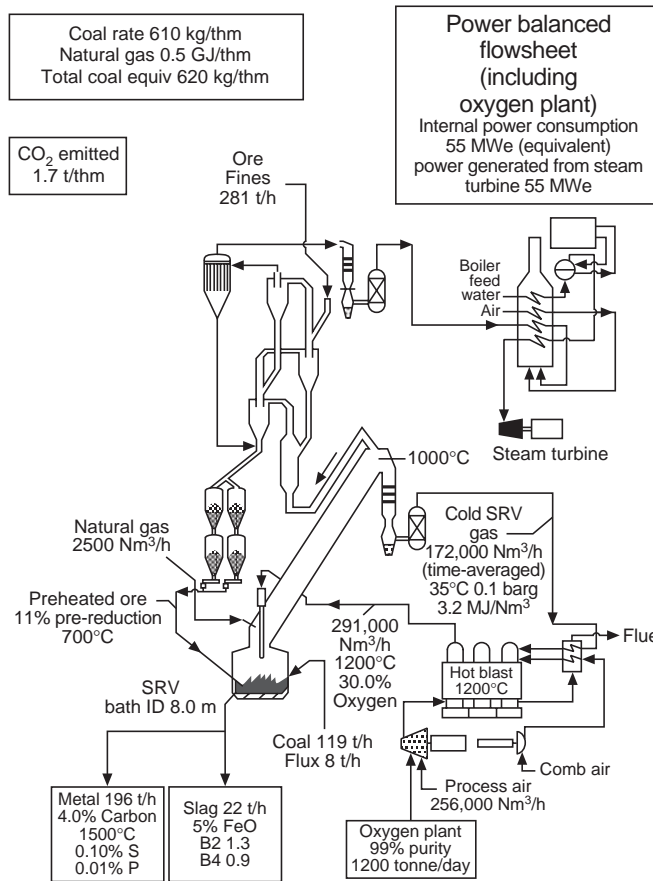**Fig. 11.18** Schematic diagram of the AISI process.

Fig. 11.19 Schematic diagram of the HIsmelt process.

droplets into the freeboard where they picked up energy from post-combustion reactions for transfer back to the bath. In 1998 HIsmelt adopted a more conventional vertical reactor design. Coal and pre-reduced iron ore are injected into the metal phase, where rapid dissolution and smelting occur. Offgassing generates a large liquid fountain with strong mixing of metal and slag. Oxygen enrichment of the hot blast is possible as back-reaction between the hot blast and dissolved carbon in the bath is minimized by the high slag volume in the liquid metal fountain. Post-combustion in the range of 50–75% is routinely achieved. A schematic diagram of the vertical steel refining vessel (SRV) HIsmelt process with inputs and outputs is presented in Fig. 11.19.

11.5.5 ROMELT[39]

The ROMELT process was developed in Russia in the 1980s. A major feature is there is no pre-reduction process step as shown in Fig. 11.20. The smelter has a water-cooled roof and sidewalls in contact with slag and conventional refractories in contact with the metal. A mixture of air and oxygen is injected through two rows of tuyeres. Coal and ore are fed by gravity. The system, simple and robust, has produced hundreds of thousands of tonnes of metal and runs continuously for weeks. ROMELT consumes more energy than other smelting processes due to the lack of pre-reduction and extensive water cooling.

11.5.6 Cyclone Converter Furnace (CCF)[40]

The CCF, initially developed by Hoogovens Staal BV in the early 1990s, is a process for melting and pre-reducing iron ore fines, as feed for a bath smelter. The ore fines, introduced at the top of the vertical furnace, are initially melted and then reduced to 15–20% by the rising smelter offgas.
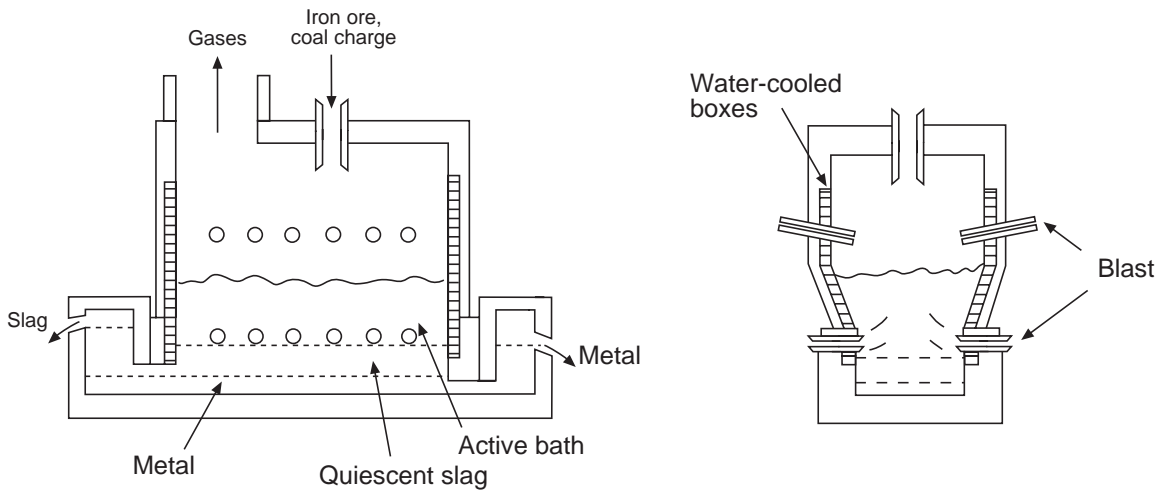
Fig. 11.20 ROMELT process.

Melting occurs due to the high temperatures achieved by combustion of the rising offgases with oxygen introduced through swirl injectors. The injectors promote mixing of the combusted gas with the ore fines for improved heat transfer. Final reduction occurs in a bath smelter. Hoogovens considered using the DIOS-type smelter for a 0.4–1.0 million tonne/yr plant.[35]

11.6 Inactive Processes of Historical and Academic Interest

11.6.1 Höganäs Process[41]

The E. Sieurin, or Höganäs, process was developed at Höganäs, Sweden in 1910 and is still in commercial use. Reduction is carried out in cylindrical silicon carbide saggers consisting of three annular spaces. Iron ore concentrate is charged into the intermediate annular space, and a fine coke and limestone mixture in the outer annular space as well as the central circular space. Saggers are heated in an oil-fired tunnel kiln and air cooled outside the kiln. A vacuum device removes the coke ash and unused coke fines from the sagger. A mechanical device lifts the metallized iron tube out of the sagger and the iron is cleaned and crushed for sale as iron powder. Total retention time of a container in the tunnel kiln is approximately 80 hours. Most of the sponge iron produced is sold as iron powder. The largest Höganäs plant in operation has two tunnel kilns and is capable of producing 38,000 tonnes annually.
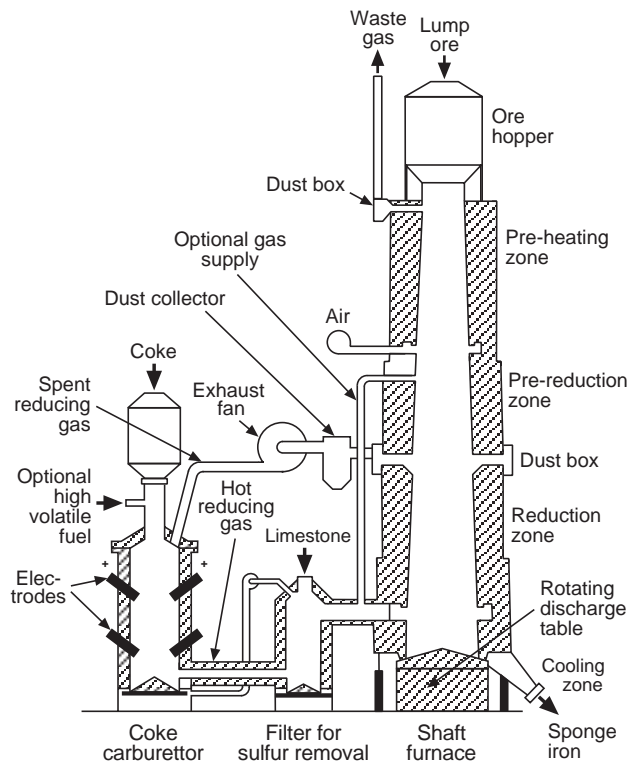

Fig. 11.21 Schematic cross-section showing the principle of operation of the Wiberg-Soderfors process.

11.6.2 Wiberg-Soderfors Process[42,43]

The Wiberg-Soderfors process, Fig. 11.21, was the earliest successful shaft furnace process. The largest commercial installation had a capacity of 18,000 tonnes/yr and went into production in 1952 in Sweden where several other plants were constructed. The largest Wiberg furnaces were 24.4 m (89 ft) tall structures of firebrick and insulation enclosed in a steel shell. Reducing gases (about 25% H_z and 75% CO) were produced in an electrically heated coke or charcoal carburetor and desulfurized in a limestone or dolomite bed. Spent gas, about 75%, leaving the reduction zone was recycled to the carburetor for regeneration. The remainder of the spent gas passed through the pre-reduction zone and was burned with air in a preheating zone at the top of the shaft. The process was abandoned due to low productivity and high cost.

11.6.3 Sumitomo Dust Reduction Process (SDR)[44]

SDR, developed by Sumitomo Heavy Industries Limited, Japan in 1971–72 to recover zinc and lead from steel plant wastes, produced DRI pellets for blast furnace use. A commercial unit, 160,000 tonnes/yr, was built in 1975 at Wakayama Steel Works.

The process consisted of feed preparation using a thickener, dry dust, bins, mixing tank, slurry basins, filter and rotary dryer; pelletizing and preheating using kneading equipment, pelletizing discs and grate-type pellet preheater; and reduction and cooling using a rotary kiln and a quench tank. Waste dusts from sinter plant, blast furnace ore bins, blast furnace top gas, cast house emissions, BOF offgas and melt shop emissions were used as feed. Due to high sulfur content, the DRI was used primarily as a blast furnace charge material and not for steelmaking.

11.6.4 Sumitomo Pre-reduction Method Process (SPM)[44]

SPM was developed by Sumitomo Heavy Industries Limited, Japan to separate zinc from steel plant wastes and recover iron units. A commercial plant, 100,000 tonnes/yr, was put into operation at Kashima Steel Works in 1975.

The SPM process was similar to SDR. In SPM, waste materials were not pelletized prior to charging to the reduction kiln. SPM employed a rotary cooler for product cooling instead of the quench tank used in SDR. The SPM product contained a high percentage of fines, which did not enable significant use in a blast furnace.

11.6.5 Kawasaki Process[45]

Kawasaki process development was initiated in 1963 with the treatment of steel plant waste dusts and sludges. The first commercial unit, 60,000 tonnes/yr, was put into operation at Chiba Works, Japan in 1968. Larger plants, 180,000 tonnes/yr, were set up at Mizushima Works in 1973 and Chiba Works in 1977.

In-plant waste dusts and sludges were blended, pelletized in a disc pelletizer, and dried and preheated in a grate-type unit. A rotary kiln was used as the reduction unit. DRI was directly cooled in a rotary cooler using high-pressure water sprays. Separation of zinc and alkali elements from iron units was satisfactorily achieved and a highly metallized DRI was produced, but its high sulfur content restricted use in steelmaking.

11.6.6 Armco Process[46,47]

This process was developed in a pilot shaft-type direct reduction plant set up by Armco Steel Corporation at Kansas City in 1963. The first commercial scale Armco direct reduction plant went into operation at Houston in 1972. The pilot plant used pebble stoves for generating reducing gas; catalytic steam reforming was used in the commercial plant. The Houston plant operated with pellets

and low sulfur natural gas for generating reducing gas. Natural gas was desulfurized prior to reforming. The reformers were operated with a steam-to-carbon ratio of 1.3.

11.6.7 Nippon Steel Process[48]

Nippon Steel Corporation initiated development of the process in 1969 using a shaft reduction furnace. Reducing gas was generated by partial oxidation of fuel oil. Pilot plant studies were conducted between 1970 and 1973, and a semi-commercial plant, 500 tonnes per day, was put into operation at Hirohata Works in 1974. This plant operated with both lump ore and pellets. Reducing gas at was generated by partial oxidation of light fuel oil using the Texaco Partial Oxidation process.

Renamed NSSC-DR, a 600,000 tonnes/yr commercial plant based on steam reforming was set up at Perwaja Steel, Malaysia in 1981. Due to problems experienced during start-up, the process was abandoned and the plant was converted to the HYL process in 1993.

11.6.8 High Iron Briquette Process (HIB)[49,50]

HIB is a modified version of the Nu-Iron process, developed by U.S. Steel in the 1950s. A commercial HIB plant with a design capacity of 1 million tonnes/yr went into operation in 1971 at Puerto Ordaz, Venezuela to produce 75% reduced briquettes for blast furnace feed. The plant consisted of three separate trains, each with 330,000 tonnes/yr capacity similar to the fluidized bed reducers in FIOR. The plant experienced serious mechanical problems after start-up, necessitating considerable modifications to the ancillary equipment. Stabilized operation of the plant commenced in 1979.

The process operated on fines as the oxide feed and natural gas as fuel and reductant. Reduction was carried out in a two-stage fluidized bed reactor. Steam-methane reformers were used to generate reducing gas. The plant was shut down in the middle 1980s and the three fluidized-bed systems were replaced by a Midrex shaft with a capacity of 830,000 tonnes/yr.

11.6.9 Plasmared Process[51,52]

To apply plasma energy to the direct reduction of iron ore, SKF Steel Engineering AB in Sweden developed a coal gasifier based on plasma technology. The DRI is produced in a shaft furnace similar to a Midrex or HYL process shaft using reducing gas produced in the plasma gasifier. A plant to produce 50,000 tonnes/yr of DRI was installed in Hofors, Sweden in 1981, Fig. 11.22. A coal/water slurry is introduced into the slagging gasifier along with recycled process gas through an injection tuyere where it is mixed with high-temperature oxidizing gas from the plasma generator. Hot gas from the gasifier is passed through a dolomite bed for sulfur removal and introduced into the shaft furnace where DRI is produced from pellets and/or
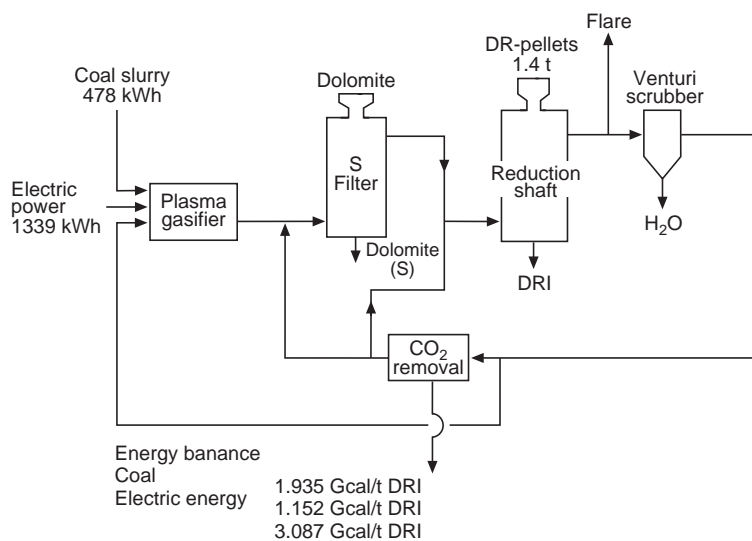
Fig. 11.22 Plasmared flowsheet in Hofors, Sweden.

lump ore. Offgas from the reduction shaft is cooled and cleaned prior to CO_2 removal and recycled to the gasifier.

The Plasmared plant was operated for several years and provided engineering data for a proposed commercial plant located near Stockholm. The project was not implemented due to environmental objections from local agencies and individuals.

11.7 Future Developments[53]

Electric arc furnaces at new steel plants require low residual iron sources to manufacture high end carbon steels made in conventional integrated mills by the blast furnace-BOF process. Reduced iron products will be a commodity and steel companies will look for ways to minimize their cost.

Future developments include:

1. Gas-based merchant plants will continue to be built where natural gas is abundant and cheap.

2. Research and development will continue on direct iron and steelmaking concepts.

3. Direct reduction operations that utilize fine ores will increase due to the lower cost of the ore feed.

4. Steelmakers without access to low-cost gas or ores will invest in DRI plants to control the cost of their iron units.

5. Inferior ores will be reduced and smelted with solid reductants to produce hot metal as charge material to adjacent EAF melt shops to provide high-end products at favorable cost.

6. Research efforts will continue on upgrading abundant, low-grade ores for direct reduction fines.

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