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Recovery and utilization of waste heat in a coal based sponge iron process

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ABSTRACT

The present work is an attempt to conserve energy in coal based sponge iron industry incorporating certain design modifications without disturbing the process technology. A typical sponge iron plant has been investigated to find out the potential areas where energy is being wasted. To recover heat from these areas two design modifications, Case-1 and Case-2, are proposed. Case-1 accounts for preheating of air using waste gas exiting from ESP. However, for Case-2 initially water is heated using hot sponge iron exiting rotary kiln and further hot water is used to preheat air. To compute coal demand of modified designs a model is developed based on heat of reactions, feed preheating, sensible and radiation losses, etc. Preheating of air up to 170 °C for Case-1 reduces coal consumption by 8.7%. Consequently, waste gas generation reduces by 16.7%. Thus, for Case-1 profit is Rs 9.6 million/year. However, for Case-2 preheating of air to 80 °C before entering the kiln reduces coal and water consumption by 7.2% and 96.3%. Consequently, cooling tower capacity is reduced by 37.2%. Due to 27.8% less profit for Case-1 in comparison to Case-2 Case-1 offers higher payback period than that of Case-2. Thus, Case-2 is selected as best proposed design.

1. Introduction

Sponge iron is the metallic form of iron produced from reduction of iron oxide below the fusion temperature of iron ore (1535 °C) by utilizing hydrocarbon gases or carbonaceous fuels as coal. The reduced product having high degree of metallization exhibits a 'honeycomb structure', due to which it is named as sponge iron. As the iron ore is in direct contact with the reducing agent throughout the reduction process, it is often termed as direct reduced iron (DRI).

It is seen that the growth of sponge iron industry in last few years is unremarkable and today India is the largest producer of sponge iron as it covers 16% of global output. According to Sponge Iron Manufacturers Association [1], India has produced around 23 million tonnes of sponge iron in the financial year of 2009–2010. Sponge iron is used as feedstock and a recognized alternative to steel scraps in iron and steel making processes. Due to certain problems inherent in blast furnace and induction furnace process such as the scarcity of steel scraps in international markets, depleting reserves of high quality metallurgical grade coking coal, environmental constraints on coke and sinter plants, and requirement of auxiliary plants with increased capital and operational intensity direct reduction technology has emerged as a potential process for iron and steel making.

Sponge iron is produced primarily both by using non-coking coal and natural gas as reductant and therefore classified as coal based and gas based process respectively. Due to promising availability of coal of 264,535 million tonnes the coal based sponge iron plants share the major amount of its production [2]. At present, there are 118 large and small sponge iron plants operating in India, among them only 3 are natural gas based and the remaining 115 plants are coal based. Among all the available options, rotary kilns have been widely used as a reactor in coal based plants and the important processes applying this technology include: SL/RN, Codir, ACCAR, DRC, TDR, SIIL and Jindal [3–6]. In the present work SL/RN process based plant is selected for energy conservation as it is oldest and most widely applied process. Many investigators considered sponge iron manufacturing process and suggested improvement in that [5,7,8].

With the availability of raw materials, high demand of sponge iron and less payback period, sponge iron industry has emerged as a profitable venture. However, due to lack of proper integration techniques, non-optimal process parameters, high-energy consumption and old running process technology, the industries are facing a setback in market. Further, it is seen that much of the heat generated in the process is lost without being recovered due to lack of heat recovery options. Thus, the energy conservation in sponge iron plants has sought the attention of many investigators.

Eriksson and Larsson [9] carried out the energy survey of sponge iron process and showed that the process is 40% energy efficient and the major loss is through exhaust gases. A few authors [10,11] found that 10–12% energy can be saved by controlled axial and radial air injection, leading to efficient combustion and improved

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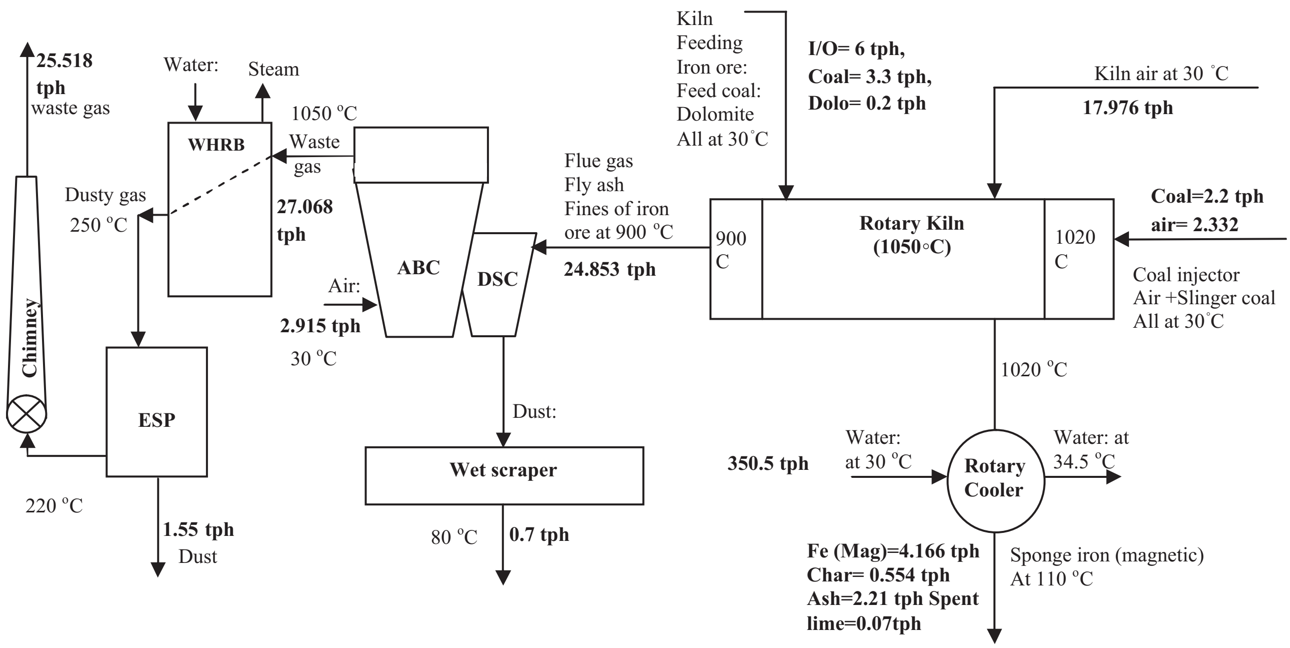

Fig. 1. Process flow diagram of sponge iron process.

heat transfer thus reducing waste gas temperatures. Jena et al. [12] reported their work on a typical plant of production capacity of 350 tpd. They pointed out that the thermal efficiency of the process is 51.3%. Considerable amount of heat is lost in the waste gas which is about 33% of the heat generated in the kiln. The above fact has also been reiterated by many other investigators [13,14]. Hajidavalloo and Alagheband [15] studied the thermal analysis of sponge iron preheating using waste energy of EAF. To improve performance of EAF, a new initiative technique has been introduced in which sponge iron particles are preheated before entering the furnace. Based on the simulation results it was found that energy consumption in the EAF was reduced up to 14%. According to Agarwal and Sood [16] almost 100,000 Nm^3/h of waste gas at 1000 °C are generated from a 100,000 TPA module of DR plant, which in any conventional plant are cooled, cleaned and let off to atmosphere. Out of the total heat generated from coal based DR plant, only 35% of the heat is used for reduction and the rest is lost with the exhaust gases. To tap this heat a waste heat recovery boiler (WHRB) was installed to generate process steam. But this recovery system was also linked with some problems as: (i) kiln operation is governed by process parameter and metallization of DRI and is not dependent on the quality of steam to be generated, (ii) during startup the steam quality is not suitable for turbine application, (iii) any interruption of kiln operation as accretion formation, shut down or unavailability of feedstock disturbs the waste gas generation which in turn effects or stops steam generation, (iv) kiln waste gas contains 35×10^6 to 40×10^6 kg/m^3 of dust particles and (v) the economic viability of the installation of waste heat recovery boiler occurs only when the installed sponge iron capacity is above 300 tpd (tonnes per day). Apart from above, Ulrich and Tondon [17] concluded that the feasibility of a WHRB for one 500 tpd module plant would be questionable. It should be viable for the plants beyond a production capacity of about 2 × 500 tpd of sponge iron.

Thus, the low value of thermal efficiency and the problems inherent in the sponge iron production serves as the motivation of the present research work to propose some better alternative than waste heat recovery boiler to utilize waste heat. Along with the energy aspect the present work also considers better utilization of water.

2. Coal-based sponge iron process

The process of the direct reduction of haematite in a rotary kiln is schematically described in the Fig. 1. Haematite and non-coking coal are fed to the kiln at controlled rates without pre-mixing and the charge moves through the kiln depending upon the rotation speed and inclination of the kiln. In combination with the feed charge, the other successive processes such as drying, preheating and reduction are controlled by means of air which is injected over the kiln length [18]. The material discharged from the kiln is cooled in an evacuated rotary cooler with water sprayed on the shell side. The cooler discharge is then separated into sponge iron, char and ash by magnetic separator. The waste gas generated in the rotary kiln is passed through dust settling chamber (DSC) and the carbon monoxide produced through incomplete oxidation is converted to carbon dioxide by supplying excess air in the after burning chamber (ABC). As the waste gas is at 900 °C, it is passed through a WHRB to tap the heat content, producing steam for power generation. This power generated brings profit to the plants and thus bridges the gap between investment and returns. After WHRB the waste gas is passed through electro static precipitator (ESP) for dust settling and is then released to atmosphere by chimney. The flow of the waste gas throughout the process is generated by an induced draft (ID) fan located near the chimney. The operating data of the process, taken from a typical plant of 100 tpd, is shown in Fig. 1.

3. Design modifications

To propose design modification one should know the possible areas of the process, where energy is being lost and can be conserved. These are as follows: (i) clean waste gas is generated in ESP from where it goes to the chimney at temperature of 220 °C and then directly to the atmosphere and (ii) hot sponge iron is being cooled from 1020 °C to 110 °C using water in the rotary cooler. Water is sprayed on the outer surface of rotary cooler and vapour generated from this process goes directly to atmosphere.

To recover heat from above areas two modifications are proposed. First modification accounts for the preheating of kiln air using waste gas exiting from ESP. However, for second modification initially water is heated using hot sponge iron exiting the

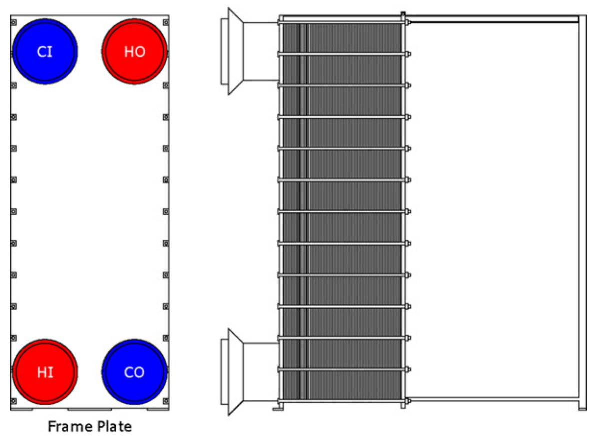

Fig. 2. Model of plate type heat exchanger.

rotary kiln and then hot water is used to preheat the kiln air. These modifications are discussed below in detail.

3.1. Case-1

In this case waste gas available at 220 °C is used to heat kiln air up to 170 °C. In fact, in real situation kiln air is preheated inside the rotary kiln by burning of coal in the presence of air. Thus, preheating air outside the kiln reduces amount of coal consumption. To bring waste gas from ESP outlet to rotary kiln an insulated duct, which is a hollow cylinder of 43 m length and 0.6 m diameter, is used.

To preheat kiln air using waste gas a plate type heat exchanger is designed using a software package HTRI (Xphe, ver. 5.0). There are 40 channels having width of 1.1 m and a spacing of 12 mm. The flow in the heat exchanger is in counter current manner. The packed length is 1.008 m. The schematic model with detailed design parameters of the heat exchanger is shown in Fig. 2.

3.2. Case-2

In this case the hot sponge iron is used to heat water up to 95 °C and further preheating of air is carried out using hot water. The proposed design is shown through following two cases: (a) heat recovery and (b) heat utilization.

3.2.1. Case-2(a) – heat recovery

The kiln discharge at 1020 °C is passed into a rotary cooler in such a fashion as air ingress is completely eliminated, lest it can cause the oxidation of the metallic iron to its oxide. The rotary cooler is an inclined cylindrical drum which rotates at a predetermined speed, moving the charge along its length. The drum is always half filled with the charge. Therefore, a design is proposed to cover the cooler half with a water bath.

The water bath is a semi-cylinder and constructed in such a manner as a gap exists for the rotation of the rotary cooler as shown in Fig. 3. The rotary cooler is rotated whereas water bath is fixed. The material of construction is stainless steel. Water at atmospheric pressure is heated to 95 °C through indirect heat transfer from the hot kiln discharge. The water bath is of 38 m length and having 0.8 m internal diameter, having cut at both the faces for rotary cooler to rotate.

3.2.2. Case-2(b) – heat utilization

The heated water can be used to pre-heat air by indirect heat transfer using hair-pin fin tube double pipe heat exchanger. Hot water is allowed to flow in tube and air is in annulus side. Longitudinal fins on tubes enhance the heat transfer from water to air.

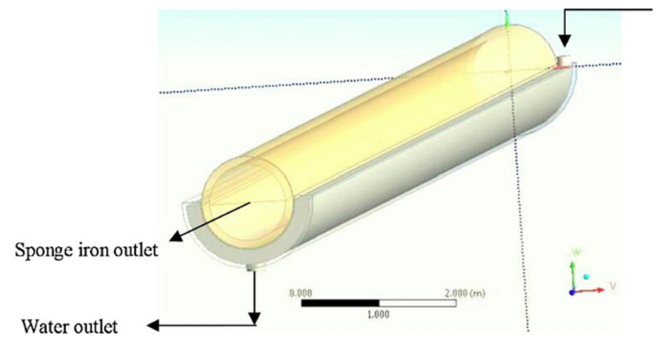

Fig. 3. Model of water bath around the rotary cooler.

The model of the heat exchanger is developed by using a software package HTRI (Xphe E ver.5). The schematic diagram with design parameters are shown in Fig. 4.

4. Model for theoretical amount of coal required in the process

The production of sponge iron is an energy consuming process and the major source of energy for the process is coal. The coal consumption is decided by energy demand of the process which depends on heat gained by the incoming feed to get pre-heated to reaction temperature, heat involved in the reaction of reduction process, heat lost through the rotary kiln wall, and latent heat required by moisture of feed material to evaporate. The estimation of total heat requirement for conventional process consists of the following expressions.

In the preheating zone inlet air and iron ore is heated up to 1020 °C and then reduction takes place in the respective zone. The sensible heat gained by air and ore is supplied by combustion of coal and computed using following equations:

$$Q_a = m_a \times C_a \times (T_r - T_a) \tag{1}$$

$$Q_o = m_o \times C_o \times (T_r - T_a) \tag{2}$$

The heat of reaction of the reduction and combustion reactions are computed using energy balance of the process. It is referred as Q_{rxn}.

The actual heat lost through the wall includes the heat lost through the kiln shell, inlet and outlet hoods, post combustion chamber and inlet area of the cooler. Therefore, the total heat loss is considered as twice as that of the kiln [5,10].

$$Q_{loss} = 2\pi D L h_r \tag{3}$$

The coal is also required to be preheated up to the reaction temperature and sensible heat involved in this is shown as:

$$Q_c = m_c \times C_c \times (T_r - T_a) \tag{4}$$

The average moisture contents in coal and iron ore are 3.58% and 2.3%, respectively, by weight and heat required to evaporate is computed as:

$$Q_w = (0.023 m_o + 0.0358 m_c) \times \lambda \tag{5}$$

In fact, 67.5% of total fixed carbon available in non-coking coal burns completely to give out the heat and remaining 32.5% do not burn. The major fraction of this unburnt carbon is lost to the atmosphere through waste gas as smoke, and rest is discharged with sponge iron from the rotary kiln. The final empirical relation for estimating coal requirement is given through Eq. (6) where NHV is 27,382.267 kJ/kg.

$$Q_a + Q_o + Q_{\text{rxn}} + Q_{\text{loss}} + Q_c + Q_w = m_c \times \text{NHV} \times (0.675) \tag{6}$$

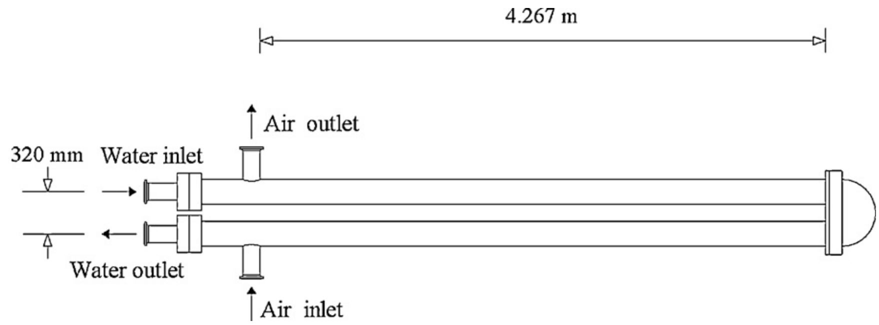

Fig. 4. Model of double pipe heat exchanger.

5. Solution technique for heat integration

To apply heat integration on the sponge iron production process the methodology involved in the computation should be known. The detailed steps of solution methodology are explained in the following steps:

Step 1: Data is collected on (i) composition and specific heat of sponge iron, iron ore, dolomite, flue gas, ESP outlet and wet scraper dust, (ii) proximate and ultimate analysis of feed coal, injection coal and char and (iii) heat of formation and heat of combustion of different components such as Fe_2O_3, FeO, CO, CO_2, Fe, C, O_2, H_2 and H_2O.
Step 2: Overall as well as equipment wise material balance are performed on the process flow diagram to obtain the mass flow rate of each stream. Component and energy balances are performed on each compound involved in the process, considering all the combustion and reduction reactions. Detailed computation of mass and energy balance of process data is shown in Appendix A.
Step 3: Air demand of the process is evaluated based on theoretical oxygen requirement of the process. It is named as $(\text{Air})_{\text{old}}$.
Step 4: The coal consumption of the process is computed using model developed through Eq. (6). Finally, air to coal ratio required for the process is computed.
Step 5: Heat integration is applied by utilizing the design modifications proposed under Section 3.
Step 6: Revised value of coal consumption is computed using design modifications achieved in Step 5. The corresponding air required is calculated using the initial coal to air ratio calculated in Step 4 and named it as $(\text{Air})_{\text{new}}$.
Step 7: If $(\text{Air})_{\text{new}} \neq (\text{Air})_{\text{old}}$, then $(\text{Air})_{\text{old}}$ is replaced with $(\text{Air})_{\text{new}}$ and Steps 4–6 are repeated.
Step 8: With this new value of coal and air consumed in the process mass balance over the process flow diagram is performed to evaluate the reduced amount of flue gas generated in the process.
Step 9: The fixed cost, operating cost, profit and payback period for the modified processes flow diagram are evaluated and compared to select the best design modification.

6. Results and discussion

In the present work two design modifications namely Case-1 and Case-2 are proposed. The detailed analysis of results predicted for these modifications are discussed in the subsequent sections.

6.1. Case-1

The waste gas exiting ESP at 220 °C is carried through an insulated duct to the rotary kiln and is used to preheat kiln air from 30 °C to 170 °C using a plate type heat exchanger.

6.1.1. Design of duct

The insulated duct is used to carry the waste gas from ESP outlet to rotary kiln. As the distance between ESP outlet and rotary kiln is 43 m in the plant insulated duct of same length is designed. The diameter of duct, which is a hollow cylinder, is considered as 1.2 m. This duct is modeled and simulated with Ansys workbench (Ver. 12) as shown in Fig. 5a. The material for insulated duct is steel with insulation of ceramic fiber at the outer surface of the duct. The waste gas inlet temperature is taken as 1173 K (900 °C). It is considered to visualize the maximum drop in temperature which is only possible at maximum heat loss through radiation. As the gas flows through the duct at 14 kg/s, the temperature decreases along the length and it drops to 1167.5 K (894.5 °C) in the insulated duct as shown in Fig. 5b. Similarly, if the waste gas enters the insulated

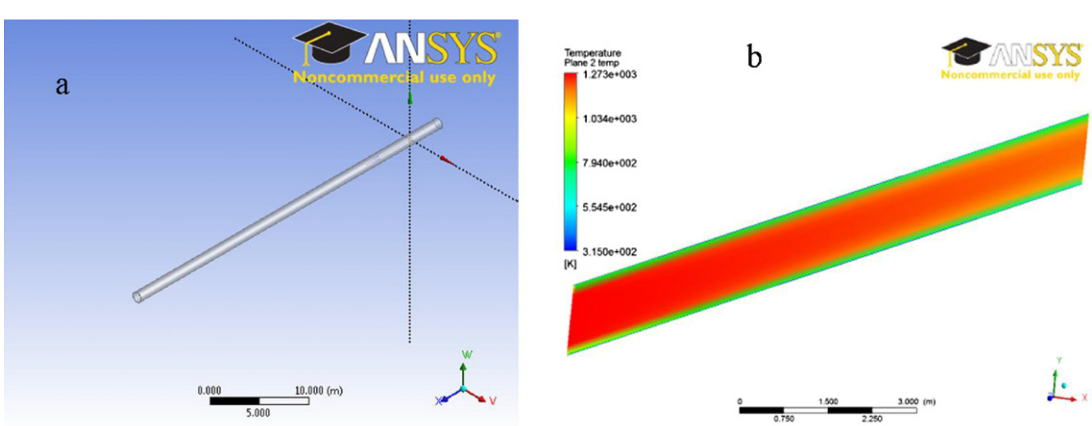

Fig. 5. Model of waste gas carrying duct.

Table 1
Comparative study of insulating materials.

Material	ΔT (K)	Cost (Rs million)
Micro porous silica	3.64	19.1
Ceramic fiber	5.48	2.5
Glass wool	6.62	0.91
Refractory brick	10	2.6

duct at 220 °C it gives 5.5 °C drop in temperature and exits the duct at 214.5 °C.

For the insulated duct as the thickness of insulation increases the temperature drop decreases whereas cost of duct with insulation increases. Thus, the insulation thickness can be optimized based on the temperature drop and the cost of insulation. For this purpose ceramic fiber blanket is chosen as insulating material, which commercially known as kaowool. The analysis shown in Fig. 6 indicates that optimum insulation thickness is 0.2 m. The cost of duct with insulation is Rs 2.53 million as shown in Tables B.2 and B.3.

Further, a comparative study is presented in Table 1 to select the insulating material based on the temperature drop and cost associated with it. For this purpose, four insulating materials such as micro porous silica, ceramic fiber, glass wool and refractory brick are considered. Among these glass wool is found as best insulating material as it costs minimum. However, it can be used only up to 500 °C. For higher operating temperature i.e. 1000 °C or more, ceramic fiber as well as refractory brick are recommended.

During the movement of waste gas through duct with 0.2 m thick insulated duct the temperature drop of 5.5 °C is predicted which causes waste gas to exit duct at 214.5 °C. The heat of waste gas at 214.5 °C is used to preheat air which is a gas–gas heat transfer. It is well-known concept that heat transfer coefficient is lower for gas–gas heat transfer, which may require substantially large area. However, the effect of lower heat transfer coefficient can be somewhat compensated by increasing the temperature difference (ΔT) between the waste gas and air. Thus, large ΔT of a range of 40–50 °C is preferred for operation when gas streams are involved in heat exchange. Considering this fact air can be preheated maximum up to 170 °C where ΔT_{min} is taken as 44.5 °C.

6.1.2. Plate type G–G heat exchanger

Preheating of air is carried out in plate type heat exchanger shown in Fig. 2. The waste gas enters into this exchanger with a velocity of 7 kg/s and is cooled from 214.5 °C to 115.18 °C and consequently, heating the air flowing at 5 kg/s from 30 °C to 170 °C. The overall heat transfer coefficient is estimated to be 52.56 W/m^2 K. Preheating of air up to 170 °C reduces the coal consumption. However, when the consumption of coal drops the quantity of air required to burn it also drops. So the solution of this problem requires trial and error approach. The results of all iterations are shown in Table B.1, which indicates that the final amount of coal required is 5.02 tph, which is 8.7% less in comparison to the existing system. The corresponding air requirement comes out to be 16.504 tph.

The design modification also reduces the amount of waste gas generated by 16.7% which is the additional benefit. The cost analysis and specification of the proposed design is represented in Table 2. If design suggested through Case-1 is implemented the production of sponge iron remains unchanged, hence, reduction in the amount of coal consumption shows a net gain for the process. Thus, due to 8.7% reduction in coal consumption annual profit of Rs 9.6 million is obtained for Case-1 as shown in Table 2. The total annual cost (TAC) for this case is predicted as Rs 107 million which is computed considering straight line depreciation method with the life of each equipment as 10 years. For cost calculation total availability of plant is considered as 8000 h/year. Details of cost computation for Case-1 are shown in Appendix B.

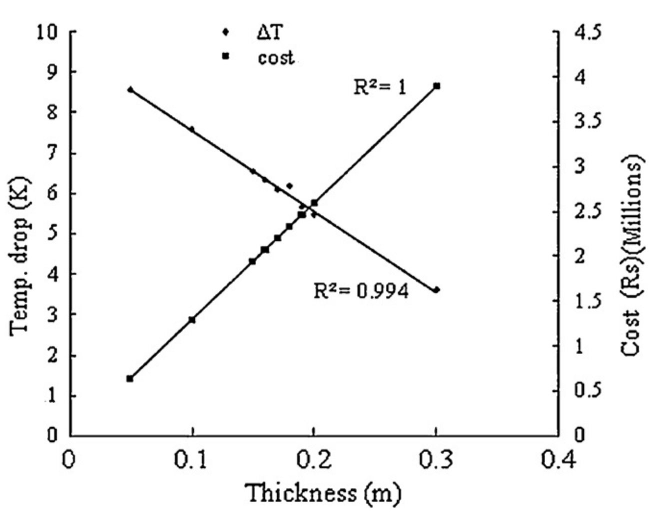

Fig. 6. Optimum thickness of insulation of duct.

6.2. Case-2

In this case the hot sponge iron is used to heat water through a water bath around rotary cooler, shown in Fig. 3, and further preheating of air is carried out using this hot water using a longitudinal fin double pipe heat exchanger, shown in Fig. 4. The design modifications are analyzed further in two cases.

6.2.1. Case-2(a): heat recovery

In the conventional sponge iron production process, water is sprayed from the top on the shell of the inclined rotary cooler to cool hot sponge iron exiting from rotary kiln. The sprayed water is collected in a large trough and sent to the cooling tower for re-use. In this process, the water temperature increases to 5–10 °C and much of the water is evaporated when it comes in contact with the hot rotary shell at around 900 °C. This cooling mechanism consumes 350 tonnes of water per hour. In the proposed design water consumption is reduced to a large extent by allowing cold water to flow through the water bath as shown in Fig. 3. This water bath is covered from the top to minimize the evaporation losses.

The water bath is designed in such manner as it covers half portion of rotary cooler where hot sponge iron is available. The residence time of the water in contact with the rotary cooler is increased in comparison to existing system and due to which water is heated to 95 °C. Applying energy balance on the water and kiln discharge the amount of water thus required drops considerably from 350 tph to 20 tph. Consequently, the capacity as well as capital cost of cooling tower is also reduced. Moreover, amount of water equal to volume of water bath should always be in the bath as hold up which comes out to be 39.187 tonnes.

The discussion on practical operability of the water bath is of great importance. As the water bath is steady and the cooler is in constant rotation on its axis, a gap is left between the walls of the water bath and rotary cooler. This gap leads to constant loss of hot water through leakage and evaporation. Thus this gap can be filled with a lubricated gasket, which can assist the rotation of the cooler and also stop the water from being wasted significantly. The size of this bath is 38 m and 0.8 m in length and internal radius, respectively, which gives the depth of bath as 0.2 m. The length of bath is slightly less than that for rotary cooler which is 40 m. The difference in length of rotary cooler and water bath is allowed for easy rotation of cooler otherwise the larger length of bath may cause problem in

Table 2
Cost analysis of Case-1.

Operating cost			Capital investment			TAC (million/year)	Profit (million/year)
Commodity	Flow rate (t/h)	Cost (million/year)	Item	Specification	Cost, million		
Water	350	5.6	Plate type HX	Ref. Fig. 2	7.42		
Coal	5.02	100.4	Duct	Length: 43 m ID: 1.2 m Insulation: 0.2 m	2.53	107	9.6

Table 3
Cost analysis of Case-2.

Operating cost			Capital investment			TAC (million/year)	Profit (Million/year)
Commodity	Flow rate (t/h)	Cost (million/year)	Item	Specification	Cost (million)		
Water	20	0.208	Double pipe HX	Ref. Fig. 4	1.05		
Coal	5.105	102.1	Water bath	Ref. Fig. 3	0.178	103.5	13.29
			Water line	Length: 60 mID: 1.2 m	0.4		
			Cooling tower	Capacity: 363.8TR	10.67		

Table 4
Comparative analysis of two cases.

Case	Coal cons. (t/h)	Water cons. (t/h)	Operating cost (million/year)	Capital investment		TAC (million/year)	Waste gas generated (t/h)	Profit (million/year)	Payback period (months)
				Item	Cost, million				
Existing	5.5	350	–	–	–	–	25.5	–	–
1	5.02	350	106	Plate type HX	7.42	107	21.234	9.6	12.4
				Ducts	2.53				
				Water bath	0.178				
				Double HX	1.05				
2	5.105	20	102.3	Cooling tower	10.67	103.5	21.601	13.29	11.1
				Water line	0.4				

smooth performance of drives which are employed to rotate the cooler. As this water is to be used in a double pipe heat exchanger, it has to be a de-mineralised water, to avoid scale formation in the tube. Thus, a water treatment plant may also be incorporated with this modification though it is beyond the study of present work.

The present design modification also aimed to recover the heat available with hot water at 95 °C to preheat the kiln air using a finned double pipe heat exchanger. The results obtained for heat recovery aspect is discussed below.

6.2.2. Case-2(b): heat recovery

The hot water from bath is carried through water lines to the double pipe heat exchanger near the rotary kiln. The water bath is placed with rotary cooler of 40 m length and double pipe heat exchanger is to be placed in between the rotary kiln which has a length of 40 m. Thus, the length of water pipe line is considered to be 60 m. The detailed design of double pipe heat exchanger is shown in Fig. 4. Hot water flowing in the inner tube of exchanger at 3.65 kg/s, heats up the air flowing in the annulus side at 4.6625 kg/s from 30 °C to 80 °C. The heat load capacity of heat exchanger is of 194.7 kW. To increase the heat transfer area on the air side longitudinal fins made of brass are designed. Total 56 fins with 1.8795 m^2 area per unit length are placed at the outer surface of inner tube in double pipe heat exchanger. The overall heat transfer coefficient is estimated to be 76.64 W/m^2 K.

The preheating of air up to 80 °C causes reduction in coal consumption. The reduced amount of coal requires less amount of kiln air. Based on the trial and error method final amount of coal required is estimated as 5.1 tph and the corresponding air requirement is found as 16.785 tph. However, this modification reduces water consumption up to 20 tph which is 96.3% in comparison to existing system. Consequently, capacity of cooling tower also reduces significantly. The capacity of new cooling tower is computed as 363.8 TR which is 37.2% less in comparison to the exiting cooling tower.

The cost analysis and specification of the proposed design is represented in Table 3. The difference in coal and water consumption in comparison to existing system shows the profit of the modification suggested in Case-2. Thus, 4.8% and 96.3% reductions in coal and water, respectively, give the annual savings of Rs 13.29 million as shown in Table 3. The cost of cooling tower as well as double pipe heat exchanger is found as Rs 10.67 million and Rs 1.05 million [19]. Costs of water bath and water line are computed using weight of material and material of construction. Total capital cost and TAC for this case are predicted as Rs 12.298 million and Rs 103.5 million, respectively. These costs for Case-2 are predicted in the manner similar to that is shown in Appendix B.

Table 5
Values of raw materials and production capacity of the plant.

Plant production capacity (tpd)	Raw material consumption (t/h)		Ratio of production to intake capacity	Reference
	Coal	Iron ore		
100	5.5	6	8.69	Present work
446.4	21.437	30	8.68	[20]

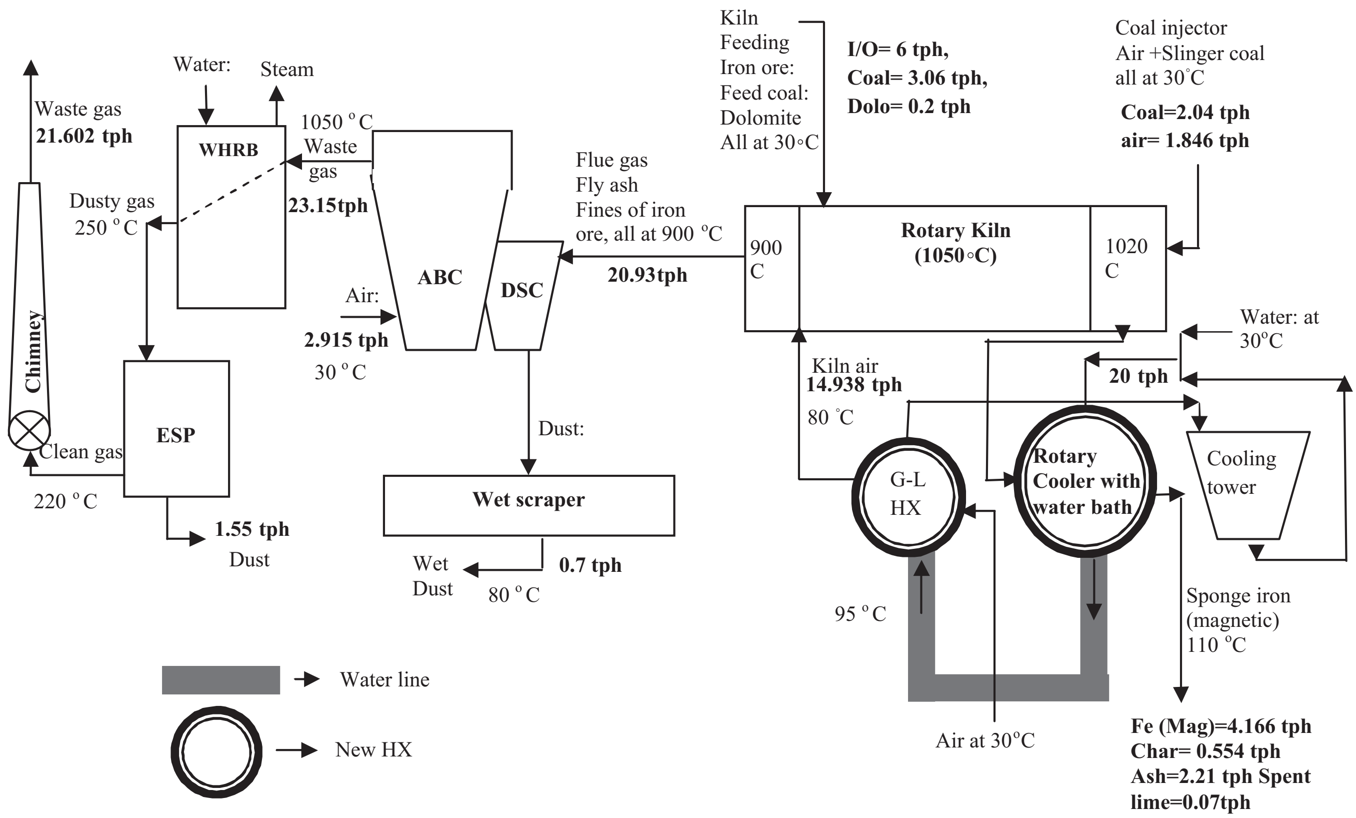

Fig. 7. Process flow diagram with design modifications.

6.3. Comparison of Case-1 and Case-2

The comparison of Case-1 and Case-2 with the existing system based on coal consumption, water requirement, operating cost, capital cost, profit, payback period and waste gas generation is registered in Table 4. It shows that though Case-1 consumes less coal and produces minimum waste gas in comparison to existing system as well as Case-2, its TAC is 3.4% more than that for Case-2. Due to 27.8% less profit for Case-1 in comparison to Case-2, Case-1 offers higher payback period than that of Case-2. Thus, based on payback period and TAC Case-2 is selected as best proposed design. The indirect benefit of this modification is that the load of cooling tower is reduced to a large extent. The modified design for Case-2 is shown in Fig. 7.

For Case-2 intake capacity of rotary kiln is reduced by 3.4% as amount of coal is reduced by 4.8%. As reduction in intake capacity is not significant the same size of rotary kiln can be used in the modified design. Similarly, sizes of post equipment to kiln are also same. However, size of cooling tower is changed as its capacity is reduced by 37.2% in comparison to existing system.

6.4. Generalization of the results

To show the effectiveness of the present modification example of a 100 tpd capacity Indian sponge iron plant is selected and thus, the results shown in this work is applicable to this plant. However, the heat integration methodologies proposed through Case-1 and Case-2 are applicable to other plants also which have different capacity. For all sponge iron plants similar trends as well as range of operating parameters such as temperature and pressure are observed. However, flowrate of feed may change. The values of flowrate of feed and capacity of two plants are reported in Table 5. These plants are placed in different part of the country. Table 5 shows that ratios of production to intake capacity of two plants are almost equal which can also be maintained for other sponge iron plants. Thus, maintaining this ratio a rough estimation of coal consumption for other plant can be carried out using heat integration technique presented though Case-2. However, the design parameters of the new equipment may change. Similar study can be applicable to other mattelurgical industries such as steel plants.

7. Conclusions

Based on the results obtained from development of design modifications for heat and water integration in sponge iron plant and also comparison of these modifications with the existing process on the basis of capital investment, coal consumption, water requirement, profit and payback period, several conclusions can be drawn as listed below:

(1) For modification suggested through Case-1 preheating of air up to 170 °C is carried out in plate type heat exchanger of 709 kW heat load having an effective heat transfer area of 213.2 m^2. It reduces the coal consumption by 8.7% less in comparison to the existing system. It thus also reduces the amount of waste gas generated by 16.7%. Due to reduction of coal consumption by 8.7% a profit of Rs 9.6 million can be earned per annum which gives payback period for this design modification is 12.4 months.
(2) For Case-2 a water bath is proposed to utilize the heat content of kiln discharge, which gets cooled from 1020 °C to 110 °C, to heat up the water to 95 °C. Air is preheated by using a double pipe longitudinal fin heat exchanger having heat duty of 194.7 kW. Hot water flowing in the inner tube of exchanger heats up the air of annulus side from 30 °C to 80 °C. Total 56 fins with 1.8795 m^2/m area per unit length are placed at the outer surface of inner tube in double pipe heat exchanger. Preheating of air to 80 °C before entering the kiln reduces coal consumption by 4.8%.
(3) Due to incorporation of water bath in Case-2, heavy evaporation losses can be minimized and this leads to a 96.3% reduction in the water consumed for cooling the kiln discharge in existing system. Consequently, capacity of cooling tower is reduced by 37.2%. The payback period of this design modification is 11.1 months.
(4) Based on payback period and TAC Case-2 is selected as best proposed design.
(5) The applicability of the conclusions can be generailized as: the main conclusion of the study is conclusion shown through point 4 where Case-2 is selected as best design among two cases. Case-2 will remain optimum design for any other sponge iron plant as operating temperature and equipment of the process are similar for other plants of different capacities. Due to varying production capacities intake capacity may be changed. Thus, for other plants saving in terms of coal and water flow rates for Case-1 and Case-2 may change; however, order of cases based on payback period will remain same i.e. Case-2 < Case-1.

Further, for Case-1 air can be preheated up to 170 °C as waste gas is available at around 220 °C and 50° difference is maintained for better heat transfer. These temperatures are almost same in other plants also. However, the heat load may change based on amount of air. The amount of air depends on coal consumption which is related to the capacity of plant. Thus, savings also vary with capacities of plants which can also be computed by maintaining the ratio reported in Table 5.

Nomenclature

A	area (m^2)
C	specific heat capacity (J/kg K)
M	mass flow rate of iron ore (kg/h)
T	temperature (K) or (°C)
Q	heat load (kW)
D	diameter (m)
G	gas
H	individual heat transfer coefficient (W/m^2 °C)
L	length (m)
tpd	tonnes per day
TPA	tonnes per annum

Greek letters

Δ	difference between two parameters
λ	latent heat of vaporization (kJ/kg)

Subscripts

W	moisture
A	air/ambient
O	iron ore
C	coal
R	reference/rediation
Rxn	reaction
Loss	heat loss

Overall mass and component mass balance around rotary kiln are shown in Tables A.1 and A.2.

See Table A.3.

Appendix A.

Table A.1
Overall mass balance on process flow diagram.

Input stream	m^3/h	tph
Iron ore		6
Feed coal		3.3
Kiln air	15,417	17.976
Slinger coal		2.2
Air with injection coal	2000	2.332
Dolomite		0.2
Air in ABC	2500	2.915
Total		34.923
Output stream	**m^3/h**	**tph**
Sponge iron		4.166
Char		0.555
Ash		2.222
Spent lime		0.07
Chimney gas		25.518
ESP dust		1.55
Wet scraper		0.7
Back spill		0.14
Total		34.923

Appendix B.

The detailed calculation for Case-1 is shown below:

Step 1

Considering kiln inlet temperature of air to be 170 °C coming from G–G heat exchanger in Eq. (6) instead the ambient temperature of air the revised value of coal is computed. Further, the corresponding value of air is calculated maintaining the initial air to coal ratio of 3.28. New value of air is substituted again in Eq. (6) and corresponding new value of coal is computed. This procedure is iterated until the air requirement converges. Table B.1 shows the iteration results of the computation.

Step 2

Cost computation for Case-1 is presented below:

Operating cost
Total working hours = 8000 h/year
Cost of water = Rs 2/tonne
Flow rate of water = 350 tonnes/h
Total operating cost of water = 350 × 2 × 8000 = 5.6 million/year
Cost of coal = Rs 2500/tonne
Flow rate of coal = 5.02 tonnes/h
Total operating cost for coal = 2500 × 5.02 × 8000 = 100.4 million/year
Total operating cost = 106 million/year

Table A.2
Component balance across rotary kiln.

Material	kg/h	%/100	Input (kg/h)			Output (kg/h)
Iron						
Feed rate	6000			Fe(m) in dri		3458.3333
Fe(t)		0.68116	4086.96	Sponge iron flow rate	4166.6667	
				Fe(M) content (%/100)	0.83	
				Fe lost as feO		226.85185
				Fe(T) (%/100)	0.9	
				Fe(m) (%/100)	0.83	
				Fe lost in flyash		401.45
				Fe_2O_3 (%/100)	0.37	
				Fly ash produced	1550	
				Other losses		0.3248148
Total			4086.96	Total		4086.96
Gangue						
Ore		0.0607	364.2	In sponge iron		320.83333
				Gangue (%/100)	0.077	
				Lost in flue dust		43.366667
Total			364.2	Total		364.2
Carbon						
Total c in Fe coal			2136.16	FC out in char		158.33333
Fe coal flow	3160			Char flow	555.55556	
C		0.676		FC (%/100)	0.285	
				FC out in flue dust		208.94
Total c in sl coal			1485	FC out in sponge		10.416667
SL coal flow	2200					
C		0.675		FC consumed in reaction		1136.0171
				F.C burnt to CO_2		728.5129
Total C in dolomite			26.06			
Dolomite flow	200			VMC burnt to CO and CO_2		354.21411
C		0.1303		Unburnt VMC		1027.4062
*F.C in FE coal	1294.02					
C in VM of FE coal	842.14					
*F.C in SL coal	948.2			C in CO_2 from dolomite		23.379668
C in V.M of SL coal	536.8					
Total			3647.22		Total	3647.22
Coal ash						
Ash in F.E. coal			750.184	Ash in char		386.66667
Ash in SL. Coal			468.38	Ash in flyash		558
				Free ash to cooler		273.89733
Total			1218.564	Total		1218.564

Table A.3
Energy balance across rotary kiln.

	Reduction reactions	Reactants (kmol/h)			Products (kmol/h)			Heat (Gcal/h)
		Fe_2O_3	FeO	CO	Fe	FeO	CO_2	
1	$Fe_2O_3 + CO \rightarrow 2FeO + CO_2$	32.9	–	32.9		65.8	32.9	−0.1006
2	$FeO + CO \rightarrow \ + CO_2$		61.7	61.7	61.7		61.7	−0.1829

	Combustion reactions	Reactants (kmol/h)					Products (kmol/h)			Heat (Gcal/h)
		C	CO	CO_2	H_2	O_2	CO	CO_2	H_2O	
3	$C + CO_2 \rightarrow 2CO$	94.6	–	94.6	–	–	189.3	–	–	3.7690
4	$2CO + O_2 \rightarrow 2CO_2$	–	94.6	–	–	47.3	–	94.6	–	−6.3564
5	$C + O_2 \rightarrow CO_2$	60.7	–	–	–	60.7	–	60.7	–	−5.7356
6	Slinger V.M.C burnt to CO_2: $C + O_2 \rightarrow CO_2$	18.5	–	–	–	18.5	–	18.5	–	−1.7481
7	Slinger V.M.C burnt to CO: $2C + O_2 \rightarrow 2CO$	11.0	–	–	–	5.5	11.0	–	–	−0.3010
8	Slinger V.M. H2 burnt to H_2O: $2H_2 + O_2 \rightarrow 2H_2O$	–	–	–	29.3	14.6	–	–	29.3	−1.7512
9	Radiation loss from the kiln wall = 3.14*d*l*k									3.1982
10	Vaporization of moisture = moisture in coal*latent heat									0.1023
11	Sensible heat lost in:									
(a)	Flue gas									9.0107
(b)	Sponge iron									0.6882
(c)	Char									0.2923
(d)	Ash									0.7316
(e)	Spent lime									0.0177
(f)	V.M									0.3684
	Total heat evolved									−16.1761
	Total heat absorbed									18.1788
	Other losses									−2.0027

Table B.1
Iteration results for coal and air requirement after heat integration.

Iteration no.	Coal consumption (kg/h)	Air requirement from air/coal ratio = 3.28
1	5034.884	16,551.50957
2	5022.875	16,512.0321
3	5020.846	16,505.3607
4	5020.503	16,504.2332
5	5020.445	16,504.0412
6	5020.435	16,504.0111
7	5020.433	16,504.0110
8	5020.433	16,504.0110

Table B.2
Calculation for steel duct costing.

Parameter	Value
IR (m)	0.6
OR (m)	0.608
Density steel (kg/m^3)	7850
Volume (m^3)	1.304833
Length (m)	43
Mass (kg)	10,242.94
Cost of steel (Rs/tonne)	27,970
Cost of duct (Rs)	286,495.1

Table B.3
Calculation for insulation costing.

Parameter	Value
OR (m)	0.6
IR (m)	0.4
Length (m)	43
Volume of insulation (m^3)	27.004
Cost (Rs/m^3)	83,262
Total cost of insulation (million) = cost × volume	2.25
Insulated duct cost (million) = insulation cost + cost of steel	2.53

Capital investment

i. Capital cost of G–G plate type heat exchanger is computed from following equation [Shenoy]:

$$\mathrm{Cost}(\$) = 30,000 + 1900 \times (A)^{0.78}$$

where area of exchanger is taken from Fig. 2

$$\text{So, } \mathrm{Cost(Rs)} = (30,000 + 1900 \times (213.224)^{0.78}) \times 48$$
$$= 7.42\,\text{million}$$

ii. Cost of steel duct is shown in Table B.2 which is obtained by multiplying cost and mass of steel.

iii. Cost for insulation is represented in Table B.3.

Total annual cost (TAC)

TAC is the summation of annual operating cost and annual capital cost. Annual operating cost for Case-1 is Rs 106 million/year. Annual capital cost is estimated assuming life of each equipment as 10 years and straight line depreciation.

So, annual capital cost = (2.53 + 7.42)/10 = 0.995 million/year

TAC = 106 + 0.995 = 106.995 = 107 million/year

Profit

$$\begin{aligned}\text{Saving of coal} &= \text{coal used in existing plant– coal used in Case-1}\\ &= 5.5 - 5.02 = 0.48\,\text{tonnes/h} = 3840\,\text{tonnes/year}\end{aligned}$$

$$\begin{aligned}\text{So, profit} &= 3840 \times 2500 = 9.6\,\text{million/year}\\ \text{And payback period} &= \text{total captial investment/yearly profit}\\ &= (2.53 + 7.42)/9.6 = 12.4\,\text{months}\end{aligned}$$

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