ENERGY CONSERVATION IN COAL BASED SPONGE IRON CLUSTER USING TOTAL SITE INTEGRATION

Ph.D. THESIS

by

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JULY, 2018

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A THESIS

Submitted in partial fulfilment of the requirements for the award of the degree

of

DOCTOR OF PHILOSOPHY

in

CHEMICAL ENGINEERING

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ROORKEE

CANDIDATE'S DECLARATION

I hereby certify that the work which is being presented in the thesis entitled **"ENERGY CONSERVATION IN COAL BASED SPONGE IRON CLUSTER USING TOTAL SITE INTEGRATION"** in partial fulfilment of the requirements for the award of the Degree of Doctor of Philosophy and submitted in the Department of Chemical Engineering of the Indian Institute of Technology Roorkee, Roorkee is an authentic record of my own work carried out during a period from July, 2015 to July, 2018 under the supervision of Dr. Shabina Khanam, Associate Professor, Department of Chemical Engineering, Indian Institute of Technology Roorkee, Roorkee.

The matter presented in this thesis has not been submitted by me for the award of any other degree of this or any other institution.

(BANDARU VENKATA RAMANAIAH)

This is to certify that the above statement made by the candidate is correct to the best of my knowledge.

(SHABINA KHANAM)
Supervisor

Date:

The Ph.D. Viva-Voce Examination of Mr. BANDARU VENKATA RAMANAIAH, Research Scholar, has been held on.. .

Chairman, SRC External Examiner

This is to certify that the student has made all the corrections in the thesis.

Supervisor Head of the Dept./Chairman, ODC

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TO

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ABSTRACT

An enormous potential for saving energy is available in coal based sponge iron plants as these are associated with high percentage of energy wastage per unit sponge iron production. To deal with the current issue of energy crisis, various design and operational modifications are proposed to sponge iron plants. For this purpose, actual data of three sponge iron industries such as Process-1, Process-2 and Process-3 being operated with different capacities in an industrial cluster of India are considered. All these three processes are SL/RN based (Stelco-Lurgi / Republic Steel-National Lead). They consist of rotary kiln, rotary cooler, dust settling chamber (DSC), after burning chamber (ABC), evaporating cooler, electrostatic precipitator (ESP), wet scrapper and chimney as important equipment.

In each process potential areas are identified where heat is lost and can be recovered and where this heat can be utilized. Considering these potential areas, two different design and operational modifications, Scheme-1 and Scheme-2, are proposed based on principles of process integration and applied in all three processes of coal based sponge iron cluster, individually. Further, two strategies, Strategy-1 and Strategy-2, are suggested for total site integration (TSI), considering all three processes as a single process to conserve energy in total site of plants of similar type where conventional methods are not applicable. These schemes and strategies vary in terms of modification, savings and investments.

Scheme-1 accounts for preheating kiln inlet streams using waste gas exiting ESP and cooling kiln outlet using same ESP exit gas. However, for Scheme-2, initially kiln outlet is cooled using kiln air and further kiln inlet streams are preheated using waste gas exiting ABC. In Strategy-1 waste gas that exits from ABC of Process-1 is used for preheating kiln feed and slinger coal of all three processes, individually. Kiln outlet stream is cooled using kiln air of respective process. Further, kiln air is preheated using waste gas to maximum possible temperature such that ΔT_{min} equal to $50^{O}C$ is maintained. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation. On the other hand, for Strategy-2 waste gas streams exiting ESPs of Process-1, Process-2 and Process-3 are mixed and used to preheat kiln feed and slinger coal of all three processes at Process-1. Further, kiln outlet stream is cooled using kiln air

of respective process. Along with this, Strategy-2 proposes waste heat recovery boiler, for power generation, in Process-1 and Process-2 using waste gas exiting from ABC of respective processes.

In the present work, the modified approach is developed to propose and solve energy conservation schemes as well as strategies. In this approach following steps are proposed: Step-1: Define a strategy for energy conservation, Step-2: Data extraction for all processes together in total site according to the strategy, Step-3: Preparation of stream table considering all processes together and selection of ΔT_{min}, Step-4: Utility targeting of total site, Step-5: Designing HEN of total site and Step-6: Modification of total site PFD. Guidelines to perform all these steps are also proposed. This method considers only six steps for TSI, which are far less as compared to conventional method where ten steps are involved. Pinch analysis is used to compute hot and cold utility requirements for schemes and strategies. In sponge iron processes, considered in the present work, utility and process streams are same i.e. coal and air where coal is burnt in the presence of air to provide necessary heat in the kiln. Therefore, once amount of hot utility changes coal and air requirements also vary. Due to this fact, flow rates of process streams (mainly coal and air) vary during solution of the problem. This requires trial and error computation technique to solve the problem. Thus, the present approach includes iterative method, if utility and process streams are same, as well as non-iterative method, if these are different. Further, as coal is used as utility and process stream in sponge iron processes, a revised model to compute its consumption is also proposed based on hot utility, heat of reactions, kiln feed and air preheating, radiation losses, dolomite decomposition, heat required to vaporize the coal volatiles, etc. For each process of total site, at the existing conditions, coal utility factor, a fractional use of total energy released from coal combustion inside the rotary kiln, is computed from modified model developed for coal combustion. Further, this utility factor is used for computation of coal consumption in all design modifications proposed. Apart from this, constant properties at average temperature in each iteration are considered in the present work.

For each scheme and strategy economic as well as operability analyses are presented, which include capital investment, coal consumption, water requirement, energy consumption, savings, net profit value (NPV), internal rate of return (IRR) and discounted payback period (DPP). Further, based on these factors along with waste gas generation and %heat recovery, selections of best scheme for heat integration and strategy for TSI are carried out. Capital investment is done to install new gas-gas and gas-solid heat exchangers, gas carrying ducts with or without insulation, conveyor belts to carry hot solid streams and forced draft fans.

Results of Scheme-1 and Scheme-2, applied for individual processes, indicate that Scheme-2 offers more coal and water savings over Scheme-1 in all three processes. Consequently reduction in waste gas generation is found more in Scheme-2 than Scheme-1. Energy ratio (ER), defined as ratio between actual energy consumed in the kiln to theoretical energy required for the reduction reactions to continue, is found less in Scheme-2 than Scheme-1. Thus, actual energy consumption, in Scheme-2, is more close to theoretical energy required than in Scheme-1. Based on these factors Scheme-2 is selected as best heat integration option for all three processes. On the other hand, results of Strategy-1 and Strategy-2, proposed for TSI of sponge iron cluster, indicate that both strategies offer better results over the existing system. However, savings in coal and water and hence, reduction in waste gas generation offered by Strategy-1 are found significantly higher as compared to Strategy-2. Excess energy consumption, which is 76%, prior to TSI, is reduced to 8% and 37% through Strategy-1 and Strategy-2, respectively. After final design Strategy-1 recovers 99.8% of waste heat available in the modified site whereas, Strategy-2 recovers 86%. Total annual cost (TAC) is 11.8% less for Strategy-1 as compared to Strategy-2. The discounted payback period for Strategy-1 is 11.65 months and that of Strategy-2 is 10.46 months. Thus, Strategy-1 is considered better heat recovery option over Strategy-2. Further, the results obtained for Strategy-1, are found significantly greater than Scheme-2 and various design modifications reported, based on single site integration, in literature. Hence, Strategy-1 is selected as the best design for cluster of any sponge iron plants and integrating as a total site is more profitable than integrating individually.

ACKNOWLEDGEMENT

It gives me immense pleasure to place on record my deep sense of gratitude and indebtedness to my supervisor, Dr. Shabina Khanam, Associate Professor, Chemical Engineering Department, IIT Roorkee, India, for giving me an opportunity to work under her esteemed guidance. Her superb guidance with enriched knowledge, regular encouragement and invaluable suggestions at every stage of the present work has proved to be extremely beneficial to me. Her painstaking efforts and immense care in going through the manuscript, her timely advice with great patience and cool mind proved a great help during the completion of present work. She is very noble person with soft and tender heart and always ready to extend to her students at any time. I consider myself so fortunate to have such a pleasant and able supervisor like her in my life. I owe a lot to Late M/s. Tirumalaraju Deepthi because of whom I came in contact with my supervisor.

I would like to express my deepest thanks to Prof. Bikash Mohanty, Professor, Chemical Engineering Department, IIT Roorkee, India, who provided his whole hearted cooperation, never ending inspiration and guidance, all blended with personal touch throughout the duration of this work. His invaluable suggestions and discussions have immensely contributed towards the completion of present work.

I would like to express my sincere gratitude to Prof. Shishir Sinha, Head, Department of Chemical Engineering, and Prof. V. K. Agarwal, and Prof. C.B. Majumdar former Heads, Department of Chemical Engineering for their kind help in departmental affairs.

I would like to express sincere gratitude to my parent institute, Maharaj Vijayaram Gajapathi Raj College of Engineering, Vizianagaram, Andhrapradesh for sponsoring me to pursue PhD under QIP. I owe a lot to the Chairman and administrative team of my parent institute. I especially thankful to Dr K.V.L. Raju, Principal of my parent institute for his whole hearted support. I would like to record very humble and sincere gratitude to All India Council for Technical Education (AICTE), New Delhi, India, for sponsoring me to pursue PhD under QIP programme. I sincerely thank QIP centre, IIT Roorkee, India, for supporting in smooth completion of my research work.

I am grateful to reviewers of my published research articles for their valuable and constructive comments.

I am thankful to my friends Mr. V. Ravindrababu, Mr. K. Malyadri, Mr. N. Srinivasarao, Mrs. Seshulata, Mr. Yenuganti Srinivas, Mr. Vuka Srinu, Mr. Sarat Chandra, Mr. Balaji, Mr. Lakhminarayana, Mr. K. Srinivasarao, Mr. Ch. Krishna and his parents for their direct and indirect support. I am also thankful to Mr. Mahesh garu and Prof. Dr. Hemalatha Thiagarajan, Professor of Mathematics, NIT Trichy, India, for their support.

I am ever grateful to my seniors Dr. Vrsakapi, Dr. Navayauvana, Dr. Sanatkumar, Dr. Harilila and Dr. B. Srinivas for their continuous support in my life, particularly in difficult times. I am especially grateful to my brother-in-law and senior Dr. P. Surya Prasad and his parents for their valuable support in my life.

I am also thankful to Ms. Anchal Tyagi, Mr. B. Uday Bhaskar, Dr. B. Mouli, Mr. MSV Prasad, Mr. B. Suryawanshi, Mr. Gaurav Kumar Silori, Mrs. Vibha Devi, Ms. Preeti Kumari, Mr. Ajay Kumar Meena, Mr. Kamal Kanth Meghwal, Mr. Ajay Sharma, Mr. Brajesh and Dr. G.V.S.K. Reddy for creating good research environment and giving me company during my Ph.D.

I would be failing in my duty if I don't mention my father, Mr. Venkata Narasaiah whose strictness helped me a lot in my schooling. I also wish to acknowledge the love, support and dedication of my mother Mrs. Nagamma towards her family and children. I am grateful to my loving sisters: Mrs. Narasamma, Mrs. Rattamma and Mrs. Sujatha, brothers-in-law: Mr Subbarao, Mr. Venkateswarlu and Mr. Suryaprasad and other family members and relatives for their continuous love and moral support. I am grateful to Mr. Y. Tirupataiah, my cousin, who lovingly guided me in selecting appropriate stream in my education after my SSC.

I am very grateful to my loving teacher Prof. P.V. Krishnan, retired professor, IIT Delhi, India, whose teachings gave me new life. I owe myself to him completely for all his sacrifices, love and care towards me.

I am always grateful to my most merciful and compassionate Lord Krishna, who is my life and soul.

(BANDARU VENKATA RAMANAIAH)

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NOMENCLATURE

A	Heat transfer area (m^2)
C_p	Specific heat capacity, kJ/kg ^OC
CP	Heat capacity flow rate, kW/^OC
D	Duct/diameter
h	Hour
H	Enthalpy (kW)/Heater
HX	Heat Exchanger
m	Mass flow rate, kg/h
P	Pressure, atm
Q	Heat load, kW
T	Temperature
tpa	Tonne per annum
tpd	Tonne per day
U	Overall heat transfer coefficient (W/m^2-^OC)/Utility factor

SUBSCRIPTS

a	Air
c	Coal
cwsrc	Covered wall-to-solid by radiation and convection
d	Dolomite
de	Decomposition
ewsr	Exposed wall-to-solid by radiation
f	Feed/Factor
gs	Gas-to-solid
gw	Gas-to-wall
hu	Hot utility
L	Losses
loss	Rotary kiln losses

m	Moisture
min	Minimum
ore	Iron ore
p	Process
r	Radiation
s	Supply
sib	Surface-to-interior of the bed
t	Target
v-char	Volatiles in char
v-coal	Volatiles in coal

GREEK LETTERS

Δ	Difference between two parameters
λ	Latent heat of vaporization, kJ/kg

ABBREVATIONS

ABC	After Burning Chamber
CC	Composite Curves
CH	Chimney
DPP	Discounted payback period
DRI	Direct Reduced Iron
DSC	Dust Settling Chamber
EAF	Electric Arc Furnace
EC	Evaporating Cooler
ESP	Electrostatic Precipitator
GCC	Grand Composite Curve
HEN	Heat Exchanger Network
HI	Heat Integration
IRR	Internal rate of return
NHV	Net heating value

NPV	Net profit value
PFD	Process Flow Diagram
PTA	Problem Table Algorithm
RC	Rotary Cooler
RK	Rotary Kiln
SI	Sponge Iron
SSiP	Site Sink Profile
SSoP	Site Source Profile
SSSP	Site Sink Source Profiles
SUGCC	Site Utility Grand Composite Curve
TAC	Total Annual Cost
TS	Total Site
TSHI	Total Site Heat Integration
TSI	Total Site Integration
TS-PTA	Total Site Problem Table Algorithm
TSST	Total Site Sensitivity Table
UGCC	Utility Grand Composite Curve
WHRB	Waste Heat Recovery Boiler
WS	Wet Scraper

CHAPTER 1

INTRODUCTION

Sponge iron, also known as direct reduced iron (DRI), is produced by direct reduction of iron ore in the presence of coal and air. Sponge iron is produced primarily using non-coking coal and natural gas as reductants. It is used widely in steel making process and resembles to a honey comb in its structure. With the availability of raw materials, high demand of sponge iron and less payback period, sponge iron industry has emerged as a profitable venture. As India has adequate coal deposits and its utilization for steel plants is considered to be of prime importance, production of coal based sponge iron is considered as a viable option [2]. In DRI production India deserves special attention, not only because the country is the largest producer of DRI while contributing 25.6% to world DRI production, but also because production is primarily coal based. Out of the global DRI production of 87.1 MT in 2017, the production of coal based DRI occupies 15.33 MT i.e. 17.6% of the total. Further, 96.5% of the total coal based DRI is produced in India only [111]. India has 374 plants with an installed production capacity of 48.63 MT/year. Amongst these 369 plants are coal based producing 75% of the total production [178].

Although most of these plants have acquired the desired level of operational efficiency, energy utilization of various units is found below to the optimum limit due to lack of proper integration techniques, non-optimal process parameters, high energy consumption and old running process technology. Hence, a large scope exists to make sponge iron industry more competitive by cutting down its internal losses using modern technology [131]. Amongst these drawbacks the problem of proper integration technique and high energy consumption are addressed by many investigators [1-3,11,17,27,52,114,157].

In coal based sponge iron processes, coal is the only source of heat and the required heat is generated when coal is burnt along with air in the rotary kiln. This heat is utilized for removing the moisture from feed materials and preheating these, metallization of iron ore, compensating radiation losses from kiln wall and decomposition of dolomite, etc. At the same time much amount of heat energy that is present in kiln outlet and waste gas is unutilized in the plant. Thus, in sponge iron process kiln feed, kiln air and slinger coal are heat sinks whereas kiln outlet and waste gas released are heat sources. Energy integration can be done across these energy sinks and sources so that much amount of coal (energy) consumption can be reduced.

The literature shows that many investigators examined the sponge iron manufacturing process and found that during operation a large amount of heat is generated [1-3,11,49,77,114,157]. They mainly proposed waste heat recovery boiler (WHRB) to generate steam to recover energy of waste gas though did not show computational analysis for this. A few authors suggested improvements in energy efficiency of the process by modifying rotary kiln design [52]. Agarwal and Sood [1] proposed to utilize the untapped energy of waste gas by installing WHRB to generate steam.

Apart from WHRB, a few investigators proposed different energy conservation schemes to utilize heat of waste gas using process integration principles [12,20,24,36,77,83,86,96,130,131,161]. Prasad et al. [130] proposed design modifications for sponge iron industry using principles of process integration, which eliminated the use of evaporating chamber resulting in 72.6% and 30.5% reduction in water and coal consumption, respectively. Kumar and Khanam [83] designed a water bath around rotary cooler to recover heat from kiln outlet and a gas-liquid finned double pipe heat exchanger for heat integration. Through these modifications they reduced coal and water consumption by 7.2% and 94.3%, respectively. Recently, profitability analysis of power generation and preheating using waste heat of sponge iron process was carried out and found that power generation was more profitable over preheating when net profits were considered [59]. The energy integration of the process can be studied using pinch technology as well as mathematical modelling [56]. The classical Pinch Design Method is primarily applied in various industries for designing of Heat Exchange Networks [60]. Along with it many investigators contributed significantly in the area of process integration principles [14,34,35,37,54,55,85,86,144]. Various authors also studied on heat integration of intermittent process streams and multipurpose batch plants [87,101,102,148]. Mathematical modelling is a promising tool to analyse and optimize operating parameters of chemical processes [4,47,84,115,118,153].

However, these studies considered only single plant for energy integration whereas, in many states of India there are hubs of coal based sponge iron plants, which are placed very near to each other in the mineral belts of India [146]. To integrate energy in such clusters of plants simultaneously a concept of total site integration can be used effectively.

Dhole and Linnhoff [50] proposed the concept of total site integration (TSI) where they developed a graphical method of site profiles. Klemeš et al. [80] extended the methodology of Dhole and Linnhoff [50] by introducing total site profiles and site utility grand composite curves. Goršek et al. [63] presented modified site sink source profiles to determine energy saving potential. Bandyopadhyay et al. [13] proposed a simplified and novel methodology for targeting cogeneration potential based on the Salisbury approximation. Varbanov et al. [165] extended the TS methodology considering individual ΔT_{min} for each process and utilities whereas, Fodor et al. [53] accounted stream specific ΔT_{min} inside each process while setting different ΔT contribution (ΔT_{cont}) and also using different ΔT_{cont} between process and utility systems. Tarighaleslami et al. [162] focused on unifying total site heat integration (TSHI) for multiple isothermal and non-isothermal utility targeting. Hackl et al. [66] applied the concept of TS to a cluster of five chemical companies and showed that 50% of savings could be achieved by moderate changes to the existing heat exchanger system whereas, 92% of changes were technically feasible. Similarly, many researchers applied TSHI techniques in various plants, e.g., heavy chemical complexes [106], steel plants [107], petro-chemical plant [89], large dairy factories [8] and bromine plant [19].

It appears from above discussion that for energy conservation industrial cluster/site considered includes plants of different process such as heavy chemical complexes [106]. However, in a number of cases, such clusters are of same process such as sponge iron, sugar, etc. As all these plants are similar, there is no relative advantage of conventional TSI, which is more suitable to integrate heat within the site of different processes.

Further, it is observed that global warming is the serious environmental issue causing enormous pressure on all countries to reduce greenhouse emissions. World's primary energy source comes from fossil fuels (coal, oil, natural gas) and its combustion gives the major source of global CO_2 and other greenhouse gas emissions. Largest source of carbon dioxide (CO_2) emissions on the planet come from coal combustion. Recent studies reveal that overall CO_2 emissions of the world in 2016 are 32.1GtCO_2. After China and USA, India is the top CO_2 emitting country in the world [74]. Iron and steel industry is the most energy intensive and highest CO_2 emitting industry contributing 27% of CO_2 emissions from global manufacturing sector [70]. Thus, there is a need to pay attention in iron and steel industry, for reducing CO_2 emissions in order to protect our environment. Over the last few years, India accounts for rise in world's total CO_2 emissions from 1.5% in 1974 to 6.6% in 2015 [21]. The primary fossil fuel energy consumption of India by coal gives strong concern on greenhouse gas emissions and climate change. Nations part of UNFCC approved Paris agreement to make best effort to limit the rise of global temperature by 1.5^{O}C. India targets to cut CO_2 emissions intensity of 33-35% by 2030 compared to the 2005 levels [38].

With these backdrops, theoretical study is carried out in the present work for heat and total site integration of three coal based sponge iron plants such as Process-1, Process-2 and Process-3 being operated in industrial cluster of India.

1.1 ORGANIZATION OF THE Ph.D. THESIS

This thesis has been organized into seven chapters. First chapter briefly introduces coal based sponge iron production and its Indian scenario. It paves the background for the current research work. In Chapter 2, a detailed literature review on different aspects of sponge iron production process has been reported. A brief review on both coal and gas reduction kinetics of sponge iron process is also included in this Chapter. As the present work is related to energy conservation in total site of coal based sponge iron processes a detailed literature review on energy conservation aspects of these processes along with studies on total site integration is presented. The chapter closes with a discussion of the research gaps and objectives of the Ph.D. thesis and its organization.

Chapter 3 presents detailed description of three coal based sponge iron plants being operated with different capacities in an industrial cluster of India that is considered in the present work. Further, physico-thermal properties of raw materials, analysis of different process streams with their composition are studied in detail. Finally, cost figures are reported for economic analysis of energy conservation modifications proposed in the current work.

In Chapter 4 process flow diagrams (PFDs) of all three processes of cluster along with material and energy balanced data are developed and potential areas, where energy is lost and where it is required, are identified. Guidelines are proposed to formulate various energy conservation modifications. Based on these guidelines four design and operational modifications are formulated for energy conservation in three sponge iron industries while considering the potential areas. To represent a particular design modification, required stream data from PFDs are extracted. Along with this, a modified model for coal consumption, to support energy integration studies, is also developed and discussed in this chapter.

Chapter 5 deals with the development of methodology to solve the design modifications proposed in Chapter 4 based on single site integration. Detailed description of the modified approach of total site integration along with its solution methodology is also presented in this chapter. This includes calculating amount of coal and air required by the processes while using these modifications and also carrying out economic analysis to select best possible design modification to existing sponge iron processes.

Chapter 6 demonstrates results predicted from theoretical studies carried out in the present work for heat and total site integration of coal based sponge iron plants. Here all four design modifications, two based on heat integration and two based on modified total site integration, are examined to select best modification based on all economical parameters such as coal consumption, water requirement, operating cost, capital investment, total annual cost, total savings and discounted payback period. Efficacies of these modifications are also computed while comparing results of these based on waste gas generation, energy ratio (ER) and %heat recovery. The results of this research are compared with that of published literature.

Finally, conclusions and future research directions are presented in Chapter 7, followed by a comprehensive list of references and appendices and a list of author's publications based on this research.

CHAPTER 2

LITERATURE REVIEW

In this chapter, a detailed literature review on different aspects of sponge iron production process has been reported. History of direct reduction process for making sponge iron dates back to some 3000 years but its first commercial exploitation took place much later with the installation of Wieberg plant in Sweden in 1952 followed by inception of a DRI plant in Mexico in 1957. However, first commercial sponge iron plant began in India in 1980 with the production of sponge Iron at Kothagudam, Andhra Pradesh [100]. Many investigators worked on different aspects of sponge iron process such as design, operation, reaction mechanism, performance of rotary kiln, selection and availability of raw materials, waste heat recovery system, etc. The literature on these aspects is presented in this Chapter. As simulation and design utilize physical properties of raw material literature review on physico-thermal properties of iron ore, coal, sponge iron, etc. has also been included. A brief review on both coal and gas reduction kinetics of sponge iron process is also included in this Chapter. As the present work is related to energy conservation in total site of coal based sponge iron processes a detailed literature review on energy conservation aspects of these processes along with studies on total site integration is presented. Finally, the modification of the process requires few new heat transfer units such as gas-gas and gas-solid heat exchangers, thus a literature review, related to these units, is also integrated in this Chapter.

2.1 DIFFERENT PROCESSES AVAILABLE FOR SPONGE IRON PRODUCTION

Sponge iron, also known as "Direct Reduced Iron" (DRI) is emerged as prime feed stock, which can replace steel scrap in electric arc furnace (EAF) as well as in other steel-making processes. It is the resulting product of solid state reduction of iron ores where degree of metallization of sponge iron is usually greater than 82%. Principal constituents of sponge iron are metallic iron, residual iron oxides, carbon and impurities such as phosphorus, sulphur and gangue (principally silica and alumina). The final product can be in the form of fines, lumps, briquettes or pellets. Sponge iron when briquetted in hot condition at elevated temperature is called hot briquetted iron.

Reduction environment is needed to remove oxygen and reduce iron oxides to iron. Based on reductants used for this purpose, sponge iron production processes can be broadly divided into two categories namely:

- Coal based processes
- Gas based processes

While majority of the worldwide sponge iron production uses natural gas as the reductant, India uses coal as a major reductant due to its immense coal reserve of 264.5 billion t. Any process can be chosen depending upon the availability of raw materials. In India, coal is available near the iron producing belts and therefore, coal based processes are preferred. As India has adequate coal deposits and its utilization for steel plants is considered to be of prime importance, production of coal based sponge iron is considered as a viable option [2]. In DRI production India deserves special attention, not only because the country is the largest producer of DRI while contributing 25.6% to world DRI production, but also because production is primarily coal based. Out of the global DRI production of 87.1 MT in 2017, the production of coal based DRI occupies 15.33 MT i.e. 17.6% of the total. Further, 96.5% of the total coal based DRI is produced in India only [111].

India has 374 plants with an installed production capacity of 48.63 MT/year. Amongst these 369 plants are coal based producing 75% of the total production [178]. There are 5 natural gas based plants producing 25% of the total annual production. 100% sponge iron demand of India is met from internal sources. Coal based sponge iron plants in India are located near the source of raw materials such as Jharkhand, Orissa, West Bengal, Chhattisgarh, Andhra Pradesh, etc. However, gas based plants are available along the west coast of Maharashtra and Gujarat [146].

2.1.1 Coal based processes

Coal based direct reduction processes involve the usage of carbon bearing materials like coal, gasified coal and coke breeze. Other methods of iron production require the use of coking coal and raw materials of stringent specifications. Since these processes can take variety of raw materials and there is plenty of non-coking coal in India, coal based processes are highly applicable. Reduction processes can be carried out in different types of reactors such as rotary kilns, rotary hearth furnaces and vertical retorts [149]. Amongst all the available options, rotary kilns are widely used as a reactor in coal based plants and important processes applying this technology include: SL/RN, Codir, ACCAR, DRC, TDR, SIIL and Jindal. All these processes consist of some common equipment such as rotary kiln, transfer chute, rotary cooler, dust settling chamber, after burning chamber (ABC), gas cleaning system and chimney.

2.1.1.1 Criticality of coal based processes

A typical process scheme for making sponge iron in a rotary kiln is presented in Fig. 2.1. Coal based direct reduction technologies involve reduction of iron oxides in a rotary kiln using non-coking coal as reductant [108]. Limestone or dolomite is used as desulphurising agent. Coal, iron ore and desulphurising agent are continuously fed to charge end of the rotary kiln and injection coal through the injection pipe at discharge end of the kiln. Through inclination and rotation of the kiln, materials move along the kiln axis and continuously discharge after processing. Processes, which occur as the charge moves towards discharge end are drying, preheating and reduction, which are all controlled by air injected along the kiln length. Normally in most of the commercial plants, air is introduced axially through air tubes fitted at different locations along the length of the kiln, in a direction opposite to flow of feed material. Apart from this, to supply desired heat for the process, air is also blowing into the kiln through central burner pipe located at discharge end. The incoming bed material, consisting of iron ore and a major portion of feed coal, is heated not only by heat transfer through radiation and convection from the wall and waste gas flowing in opposite direction to the charge but also due to heat generated by combustion of coal within the bed of material.

Kiln is divided into two parts: preheating zone and reduction zone. Nearly, half of the portion (lengthwise) of the kiln is considered as preheating zone. Length of preheating zone is dependent on the kiln throughput and other operating conditions. In general, it depends on kiln diameter and length. For larger kiln diameters the preheating time would increase remarkably [17]. This is because the ratio of surface area of charge bed across which heat transfer occurs to volume of charge decreases as kiln diameter increases. In the preheating zone, materials charged are dried and incoming bed of iron ore is heated to reaction temperature of 900-1000°C. Chatterjee and Biswas [27] pointed out that in the reduction zone, beyond 750-760°C, iron ore gets reduced. It was observed that even up to 55-60% of total length of the kiln was utilized only for drying and preheating the charge. Therefore, reduction was actually taking place in only 40% of the total length of the kiln.

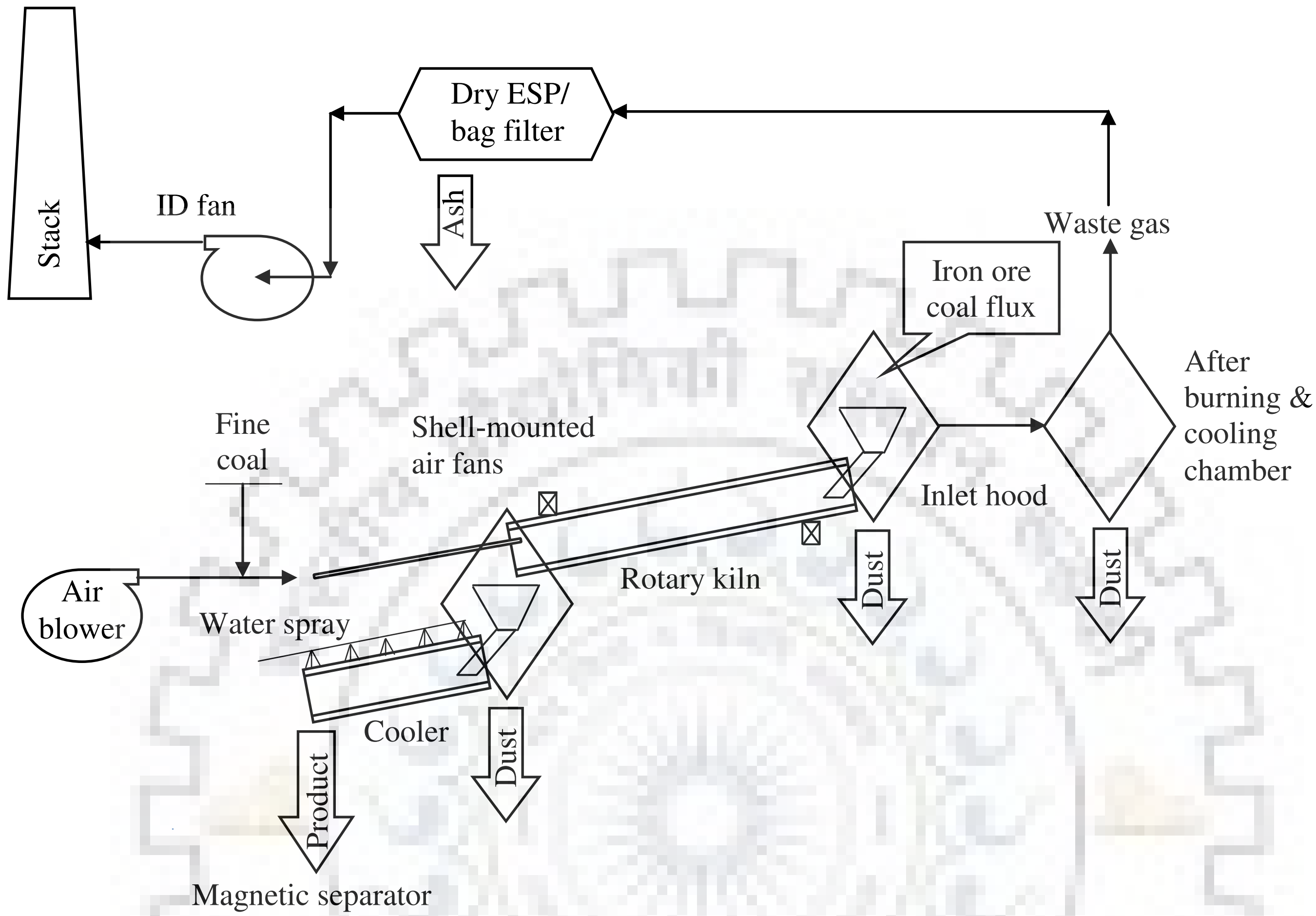
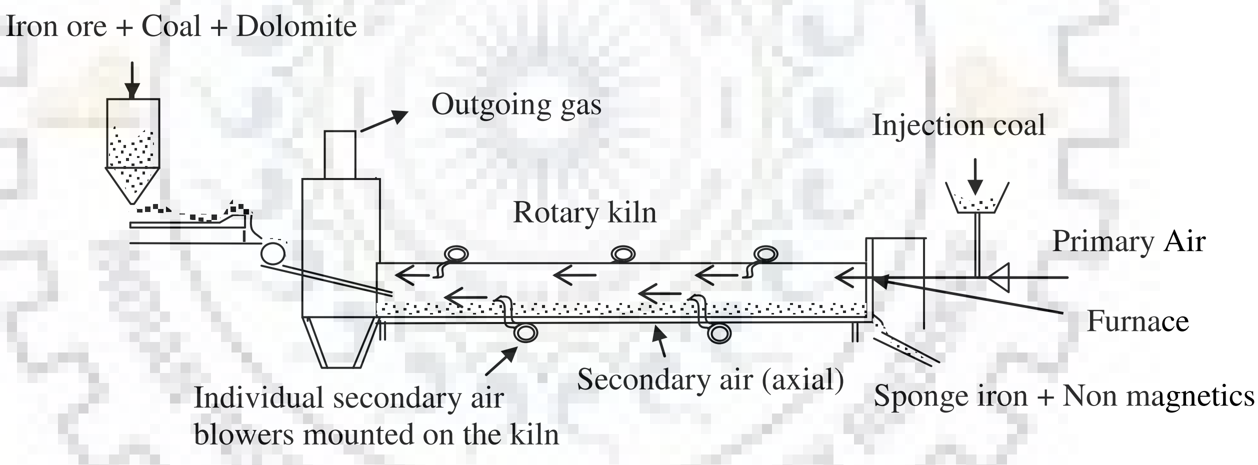


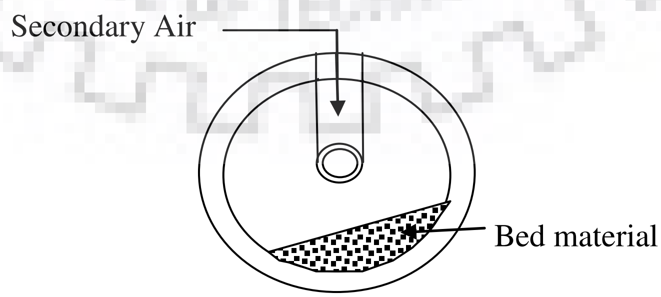
Fig. 2.1 A concise schematic representation of a rotary kiln sponge iron plant [134]

The remaining part of unburnt feed coal within the bed of material and some fresh amount of injection coal, which is injected pneumatically over the heated bed along the outlet portion of kiln, serves to reduce the heated ore. In this zone, the main portion of oxygen contained in iron ore, is removed leaving metallic iron (Fe) and some iron oxide (FeO) behind. Oxygen is removed by carbon monoxide and carbon dioxide, in turn, is converted to carbon monoxide by means of the carbon from the reductant. In order to cover the heat requirement for the process reactions, excessive carbon monoxide evolving from the charge and a small portion of carbon are burned with the air supplied by air tube as shown in Fig. 2.2 [27].

Moisture of feed materials is driven off at initial stage of the preheating zone and most of the volatiles from coal and CO_2 of dolomite are released. Volatile contents of coal, ranging from 28-32%, are volatilized at temperatures up to 600°C where any reducing reaction has not started. One part of these volatiles escapes into the gas space above material bed and the other part is directly burned within the material charge where these are used for direct heat up. Volatiles in the gas space are burned by air admitted through air pipes and thus, supply energy for heating up the kiln charge. Some of these volatiles can be used to preheat the kiln for which a surplus of hot gas is passing through the kiln. However, major portion of these volatiles leaves the kiln with the waste gas without contributing anything to the process. It is to be treated by a costly post combustion and cooling procedure to condition it for the de-dusting system. It is a fact that with the higher coal consumption heat content of the kiln waste gas increases, which is caused by non-utilization of coal volatiles in the process. Therefore, it is only consequent to equip rotary kiln with waste heat recovery systems, which make use of this otherwise wasted energy [17].



(a) View along the kiln length



(b) Cross sectional view

Fig. 2.2 Essential features of a reduction kiln in a coal based sponge iron plant

At discharge end of the kiln, the carbon monoxide generation within the material charge i.e. reducing atmosphere is falling short due to the almost completed reduction. At the same time injection coals are injected to these processes, which is burnt to makeup the lack of carbon monoxide otherwise supplies energy in rest of the kiln volume. With this practice about 30% of total coal is fed through injection to the reduction zone where fresh coal is immediately subjected to temperatures above 800°C at which reduction reactions of sponge iron take place [167]. Balance of 70% of coal is fed at the charge end of the kiln.

A study of heat transfer mechanism inside a rotary kiln allows estimating accurately heat gained by the bed material. Mechanism of heat transfer is very complex because it does not involve only gas and solid, but also wall of the rolling kiln. Apart from conduction through kiln wall, entire heat transfer involves five following components as shown in Fig. 2.3 [17]:

- Gas-to-solid by radiation and convection (Q_{gs})
- Gas-to-wall by radiation and convection (Q_{gw})
- Exposed wall-to-solid by radiation (Q_{ewsr})
- Covered wall-to-solid by radiation and convection (Q_{cwsrc})
- Surface-to-interior of the bed by bulk mixing through conduction and convection (Q_{sib})

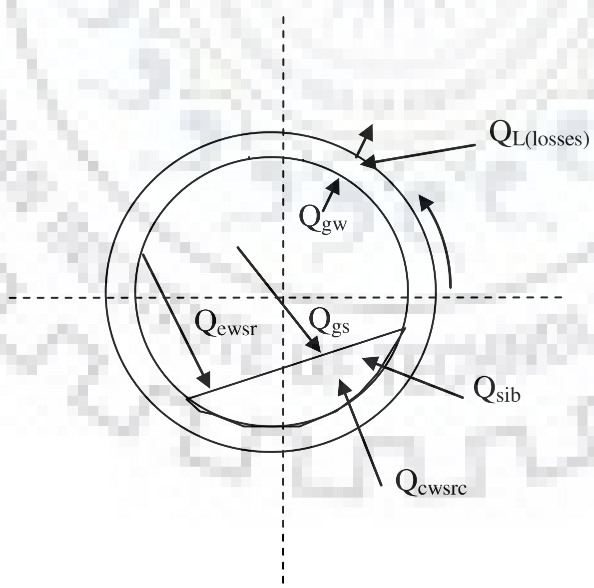


Fig. 2.3 Heat flow path in a rotary kiln

Ventor and Saayman [167] pointed out that at the initial stage ISCOR, South Africa, used a high quality iron ore and low grade coals. Shortly, it became evident that dust losses in the kiln waste gas were up to three times more than what was expected. This resulted in an overloading of dust handling equipment and feed rate of the kiln was reduced to suit the conditions.

Large quantity of waste materials are generated by coal based sponge iron plants in the form of kiln char and fly ash [98]. This material has very little fuel value as it contains a large percentage of ash and it is almost free of volatiles. However, it is realized that char, fly ash and a considerable quantity of coal fines, separated in coal preparation plant, can be utilized to enhance overall economy of the sponge iron plant.

Patnaik et al. [124] shared their experience on the operating experience of the Sponge Iron India Ltd. plant at Polancho. They observed that the success of processes depends on input quality of raw materials i.e. iron ore and coal. The operating experience of authors concluded that use of high grade iron ores with low to medium ash coals is essential for the success of operations. To avoid accretions it is necessary to ensure that gas temperature inside the kiln should be kept within 1100-1120°C and bed temperature should not exceed the outer limit of 1020°C. In spite of all precautions taken in maintaining correct air and temperature profiles, it has been noticed that material build up starts after a period of 60 days, which gets aggravated progressively due to fluctuating raw material quality and interruptions caused by equipment and power failure. Coal with the reactivity 1.0 requires an operating temperature of approximately 1050-1060°C whereas, coal with reactivity of 2.5 only requires 990-1000°C [51]. At kiln temperature 1060°C, required for a lower reactive coal, energy losses are higher and to be compensated by higher combustion rates. In addition, more carbon has to be made available within the bed material in order to ensure high reduction rates and to avoid reoxidation of sponge iron at discharge end of the kiln. Consequently, coal consumption with low reactive coal is higher than with medium reactive coal.

2.1.1.2 SL/RN process

The Stelco-Lurgi/Republic Steel-National Lead (SL/RN) process is a widely used coal-based DRI-making process, which uses a rotary kiln. It is one of the most prominent processes, applied to Direct Reduced (DR) coal based process. The largest commercial SL/RN rotary kiln was designed by Stelco, which was installed in Ontario, Canada, in 1975 [147]. Specifications of the kiln are

125 m length and 6 m inside diameter. Energy consumption in this kiln is 22 million kJ/t. A number of plants are operated in India using SL/RN process.

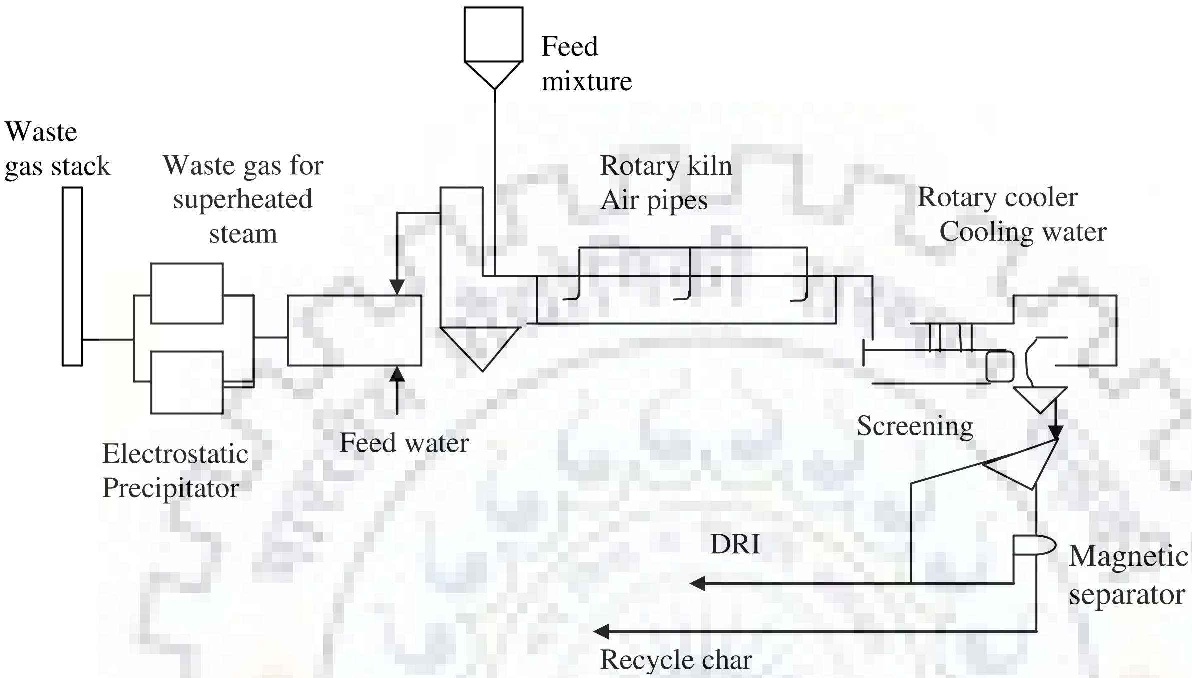


Fig. 2.4 Process flow diagram of SL/RN process

Salient Features of the SL/RN process include:

- Flexibility with regard to the type of iron-bearing materials, which can be used ranging from lump iron ore/pellets to iron sands and steel plants wastes.
- Ability to use a wide variety of solid fuels ranging from anthracite to lignite and even charcoal
- Improved heating of the charge by submerged air injection in the pre heating zone of the reactor
- Optimized facilities for counter current coal injection
- Waste gas conditioning by controlled post combustion followed by waste heat recovery through steam generation.

The process flow diagram (PFD) of SL/RN process is shown in Fig. 2.4. The process operates at high temperature and atmospheric pressure. Materials are fed through fixed inlet pipe to the inclined rotary kiln. The charged materials then progressively move towards discharge end and heated up to 1000-1100^{O}C. To maintain a uniform temperature by burning combustibles released from the bed, air is blown through mounted fans along the length of the kiln. Air is introduced axially in to the kiln and additional combustion air is blown into the kiln through a central airport of discharge end. Solids are discharged from the rotary kiln via transfer chute into a sealed rotary cooler where it is cooled indirectly to 110^{O}C by spraying water over its shell surface. The product is then transported by conveyor belt to magnetic separators to separate iron from char and ash. Waste gases released from the kiln through feed end pass through dust settling chamber (DSC) and after burning chamber (ABC). Further, it is cooled and cleaned in ESP and released to the atmosphere or used to generate steam through waste heat recovery boilers (WHRB).

2.1.1.3 CODIR Process

Coal Ore Direct Iron Reduction (CODIR) process was developed on the basis of experience gained in Krupp-Renn and Waelz process, used for reduction of zinci-ferrous raw materials [32]. This was later extended to the reduction of high grade lump ores. A unique feature of the CODIR process is counter current injection of coarse coal in the size range of 5-25 mm and even up to 35 mm for highly reactive coals. Major part of the total coal (51-66%) is injected from discharge end of the kiln such that bed temperature doesn't fall below 950^{O}C-1000^{O}C. All secondary air tubes are positioned in such a way that air flows towards inlet end of the kiln. In this process coal volatiles can be used for reduction through coal injection system that reduces the requirement of coal to a minimum. This reduces energy consumption from 20 GJ/t DRI to 15 GJ/t DRI [147]. Another distinguishing feature of this process is use of direct mist of water inside the rotary cooler instead of indirect spray of water on the cooler shell. Water mist is controlled in such a way that hot sponge iron is cooled without loss of its significant metallization.

2.1.1.4 ACCAR process

The Allis Chalmers Controlled Atmospheric Reduction (ACCAR) process was developed in l972 at Niagara metals, Canada. The heart of the process is a ported rotary kiln in which natural gas/fuel oil without any external reforming can be used, either singly or in combination with coal as the primary reductant [32,88]. Except for this fundamental difference, i.e. use of gas and coal or oil and coal, keeping coal percentage at a maximum of 80% in both cases, the process is similar to the SL/RN and Codir processes. In the ACCAR process, coal is charged along with the oxide feed, while oil/gas is injected directly under ore bed through radial ports, symmetrically arranged in equally spaced rows. Under bed injection facilities are provided at 20% of the length of the kiln from inlet end. The arrangement of ports and valves is such that oil/gas and air are alternatively injected through the same port. When a particular port is under the charge bed, oil /gas is injected and when the same port approaches above bed position, air is injected through it. This radial oil/gas cum air distribution system is claimed to permit closer control of the temperature profile along the entire length of the reactor.

ACCAR process makes use of 10-20% gases and liquid fuels along with coal as primary reductant resulting in high degree of metallization at low operating temperature, which is the main difference otherwise the process is similar to SL/RN and CODIR process. Along with this lower energy consumption is achieved over other rotary kiln DR processes. Sponge iron produced from ACCAR process contains extra carbon, which is an advantage for electric arc furnace (EAF) steelmaking.

2.1.1.5 DRC process

The DRC (Direct Reduction Corporation) process was developed on the basis of the Hockins process for synthetic titanium dioxide production, originally by Western Titanium Ltd. in Australia [32]. The feed in this process consists of lump ore/pellets, coal and limestone, which are continuously fed to a refractory lined kiln rotating at less than one rpm. Like all rotary kiln processes, combustion air is supplied by shell mounted fans using internal air tubes (secondary air pipes). Coal is blown from the discharge end to supplement the heat generated by the carbon monoxide escaping from the bed. Thermocouples installed along the kiln length accurately measure temperatures and close control of temperature is made possible by monitoring amount of air introduced through air tubes. The product is discharged via a sealed transfer chute to an indirectly cooled rotary cooler and then sent for screening and magnetic separation. Products comprise coarse and fine DRI, non-magnetic fine waste as well as char, which are recycled in the similar way as carried out in all other rotary kiln processes.

Distinguishing features of the DRC process include the followings:

(i) In preheating zone, air is introduced axially in a direction opposite to flow of the feed material whereas in reduction zone it is in a direction opposite to flow of the gases. It is claimed that this method of air injection inside the kiln ensures optimum heat evolution and heat transfer in various zones.

(ii) A part of the coal and char is charged pneumatically into the kiln from discharge end, but unique feature is that the trajectory of counter current reductant injection can be altered by changing the position/inclination of the delivery pipe

2.1.1.6 TDR process

The TISCO Direct Reduction (TDR) process technology developed by Tata Steel was commercialized in a 300 tpd plant, which was upgraded with the association of Lurgi to 350 tpd [32].

Salient features of the TDR process are:

(i) It uses non-coking coal exclusively as the reductant. Fuel oil is used only for preheating the kiln at the beginning of the operation.

(ii) The process prescribes dolomite in a specified size range and proportion as the flux for scavenging sulphur.

(iii) Normal carbon content of the product is around 0.01-0.012%. However, this can be increased to some extent by spraying a small quantity of oil or coal-oil slurry using a special lance introduced through the discharge hood of the kiln.

(iv) There is a provision for radial and axial injection of air up to 30% of kiln length from the inlet end and in remainder of the kiln, air is introduced axially through secondary air tubes protruding up to the centre line of the kiln.

(v) Trajectory and amount of counter current coal injected from the kiln discharge end helps in maintaining a flat temperature profile and proper reducing conditions along the entire length of the kiln.

2.1.1.7 Jindal process

The Jindal process of sponge iron production was developed by Jindal strips Ltd., Hissar in a 15,000 tpa pilot plant [32]. Based on the pilot plant work, Jindal strips installed a commercial plant at Raigarh, which started production in March, 1991. Today, Jindal steel and power has six kilns of 300 tpd and four kilns of 500 tpd at Raigarh making it the largest coal based plant not only in India, but the world as a whole.

Lump ore is charged together with coal and dolomite into the inclined rotary kiln. The discharge moves through the kiln counter currently to hot gases and is heated to reduction temperature. About 40% of coal is charged with iron ore from the inlet side and 60% is injected through the discharge end. Depending on properties of ore and coal, temperature of the charge in the isothermal zone is maintained between 1000-1080°C. After a retention time of 8-10 hours in the kiln, sponge iron is discharged via a transfer chute into a cooler. Here, the material is cooled below 100°C by indirect water sprays and discharged in a manner similar to all rotary kiln processes. Waste gases are passed through WHRBs to recover the sensible heat for producing steam, which in turn is used for generating power. It is claimed that Jindal technology is versatile so far and it can use even inferior grades of coal and iron ore and yet produce sponge iron with more than 90% metallisation. Salient features of this process are:

(i) 55-60% of total coal injected from the discharge end to maintain proper carbon distribution in the kiln. It is claimed that as a result, C/Fe ratio in the feed end is as low as 0.42-0.44.

(ii) Coal up to 30% ash can be used successfully

(iii) The placement of secondary air tubes within the kiln is completely different than that of other coal based technology

(iv) For first time in the world, blast furnace gas is used in this process, which results in better reduction, reduced accretion formation and reduction in the specific coal consumption (250-300 kg/t of DRI)

(v) Specific energy consumption is around 4.2-4.55 GCal/t whereas, specific power consumption is approximately 55 kWh/t

(vi) Steam production by utilization of heat of waste gases from the rotary kiln, which is used in power generation

(vii) Char from the DR plant and rejects from coal washery are also utilized for power generation in a fluidised bed combustion chamber.

2.1.2 Comparative study of coal based processes

Characteristic features of different coal based DR processes are compared in Table 2.1. It shows that energy requirement in SL/RN process is minimum for the given capacity. All processes use rotary kiln, however, SL/RN process is selected for further work as it is oldest and most widely used sponge iron production process in India. The rotary kiln has some advantages, with respect to the process and the product it makes as:

(i) Simultaneous mixing of solid charge while heating is possible in rotary kiln that helps in the dilution of CO_2 concentration formed around the iron ore/sponge iron particles. This is necessary for the reduction reactions to proceed.

(ii) Rotary kiln can tolerate heavily dust-laden gas as a large freeboard volume (about 85%) is available above the solid charge, thus, it would be best suited for utilising the Indian high ash non-cooking coals [143] if it is designed suitably.

(iii) Rotary kiln can serve the dual purpose of a coal gasifier as well as an ore reducer. Preparation of reducing gas from coal is an expensive step, which is commercialized as coal gasification based DR process. Therefore, rotary kiln DR process has proved commercially viable, even with low productivity per unit volume.

(iv) Temperature of reduction of iron oxide is much lower in rotary kiln, which is about 1000^{o}C as compared to 1500-2000^{o}C in blast furnace. Hence, less energy is required for bringing the reactants to reaction temperature [32].

Table 2.1 Comparative study of coal based direct reduction processes [26]

Process/ Parameters	SL/RN	Codir	Accar/ OSIL	TDR	DRC	Jindal	SIIL
Reactor	Rotary kiln	Rotary kiln	Ported Rotary kiln	Rotary kiln	Rotary kiln	Rotary kiln	Rotary kiln
Reactor length (m)	125	74	80	72	80	70.8	40
Reactor ID (m)	6	4	4.8	4.2	5	3.8	3
Capacity	3,60,000 tpa	1,50,000 tpa	1,50,000 tpa	300-500 tpd	1,50,000 tpa	300 tpd	100 tpd
Oxide feed	Lump ore and pellets	Lump ore and pellets	Lump ore and pellets	Lump ore and pellets	Lump ore and pellets	Lump ore and pellets	Lump ore and pellets
Feed end temperature (^oC)	-	-	825	800	800	750-900	650-800
Discharge end temperature (^oC)	-	-	1080	1000	1060	1010-1050	900-1000
Energy requirement (GJ/t-DRI)	14.63-20.9	15	15.5	18.5-21	16.5-18.5	17.5-19	

2.2 POWER AND COAL CONSUMPTION PATTERNS OF THE INDUSTRY

In general, around 35-120 units of electricity are used per t of DRI produced in coal based plants [31]. High power consumption is found in TDR process due to wet dust cleaning system and sludge handling [32,113]. The power consumption for Tata Sponge, Jindal and small plants are 130, 55 and 45 units per t DRI, respectively (kWh = unit) [174]. Earlier, power consumption in Tata Sponge plant is 110-130 units and later it is reduced to 80-90 units per t of DRI produced as system is programmed with logical operated drives and computers are replaced with giant panels. It is also reduced while employing dry gas cleaning system (electrostatic precipitator) [32]. The coal consumptions for different plants are in the range of 1.15-1.45 t/h per t of DRI produced as shown in Fig. 2.5 [26,146].

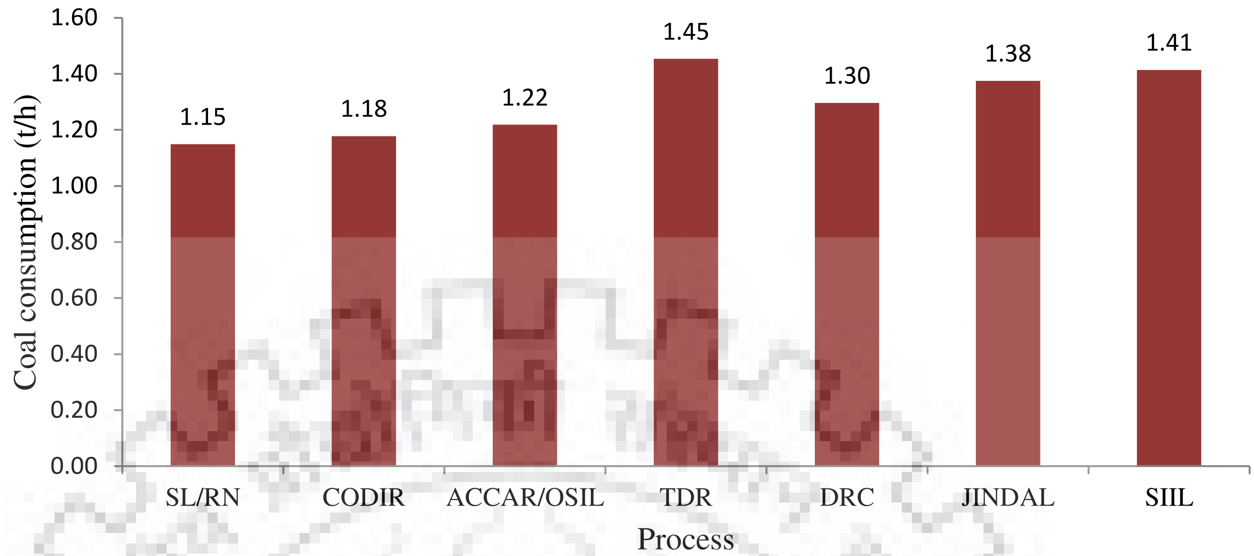


Fig. 2.5 Coal consumption in sponge iron processes

2.3 PHYSICAL PROPERTIES OF RAW MATERIALS

For development and solution of energy conservation schemes and strategies, it is necessary that physico-thermal properties of iron ore, sponge iron and coal such as density and specific heat capacity should be known a priory. These properties of materials generally depend on the presence of different constituents and its composition in the material. A brief review of literature on above properties is presented below:

2.3.1 Iron ore

Gronvold and Sawelsen [65] measured the specific heat capacity of iron ore (Fe_2O_3) or hematite. They measured specific heat of Fe_2O_3 in temperature range 300-1050 K using adiabatic shield calorimetry with intermittent energy inputs and temperature equilibration between inputs. They further compared results with that of Brown and Furnas [22,23] and found reasonable agreement with these published studies. The temperature dependent value of C_p of Fe_2O_3 is:

$$C_p = 24.72 + 0.01604T - \frac{423400}{T^2} \quad kCal/kmol°K$$

The density of iron ore is 5206 kg/m^3.

2.3.2 Sponge iron

Hajidavalloo and Alagheband [68] determined the thermal properties of sponge iron. According to them specific heat capacity (C_p) of sponge iron is 0.46 kJ/kg K. Generally, value of C_p of sponge iron depends on the composition of constituents present in it and C_p value of each constituent. Constituents present in sponge iron are Fe, C, P, S, etc. [69].

2.3.3 Coal

Horace et al. [72] determined the value of C_p of different coals experimentally in the temperature range from 23^{o}C to 223^{o}C. They observed that the specific heat of coal was hardly capable of precise determination as coal was a conglomerate of varying composition, unstable and easily subject to change by heat and exposure to air. Speight [156] found variation of C_p of coal with moisture content, carbon and its volatile matter. The author proposed an empirical relationship between C_p and elemental analysis (%wt) of coal as:

$$C_p = 0.189C + 0.874H + 0.491N + 0.360O + 0.215S$$

Where, C, H, N, O and S are respective amounts (% w/w) of elements in the coal

2.4 SELECTION OF RAW MATERIALS

Principal raw materials used for sponge iron production are iron ore, non-coking coal and lime stone/dolomite if coal is used as the reductant. Dolomite/Limestone acts as a desulphuriser in the process, removing sulphur from the feed mix during the reduction process. It is mixed in small proportion along with other raw materials before charging into the kiln [32,174]. The quality requirements of iron ore and non-coking coal for sponge iron plants are well documented in the relevant Indian Standards [5] as well as other literatures [28,29,33] and are presented in Tables 2.2 and 2.3. Typical raw material characteristics are:

1. Iron ore lump: Fe > 65%, (SiO_2+Al_2O_3) < 3.5%, S < 0.02% and P < 0.035%. Its size should be 5-20 mm.
2. Coal (dry basis): Fixed C > 42.5%, Ash < 27.5%, VM < 30%, S< 1.0%, Moisture < 7%, Caking index < 3. Its size should be 0-20 mm.
3. Limestone contains SiO_2 < 8%, CaO > 46% and MgO should be 8-10%.
4. Dolomite consists of SiO_2 < 5%, CaO > 28% and MgO should be 20%.

2.4.1 Iron ore

Chemical, physical and metallurgical properties of iron ore and its desirable limits are shown in Table 2.2 [11,18]. On the basis of essential and desirable limits of properties iron ore can be selected. Iron ore lump mainly consists of Fe, SiO_2+Al_2O_3, S and P. Its size should be 5-20 mm. Unlike in the conventional steel melting processes, the gangue content of iron ore cannot be separated as a slag. Therefore, it becomes imperative to select an ore with high Fe content and low gangue content, to optimize yield during steel making. Apart from this, to ensure a better kiln campaign life and output, iron ore is subjected to undergo series of tests such as shatter, abrasion indices, and reducibility, etc.

Table 2.2 Criteria for selection of iron ore for rotary kiln DR processes

i) Chemical properties (Dry loss of ignition free basis)		
	Essential limits	Desirable limits
Fe (Total)	62% min.	65% min.
$SiO_2 + Al_2O_3$	7% max.	5% max.
CaO + MgO	2% max.	-----
P	0.08% max.	0.04% max.
ii) Physical strength		
	Essential limits	Desirable limits
Size in (mm)	5-25 (-5 mm = max. 5%) (+25 mm = max. 10%)	5-20 (5-10 mm should be less than 20%)
Shatter index	96% min.	96% min.
Tumbler index	85% min.	85% min.
Abrasion index	5% max.	5% max.
iii) Metallurgical properties		
	Essential limits	Desirable limits
Static reducibility	0.5% per minute	0.6% per minute
Dynamic reducibility	90% (degree of metallization)	95% (degree of metallization)
Reduction degradation (-1 mm) (-5 mm)	 5% max. 20% max.	 3% max. 15% max.

Chatterjee [31] recognized that India is endowed with widely distributed and vast reserves of high grade hematite ore, over 11,000 mt. Stringent selections has to be made to meet the specific requirements of the sponge iron industry with regard to purity, sulphur content and reducibility as well as reduction and thermal degradation indices.

2.4.2 Coal

Desirable limits of coal to be used in rotary kiln are shown in Table 2.3 [31]. Reserves of non-coking coal appear in significant size and constitute 82% of total reserves coal in the country. Table 2.4 shows total reserves of Indian coals [27]. Most of the coking coals are characterized by high ash content and poor washability. Considering mining and washing losses, only about 50-60% of total proved reserves of coking coal are considered available for the production of metallurgical coke. On this basis the net available proved reserves of prime and medium coking coal have been worked out at about 2150 mt and 2300 mt, respectively. It is reported that with 4% growth rate in hot metal production the above quantity may last for another 60-70 years. The country thus suffers from some inherent limitations in its blast furnace iron making.

Table 2.3 Criteria for selection of coal for rotary kiln DR processes

i) Chemical properties (Dry loss of ignition free basis)		
	Essential limits	Desirable limits
Moisture	8% max.	5% max.
Ash	....	24%
Volatiles matter	....	32±2% max
Sulphur	....	0.08%
ii) Physical strength		
	Essential limits	Desirable limits
Size in (mm)	-----	3-25 mm(feed coal)
Caking index	5% max.	3% max.
Swelling index	1% max.	1% max.
IDT of ash	1150°C.	1200°C
iii) Metallurgical properties		
	Essential limits	Desirable limits
Reactivity	2 cc/gm/s	2.5 cc/gm/s

Table 2.4 Indian coal reserves

Description	Proved	Indicated	Inferred	Total (mt)	Percentage
Prime coking coal	3653.2	1237.3	359.4	5249.9	6%
Medium coking coal	3848.9	4275.2	1252.5	9376.6	11%
Simi or weakly coking coal	1554.5	2452.7	714.8	4722.0	6%
Non coking coal	12290.3	22863.9	28196.8	63351.0	77%
Total	21346.9	30829.1	30523.5	82699.5	100%

However, an abundance of non-coking coal available in the country can be sufficiently used for sponge iron production. A large % of these huge reserves are made up of non-coking coals with high ash content, low ash fusion temperature and relatively low reactivity. All of these, both individually and collectively, have adverse effects on the operation of coal based sponge iron plant. Because of the quality restrictions imposed on non-coking coal, which can be satisfactorily used as a reductant in rotary kiln, these are rather limited in this country. These are available only in some parts of the country, which need very careful consideration before using for large number of coal based sponge iron plants in India. The report of an expert committee to assess the requirement and availability of coals suitable for the sponge iron industry highlights that "the type of non-coking coal required by this industry is, perhaps, as scarce as good quality coking coal" [44]. Over and above this, even coal identified as suitable for the sponge iron industry, are not always made available to sponge iron plants since they are linked to traditional users like the railways, cement producers, power plants and refractory manufacturers.

2.5 STUDIES CONDUCTED FOR TOTAL SITE INTEGRATION

Energy conservation in a process plant is the primary demand of present days. To carry out it, process integration principles are used successfully since many decades [81,82]. However, instead of a single plant, energy conservation in the whole site of plants is more profitable, especially in cases where plants are located nearby, as these sites consider energy conservation opportunities available in all individual plants collectively. To conserve energy in such clusters of plants simultaneously a concept of total site (TS) integration is used effectively. Higher-level Heat Integration can be performed using the site utility system for exchanging utilities. This is known as total site heat integration (TSHI) — first introduced by Dhole and Linnhoff [50]. This methodology has evolved from the original applications of pinch analysis. However, TSHI extends considerably beyond the boundaries of the individual process. It addresses the task of optimizing the design of each process and the utility infrastructure in the context of the complete factory (Fig. 2.6).

Dhole and Linnhoff [50] introduced the concept of total site integration where they developed a graphical method of site profiles, which were analogous to the composite curves for individual processes. They used a graphical approach to optimize the overall site utility system. Since a pocket in the grand composite curve (GCC) indicates the intra-process heat recovery potential, they proposed to remove these pockets of the individual GCCs first. Above and below pinch portions of the modified GCCs are shifted appropriately to construct the site source and the site sink profiles as shown in Fig. 2.7. By minimizing exergy loss between the utilities and the total site source and sink curves, the plant fuel consumption and electricity generation are optimized. However, the method did not take into account capital cost of heat exchanger network (HEN) in optimization and did not provide a clear picture for how the steam usage is distributed for the inter-process heat recovery.

Raissi [138] developed analytical tools that helped in understanding interactions between site fuel, heat recovery and co-generation. Further, Hui and Ahmad [73] developed a graphical based procedure for optimizing the overall site utility system and saved 9.2% total cost. They considered capital cost for the heat exchanger networks along with fuel and power costs in the utility system simultaneously.

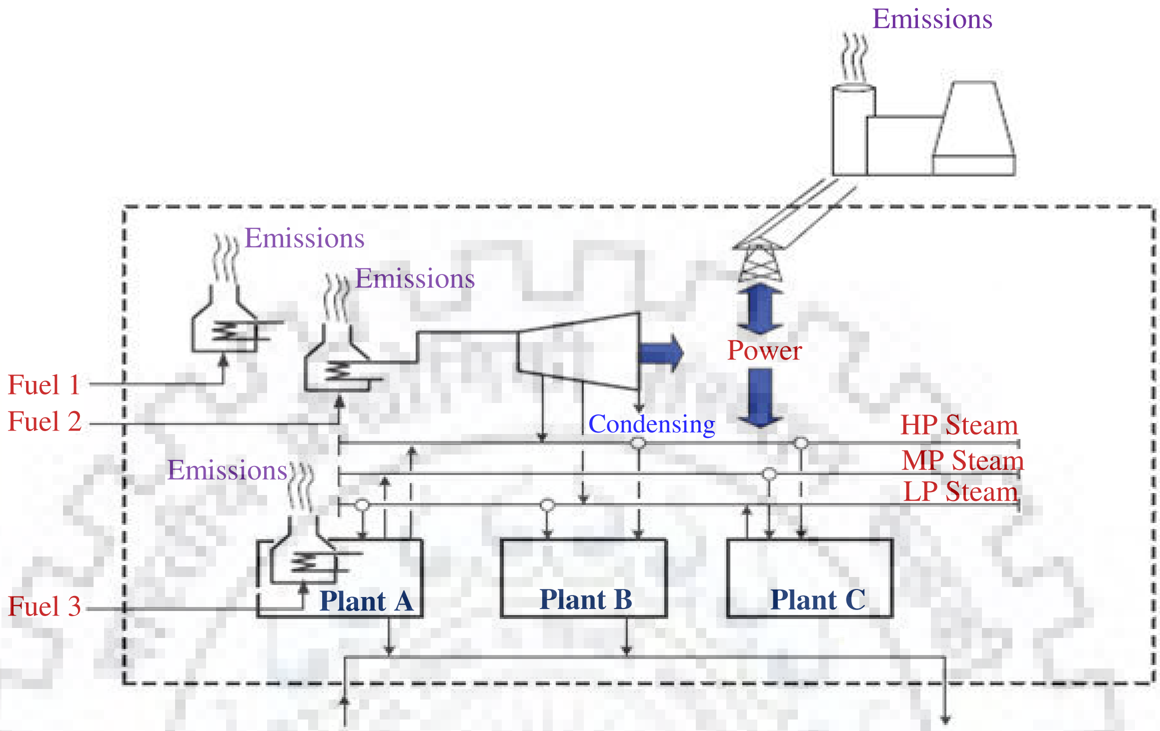


Fig. 2.6 Typical configuration of an industrial Total Site

Klemeš et al. [80] extended the methodology of Dhole and Linnhoff [50] while introducing total site profiles and site utility grand composite curves (UGCC), which consisted of hot and cold utility composite curves. According to them, total site (TS) pinch is the point where the cold utility composite curves (CC) first intersects with the site sink profile (SSiP) or when the hot utility CC first intersects the site source profile (SSoP). The UGCC provides a visual illustration of the external utility requirements. These graphical tools are used to set targets for steam usage and generation by site processes, steam required to be produced by the boilers, and shaft work produced by the steam turbines. The amount of heat recovery that can take place on the total site through the steam mains can be derived from the Total Site Composite Curves. Heat recovery depends also on the number of steam levels as shown in Fig. 2.8. The limit to heat recovery is reached when the two site composite curves touch and cannot be shifted further. Total Site Pinch divides the problem into net heat sink (above) and net heat source (below). The remaining site sink profile heat demand is met by supply of steam from a central boiler. Below the site pinch excess heat is removed by cooling (water) or produces very low pressure steam. The area enclosed by the

Site Composite Curves is proportional to the cogeneration potential of the site utility system. They applied this technique to large industrial complexes, such as chemical process factories and oil refineries, and saved 30% energy and 10% capital cost.

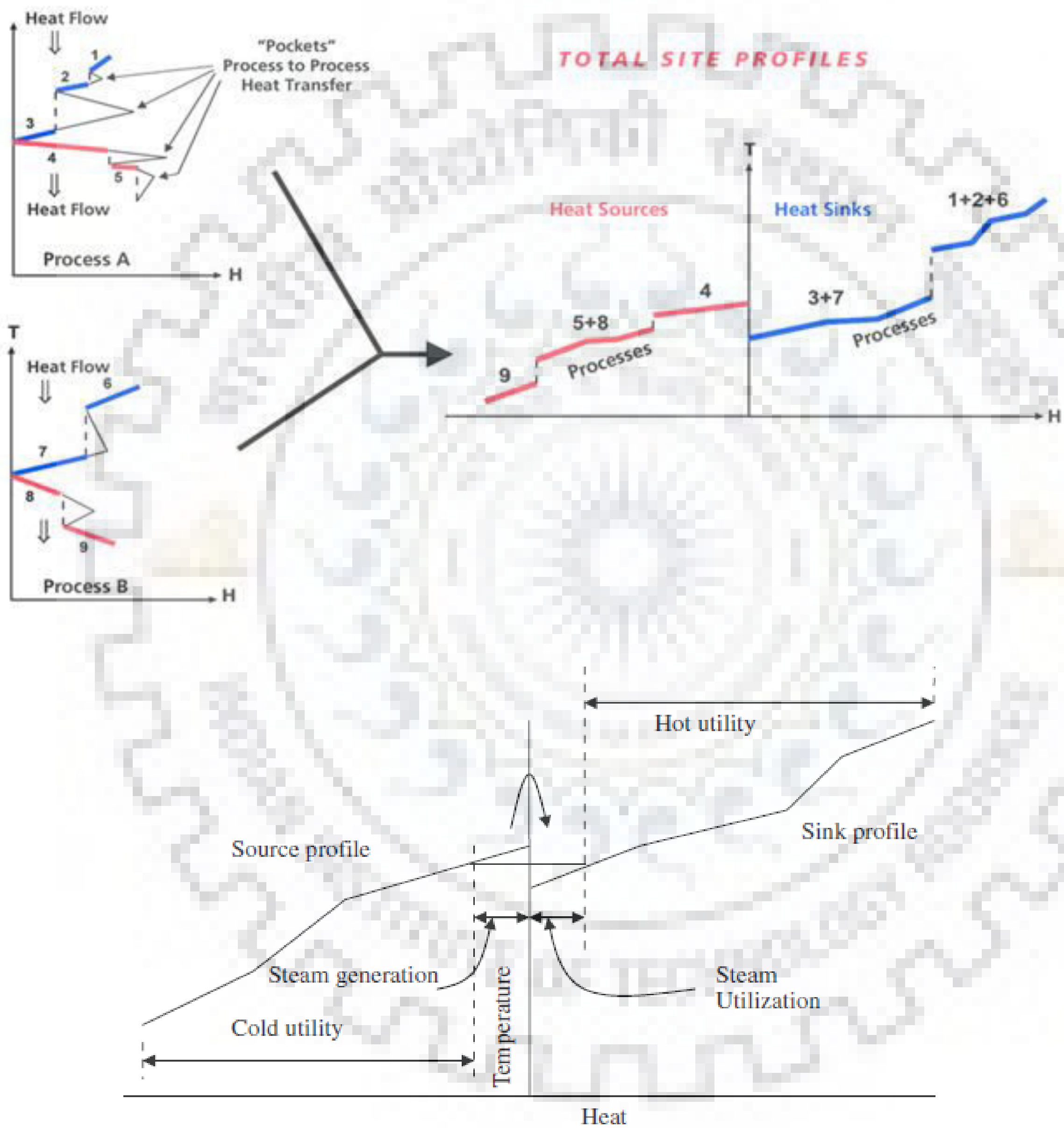


Fig. 2.7 Construction of Site Sources-Sink Profiles for two processes

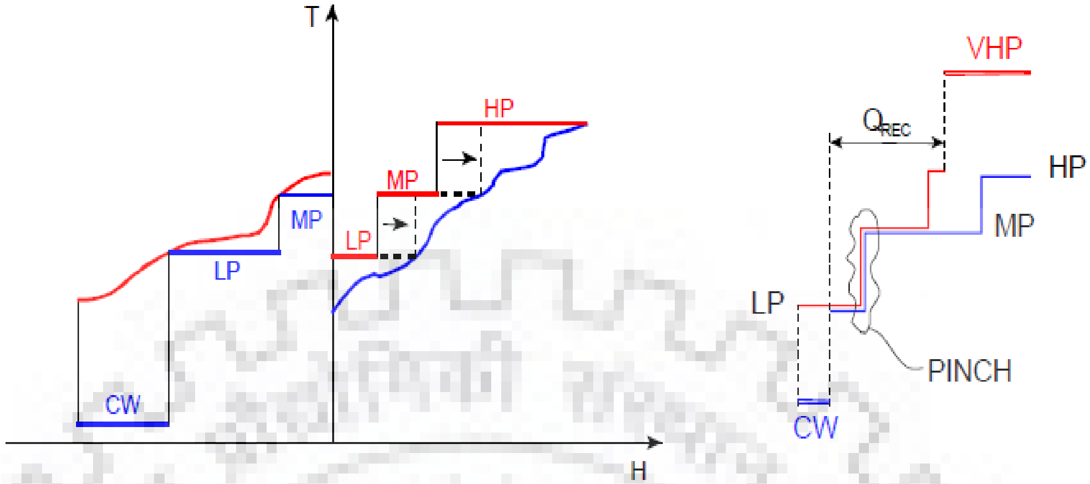


Fig. 2.8 Site composite curve and total site pinch

Goršek et al. [63] presented modified site sink source profiles (SSSP) to determine energy saving potential and designed cogeneration system in an existing site of specialty chemicals while saving 9% hot and 5% cold utilities. Matsuda et al. [106] applied TSI concept to a large industrial area in Japan, which included 31 sites in two blocks, A and B, to further improve performance of highly efficient process plants and found huge amount of energy saving potential in both blocks A and B. Bandyopadhyay et al. [13] proposed a simplified and novel methodology for targeting cogeneration potential based on the Salisbury approximation while utilizing the concept of multiple utility targeting on the site grand composite curve (SGCC). They incorporated assisted heat transfer in generating SGCC. However, a new methodology was proposed by Manesh et al. [105] to determine cogeneration potential of TS utility system, which allowed targeting of shaft work produced and degree of superheat at steam boiler.

For integration of renewable energy with variable supply and demand in industrial sites a graphical targeting methodology was proposed by Varbanov and Klemeš [166]. They introduced time slices into the TS description, with a heat storage system for accommodating these variations. However, numerical algorithmic targeting methodology proposed by Liew et al. [91] for the same purpose of integrating renewable energy with variable supply and demand in industrial sites was found advantageous from the perspective of efficiency and precision as compared to graphical TSHI approaches.

Hackl et al. [66] applied the concept of total site to a cluster of five chemical companies for improved energy efficiency. They designed a site-wide utility system using the proposed concept and showed that 50% of savings could be achieved by moderate changes to the existing heat exchanger system whereas, 92% of the changes were technically feasible. Varbanov et al. [165] extended the TS methodology to allow specification of individual ΔT_{min} for each process and between process and utilities. This offered a step toward providing more flexibility and obtaining site utility targets, which were more appropriate to the individual heat transfer properties of the various site processes. However, Fodor et al. [53] considered stream specific ΔT_{min} inside each process while setting different ΔT contribution (ΔT_{cont}) and also using different ΔT_{cont} between process streams and utility systems.

Matsuda et al. [107] applied TS based pinch technology to target energy saving potential in a large steel plant and identified large energy saving potential, especially in cooler side (power generation of 21.1 MW). Atkins et al. [8] used Heat recovery loops (HRL) to indirectly integrate large industrial sites, which involved low pinch temperatures, such as dairy factories. They found that the effect of thermal storage volume on the amount of heat recovery was highly dependent on hot temperature of loop and nature of the variability of those streams connected to HRL. Kapil et al. [78] introduced a new method to estimate cogeneration potential for site utility systems along with site-wide low-grade heat recovery by a combination of bottom-up and top-down procedures. Ghannadzadeh et al. [61] presented an Iterative Bottom-to-Top Model (IBTM) as a shaft work targeting model to target the cogeneration potential for site utility systems. As IBTM is based on constant isentropic turbine efficiency, the estimation of cogeneration potential depends on accuracy of the efficiency.

Liew et al. [93] developed a numerical technique for multiple utilities targeting in Problem Table Algorithm (PTA) and TS-PTA. They also proposed total site sensitivity table (TSST) for analysing a site's overall sensitivity to plant maintenance shutdown and production changes. They [94] extended TSST for planning TSHI centralized utility system to characterize effects of plant maintenance shutdown, to determine operational changes needed for the utility production and to plan mitigation actions. Further they [92] developed an extended TS-PTA for targeting the total site utility considering water sensible heat. Hackl and Harvey [67] applied the TS Analysis as part of a holistic approach using a framework of industrial clusters for energy efficiency and improved energy collaboration. Pouransari et al. [127] performed site-scale process integration on large chemical plant having three different processes. They represented energy requirements with different heat transfer interfaces and introduced a multi-level data extraction scheme based on these different heat transfer interfaces.

Chew et al. [40] listed numerous effects such as design, operations, reliability, availability and maintenance and economic issues that could significantly affect integration of TS. A heuristic approach was used for developing a matrix containing these effects. Nemet et al. [119] developed a mathematical model for total site heat integration considering issues such as heat losses, piping cost and capital that increase due to higher pressures and temperatures. Chew et al. [41,42] developed a TSHI methodology and explored effects of process modification for achieving better energy recovery. This was extended by Chew et al. [39] for incorporating horizontal pressure drop in steam network. Chew et al. [41] developed a systematic TSHI methodology to target decreasing the capital cost of heat transfer units at TS. Keep hot stream hot and cold stream cold principles are applied to favourably change the TSP shape to provide a larger temperature driving force to further reduce heat transfer area (HTA) and capital costs.

Wang and Feng [170] analyzed the distance related aspects in interplant heat integration and effects of these on total annual cost (TAC). Wang et al. [169] proposed a graphical methodology for determining energy target of interplant heat integration with different connection patterns while considering water as the heat transfer medium. Further, Wang et al. [168] developed a strategy that combined direct and indirect interplant heat integration to reduce TAC.

A TS retrofit framework has been proposed by Liew et al. [89] which considered both individual process and TSHI to improve HEN configuration of an existing petrochemical plant. They found significant energy savings when both direct and indirect heat recovery retrofit options are evaluated. The TS Plus–Minus principle is used in this framework to screen the retrofit options prior to the detailed analysis. The proposed retrofit options for TS reduced hot utility and cold utility by 28% and 23%, respectively, of current consumptions. Boldyryev and Varbanov [19] analysed energy efficiency of bromine production site and proposed pathways for reduction of heating and cooling demands. They reduced hot utility by 57% and cold utility by 97% of current consumptions with improved heat integration. Liew et al. [90] reviewed developments of heat integration and TSHI until 2016. Tarighaleslami et al. [162] focused on unifying TSHI for multiple isothermal and non-isothermal utility targeting.

2.6 GHG EMISSIONS FROM SPONGE IRON PLANTS: INDIA

Global warming is the serious environmental issue causing enormous pressure on all countries to reduce greenhouse emissions. Worldwide there is an increasing emphasis on environmental issues. Iron and Steel industry is growing globally and so are its related environmental issues. The sponge iron industry emits gaseous pollutants. World's primary energy source comes from fossil fuels (coal, oil, natural gas) and its combustion gives the major source of global CO_2 and other greenhouse gas emissions. Largest source of carbon dioxide (CO_2) emissions on the planet comes from coal combustion. Recent studies reveal that overall carbon dioxide emission of the world in 2016 is 32.1 $GtCO_2$. After China and USA, India is the top CO_2 emitting country in the world [74]. Iron and steel industry is the most energy intensive and highest CO_2 emitting industry contributing 27% of CO_2 emissions from global manufacturing sector [70]. China, Japan and India are top three steel producers in the world. Thus, there is a need to pay attention in iron and steel industry, for reducing CO_2 emissions in order to protect our environment. Over the last few years, India accounts for rise in world's total CO_2 emissions from 1.5% in 1974 to 6.6 % in 2015 [21]. The primary fossil fuel energy consumption of India by coal gives strong concern on greenhouse gas emissions and climate change. Nations part of UNFCC approved Paris agreement to make best effort to limit the rise of global temperature by 1.5 degree Celsius. India targets to cut CO_2emissions intensity of 33-35% by 2030 compared to the 2005 levels [38]. India's thrust for expanding her manufacturing sector and increase in energy burden with respect to the development enhances the CO_2 emissions until it strongly focus on more efficient use of energy and energy saving.

The steel making process with general air emission targets for India is discussed in a number of studies including the World Bank Group [172] and Chatterjee [30]. The reports suggest that carbon dioxide emissions from steel production, ranging between 5% and 15% of total country emissions in key developing countries (including India) will continue to grow. This is particularly true in developing countries where outdated, inefficient technologies are still in use to produce iron and steel' [135]. Therefore, energy consumption and efficiency and/or intensity reduction has much more value to control emissions from manufacturing Iron and Steel. Developing countries accounted for 42 percent of Iron and Steel production in 2003 [15] and shared 30% of direct industrial carbon dioxide emissions in 2006 [164]. These statistics are burning issues in the context of global warming, climate change, industrial energy efficiency and carbon trading. Zahan and

Upadhyaya [183] analysed sources of fuel consumed in the basic metal industry (Iron and Steel together with non-ferrous) and reports carbon emissions from such energy sources. India, the gigantic producer in South Asia, has emerged as the fifth largest producer of steel in the world and the second largest producer of crude steel, as well as DRI in recent years [7]. Though the sponge iron process eliminated the need of polluting coke ovens and sinter plants of the conventional blast furnace route of the integrated iron and steel plants, in direct reduction itself, pollution and generation of wastes have been causing concern in industrial belts [25].

In DRI production India deserves special attention, not only because the country is the largest producer of DRI while contributing 25.6% to world DRI production, but also because production is primarily coal based. Out of the global DRI production of 87.1 MT in 2017, the production of coal based DRI occupies 15.33 MT i.e. 17.6% of the total. Further, 96.5% of the total coal based DRI is produced in India only [111]. India has 374 plants with an installed production capacity of 48.63 MT/year. Amongst these 369 plants are coal based producing 75% of the total production [178].

Leading states in sponge iron production are Odisha, Chhattisgarh, West Bengal and Jharkhand. Coal based sponge iron industry is prone to pollution because it is dry thermal reduction process. It is prone mainly to air pollution and also water and noise pollution. The source and types of pollution in the process is presented in Table 2.5 [179]. India has large deposit of iron ore, low cost labour force, highly qualified technical manpower, and geographically advantageous location in south Asia. However, it could not manage the technological change efficiently for enhance production and minimize pollution. Full proof air pollution abatement system for such units is yet to be arrived at. In spite of installation of emission control system, the sponge iron units are also causing environmental ecological disturbances.

Xue et al. [181] showed that Air pollution impact of sponge iron units on the state capital city, Raipur, located 30 km south to these clusters become a high profile issue for both the state and national Governments. Das et al. [46] discussed Environmental management at the selected sponge iron industry in Sundargarh-Jharsuguda area, Orissa, India. The causes of adverse environment impact are identified and actions for mitigation are described. Besides regular pollution control measures, the industry has launched a pilot project to implement CDM protocol, which has dual advantage of waste based power production and carbon Emission Reduction Credit. Swar [159] presented the growth of sponge iron units in Orissa, their pollutions problems present environmental management practices, pollution control regulating framework and enforcement mechanism and strategies for further improvement to prevent pollution. Meikap BC [109] studied on air pollution problems and control measures in steel making through DRI route. Bandhopadhyay et al. [10] reported that Metallurgical and mineral processing industries are always known to be major contributes to environmental pollution. Amongst them the iron and steel sector finds predominance simply because of the significant volume of effluents, emission and solid wastes generated from the various process streams. A large number of innovations in waste management result in implementation of integrated waste management plans in the steel sector as well as development of many value added products. According to Chakravorty [25], suspended particulate matter content of treated DRI gases of rotary kiln are reported by some of the industries to State Pollution Control Board as to be less than 150 mg/Nm^3 in all the plants, irrespective of the pollution control devices applied. However, there appears a gap between the reported ones and recently analysed ones.

As India is the largest producer of sponge iron, there is a need to pay attention in this plant for reducing energy consumption and thus, CO_2 emissions so that minimum waste gas is released into the atmosphere and thereby making the sponge iron production process more environment friendly.

Table 2.5 Sources and types of pollution from the coal based sponge iron process [179]

Process Activity	Source of pollution	Release to the Environment	Pathways	Types of pollution
Raw material unloading, stack piling	Unloading, stack piles, conveyor transfer points	Coal dust, iron ore dust, dolomite dust	Air	Air pollution
Crushing and screening of Raw material	Crusher and screen house	Coal dust, iron ore dust and noise.	Air	Air pollution
Rotary Kiln operation at 1100^{O}C	Kiln waste gasses	Heat, dusts, SO_2, NO_x, CO	Air	Air pollution
Rotary Cooler	Waste water	Heat, water	Water	Water pollution
Product separation and handling	Screens, magnetic separation	Dust, noise	Air	Air Pollution
Dumping of waste (Charity on the ground)	Dump Sites	Dusts	Air	Air pollution
Plant sanitary waste water	Toilets, canteen	Waste water	Pipe line open drain	Water pollution

2.7 STUDIES CONDUCTED ON REDUCTION KINETICS OF SPONGE IRON PROCESS

2.7.1 Coal based sponge iron processes

For coal based sponge iron processes, a range of experimental and theoretical studies have been done on iron ore reduction of coal ore mixtures and its kinetics. Takenaka et al. [160] developed a reactor model in which the reduction rate equations were derived from a three-interface model and involved both mass and heat balances. The model rate parameters were determined according to a parameter fitting method. Xingguo et al. [180] developed a model to predict the effect of all possible factors like preheat temperature of stock, fuel and air; mass ratio of gas phase to solid bed; size of charged materials; rotation speed on the kiln efficiency of rotary kiln processes. Lin et al. [95] studied kinetics of direct reduction of Chrome iron ore and found that in the main reaction region, direct reduction is controlled by a phase boundary reaction. They developed kinetic equations and results obtained with isothermal and non-isothermal methods are found same in the main reaction region. Agrawal et al. [2,3] studied the effect of cold bonded ore–coal composite pellets on the productivity of rotary kiln sponge iron making plant and compared with lump ore operation. They found that kiln productivity increased nearly two folds, coal requirement and residence time in the rotary kiln was reduced compared with lump ore operation.

Coetsee et al. [45] developed non-isothermal and non-isobaric mathematical model to investigate rate-limiting steps for reduction in magnetite-coal pellets and found that heat transfer was not the rate-limiting step. Liu et al. [97] examined direct reduction of iron ore with an Australian coal using advanced experimental techniques and found that coal devolatilization and iron oxides reduction occur simultaneously during heating of mixture. They postulated that H_2 and CO gases produced from coal devolatilization and char gasification were responsible for the reduction of iron oxides. The evolution of CO for coal commences at a temperature of 450^{O}C and ends at temperatures of around 950^{O}C. The highest rate of evolution is attained at 720^{O}C. On the other hand, H_2 commences at 495^{O}C reaches the highest rate of evolution at 785^{O}C and ends at 1000^{O}C.

A mathematical model of the coal-based direct reduction process of iron ore in a pellet composed of coal and iron ore mixture was investigated using finite-control volume method by Shi et al. [151]. They reported that the effect of convection on temperature, average reduction and concentrations of reaction components are small. It was also found that overall uniform surrounding temperature distribution is better for reduction. Mohanty et al. [116] optimized the performance of rotary kiln of OSIL based on production quantity and product quality using optimization procedure by ANN. Sun and Lu [158] studied kinetics of the carbothermic reduction of iron oxides in a composite pellet made of taconite concentrate and high-volatility coal by means of mathematical modelling that simultaneously considered the transfer rates of both mass and heat, and rates of chemical reactions. They found that rate of carbothermic reduction of iron oxides increases with an increased furnace temperature and decreased pellet diameter and overall rate is predominantly limited by heat transfer within pellet. From kinetics viewpoint, the optimum composition of the composite pellet is approximately in accordance with the stoichiometry, when CO is assumed to be the sole oxide of carbon in gas.

Zuo et al. [185] investigated reduction behavior of iron ore using biomass char and compared with that of coal and coke. Compared with the pulverized coal and coke, biomass char has a higher reactivity and reaction temperature of hematite reduced by biomass char is at least 100 K lower than that of coal and coke. Man et al. [104] reported that reduction process was diffusion controlled below 900^{O}C; however, at higher temperatures up to 1100^{O}C, the reduction was phase-boundary controlled. Man et al. [103] studied the mass loss and direct reduction characteristics of iron ore-coal composite pellets and found mass loss rates are influenced by temperature (between 800^{O}C-1100^{O}C) and not by heating rate.

2.7.2 Gas based sponge iron processes

Many researchers carried out extensive study on sponge iron processes outside of India also. Most of these studies are on gas based sponge iron processes. Parisi and Laborde [123] proposed a model for solid-gas counter-current moving bed reactor used in direct reduction of iron ore with H_2 and CO gas mixture. It was found that if complete metallization is reached, the production would be lower by 30%. Also if the production increased, the metallization level would fall below the level required by the steel mill. Production of reduced iron is found increasing with increasing CO/H_2 ratio. Piotrowski et al. [125] investigated the effect of gas composition on kinetics of iron oxide reduction from Fe_2O_3 to FeO at temperature range from 973K to 1183K through thermo gravimetric experiments. Once a thin layer of lower iron oxides (magnetite, wustite) is formed on the surface, the data analysis indicates that the mechanism clearly shifts from surface control to diffusion control. The reaction rate increases with temperature increasing and decreases with the CO content in the inlet gas. The comparison in activation energies proved better reducing capabilities of H_2 compared to CO.

Weiss et al. [171] investigated reduction kinetics of iron ore fines under fluidized bed conditions with hydrogen-rich gas mixtures in the temperature range of 400-700^{o}C, and the influence of both temperature and reducing gas composition on the reaction rate was considered. They found that for reduction of hematite to magnetite, the order of reaction of H_2 was 0.036. Reaction orders were 0.09 and 1.89 for the reduction of magnetite to wustite and wustite to iron, respectively. Nouri et al. [121] presented non-isothermal, steady state and heterogeneous model for simulating the behavior of direct reduction moving bed reactor for the production of sponge iron. They found that by increasing the H_2/CO ratio and solid flow rate, solid conversion decreased. The solid conversion increased by increasing gas flow rate and reduction of small iron ore pellets required lower residence times to reach complete conversion compared to pellets of larger diameter.

Yi et al. [182] studied the effect of gas composition and temperature on gas-based direct reduction of iron ore pellets. They found increase in reduction rate with increasing in temperature and hydrogen content of reducing gas. At 950^{o}C, for most part of reduction process (>90%), interfacial chemical reaction is the rate controlling step whereas in the last stage (<10%), diffusion of reducing gas becomes the controlling step. Recently, Rahimi and Niksiar [136,137] proposed a grain based model (GBM) and observed that solid conversion predicted by GBM is slightly higher and showed more accurate prediction of gas temperature than that of the shrinking core based model? Mousa et al. [117] investigated the reduction behavior of natural lump iron ore using different gas mixtures (RCOG, OCOG, OCOG-BOFG, RCOG-BOFG) and compared with that of reformed natural gas, which was applied to the commercial direct reduction processes. The reduction degree was increased with temperature to different extents based on the applied gas mixture and reduction with reformed coke oven gas (RCOG) showed higher efficiency due to higher H_2 content. Zhang et al. [184] studied the reduction behavior of fine iron ore particles in fluidized bed and identified significant stickiness between iron ore particles within temperature range from 600^{o}C to 675^{o}C.

2.8 STUDIES CONDUCTED FOR ENERGY CONSERVATION IN COAL BASED SPONGE IRON INDUSTRY

The sponge iron production process which involves reduction of iron oxide at higher temperatures. Today, India is the largest producer of sponge iron in the world while contributing 25% to world DRI production due to adequate availability of coal deposits [110]. Sponge iron plants are highly energy intensive and energy utilization in these plants is found below to the optimum limit. As energy conservation in process plant is the primary demand of present days many authors considered energy conservation and recovery aspects of this industry.

Bandyopadhyay et al. [11] discussed that if very high volatile matter coals were used in sponge iron making most of the volatile matter was lost at the kiln inlet when coal encounters counter flowing hot waste gases. Due to this 30-40% of the total energy joins the waste gases. Similar facts were also drawn by Elsenheimer and Serbent [51]. They proposed many options to recover energy of waste gas that has 40% of sensible or chemical energy in it. However, by installing these options, they did not show that how much energy can be saved and what are their practical implications. The energy cost is the principal cost factor in sponge iron plant as energy requirement ranges between 14.63 GJ/t to 20.9 GJ/t Ulrich and Tandon [163].

Considering waste gas composition, air requirement, dust loss, etc. Jena et al. [77] suggested a quantitative analysis in 14.07 t/h capacity coal based sponge iron plant. They found that 174.28 GJ/h of heat was generated inside the kiln due to combustion and chemical reactions and heat value of coal that entered the kiln was 323.2 GJ/h resulting in a thermal efficiency of the process to be 53.9%. The authors suggested a waste heat recovery system to recover heat of waste gas that contained 33% of heat generated in the kiln. However, they did not show the computation to meet this fact.

Agarwal and Sood [1] considered energy consumption aspects in these plants and reported that out of total heat generated from combustion of coal inside the rotary kiln, only 35% was used in the reduction of iron ore and remaining amount of energy goes out of kiln as waste. They proposed to utilize the untapped energy of waste gas by installing WHRB to generate steam. However, this heat recovery system was also linked with some problems such as: (i) Steam quality is not suitable for turbine application during start up (ii) generation of steam gets affected when kiln operation is interrupted due to accretion formation, shut down and unavailability of feedstock (iii) additional capital is required (iv) In normal boiler operation, quality and quantity of steam is controlled by boiler firing whereas in WHRB has to perform under kiln operating parameters (v) kiln waste gas contains large amount of dust particles.

A few authors suggested improvements in energy efficiency of the process by modifying rotary kiln design. Biswas et al. [17] suggested that 10-12% amount of energy could be saved by controlling axial and radial air injection. They also presented the potential for energy savings in sponge iron plants through the estimation of heat transfer rates using temperature profile of the kiln. Further, authors suggested use of air jet seals instead of mechanical seal arrangement in the rotary kiln and cooler for reduced power consumption. The literature indicates that the power consumption, for Indian sponge iron industries, ranges from 45-130 kWh/t [123]. Eriksson and Larsson [52] carried out an energy survey in sponge iron plant for identifying largest energy losses mainly through exhaust gases. To enhance energy efficiency of the process, which was 40% of the total efficiency, they suggested various ways such as using energy which was otherwise wasted for internal purposes, supplying it externally or using for power generation. They also proposed modification in rotary kiln design to increase efficiency.

Misra and Ipicol [114] found that during the operation in the coal based sponge iron plant, tremendous amount of heat was generated and a significant part of this heat associated with the waste gas, remained unutilized. On the other hand, Mignard and Pritchard [112] reviewed the sponge iron process for transmission and storage of remotely generated marine energy. Thermal analysis of sponge iron was discussed by Hajidavalloo and Alagheband [68] through EAF route, where raw materials were preheated using waste gas energy before entering the furnace. To improve its performance they introduced a new technique that resulted in 14% reduction in energy consumption and 13% increase in production.

Prasad et al. [129] considered a conventional coal based sponge iron plant from India and identified potential areas for energy conservation. They developed a model for coal consumption and proposed design modification based on heat integration. It resulted in the reduction of energy consumption from 5.57 to 4.9 times in comparison to the existing system. Prasad et al. [130] proposed design modifications for sponge iron industry using principles of process integration, which eliminated the use of evaporating chamber resulting in 72.6% and 30.5% reduction in water and coal consumption, respectively. Prasad et al. [131] proposed various design modifications with an aim to integrate energy in coal based sponge iron plant. They found that the existing system consumed 7.1 times more energy than the theoretical value whereas best modification reduced this difference up to 6.2.

Kumar and Khanam [83] designed a water bath around rotary cooler to recover heat from kiln outlet and gas-liquid finned double pipe heat exchanger for heat integration. Through these modifications they reduced coal and water consumption by 7.2% and 94.3%, respectively. Sahu et al. [145] designed waste heat recovery system and gas-gas plate heat exchanger to integrate the heat of stack gas in the sponge iron process. Prasad et al. [128] designed a gas carrying duct and rotary drier for utilization of waste heat in sponge iron plant which reduced the load on the kiln by 13.15% in comparison to the existing one to produce 1 t DRI.

Loganathan and Sivakumar [98] developed a waste heat recovery steam generator for sponge iron industry and saved approximately 50.2 paisa per kg of steam generation as compared with fired boiler. Coal based sponge iron plants have very high average heat to power ratios, often in the range of 25:1 to 31:1 [43]. This reflects that coal based sponge iron plants use mostly thermal energy in comparison to electrical energy. Dey et al. [48] observed that process efficiency of a conventional coal based sponge iron industry was 51.31%. They also found that the actual energy consumption was 45.2% more than the theoretical and 43.5% of energy was lost with the exhausts. So, all these studies are aimed to reduce the coal consumption in these plants which is the only source of heat.

2.9 RESEARCH GAPS

As mentioned in above sections India is the leader in sponge iron production. India has 374 plants with an installed production capacity of 48.63 MT/year. Amongst these 369 plants are coal based producing 75% of the total production [178]. Although most of these plants have acquired the desired level of operational efficiency, energy utilization of various units is found below to the optimum limit due to lack of proper integration techniques, non-optimal process parameters, high energy consumption and old running process technology. Hence, a large scope exists to make sponge iron industry more competitive by cutting down its internal losses using modern technology. As discussed in Section 2.6, few investigators proposed different energy conservation schemes to utilize heat of waste gas using process integration principles. However, these studies considered only single plant for energy integration whereas, in many states of India there are hubs of coal based sponge iron plants, which are placed very near to each other in the mineral belts of India [146]. Hence possible scope for further research exists for energy conservation in these clusters of coal based sponge iron plants as these sites consider energy conservation opportunities available in all individual plants collectively.

To integrate energy in such clusters of plants simultaneously a concept of total site integration can be used effectively. But, it appears from literature that no author has reported the application of Total Site Integration concept in sponge iron industry. Also, for total site integration industrial cluster/site considered in literature, as seen in Section 2.7, includes plants of different process such as heavy chemical complexes [106]. As all the plants in sponge iron clusters are similar, there is no relative advantage of conventional TSI, which is more suitable to integrate heat within the site of different processes. Further, it is observed that in many processes coal is used as utility stream where it is heated up to reaction temperature and then provides heat through combustion. Thus, for such process separation of process/utility data is difficult, which makes integration of the utility network infeasible through conventional TS integration concept. Hence scope for further research is possible to develop a modified approach for energy integration in TS of similar plants.

Also there is a need to develop a modified model to compute coal demand for various energy conservation schemes as previous authors didn’t accounted all the possible factors in it.

2.10 RESEARCH OBJECTIVES

Improvement in energy efficiency in the sponge iron production process can help these plants in dealing with the current issues of energy and water crisis. Also there is a need to pay attention in coal based sponge iron plants for reducing CO_2 emissions. This serves as the motivation for the present work where efforts are made to reduce coal (energy) consumption so that minimum waste gas is released into the atmosphere thereby making the sponge iron production process more environment friendly. Considering the above research gaps and motivation to reduce energy consumption thereby CO_2 emissions theoretical study is carried out in the present work for heat and total site integration of three coal based sponge iron plants such as Process-1, Process-2 and Process-3 being operated in industrial cluster of India, with following objectives:

1. To identify potential areas where energy is lost and can be conserved and where this heat can be utilized in total site of sponge iron plants individually or collectively.
2. To apply heat integration principles for development of energy conservation schemes in all three processes of sponge iron cluster, individually, considering potential areas.
3. To develop a modified approach of total site integration for energy conservation in the cluster of plants of similar type. Further, to propose energy conservation strategies based on the modified approach of total site integration
4. To develop a modified model for computation of coal demand of proposed schemes and strategies.
5. To compare results of total site integrated systems with individually integrated systems, existing site data and published literature.

CHAPTER 3

PROCESS DESCRIPTION AND OPERATING DATA

For the purpose of energy integration, three coal based sponge iron plants such as Process-1, Process-2 and Process-3 being operated with different capacities in an industrial cluster of India are considered in the present work. All these three processes are SL/RN based (Stelco-Lurgi / Republic Steel-National Lead) and the detailed description of these three processes along with operating parameters, role of major equipment and reaction mechanism involved in these processes are discussed in this chapter. Further, physico-thermal properties of raw materials, analysis of different process streams with their composition are studied in detail. Finally, cost figures are reported for economic analysis of energy conservation modifications proposed in the current work.

3.1 FLOW SHEET OF THE PROCESS

Process flow diagram (PFD) of a typical coal based sponge iron process is shown in Fig. 3.1. Different streams in PFD are assigned individual numbers. The stream-wise data such as stream number, mass flow rate of stream and its temperature are presented in Table 3.1 for all three processes of total site. These data, priory to material and energy balance, are obtained from the site directly.

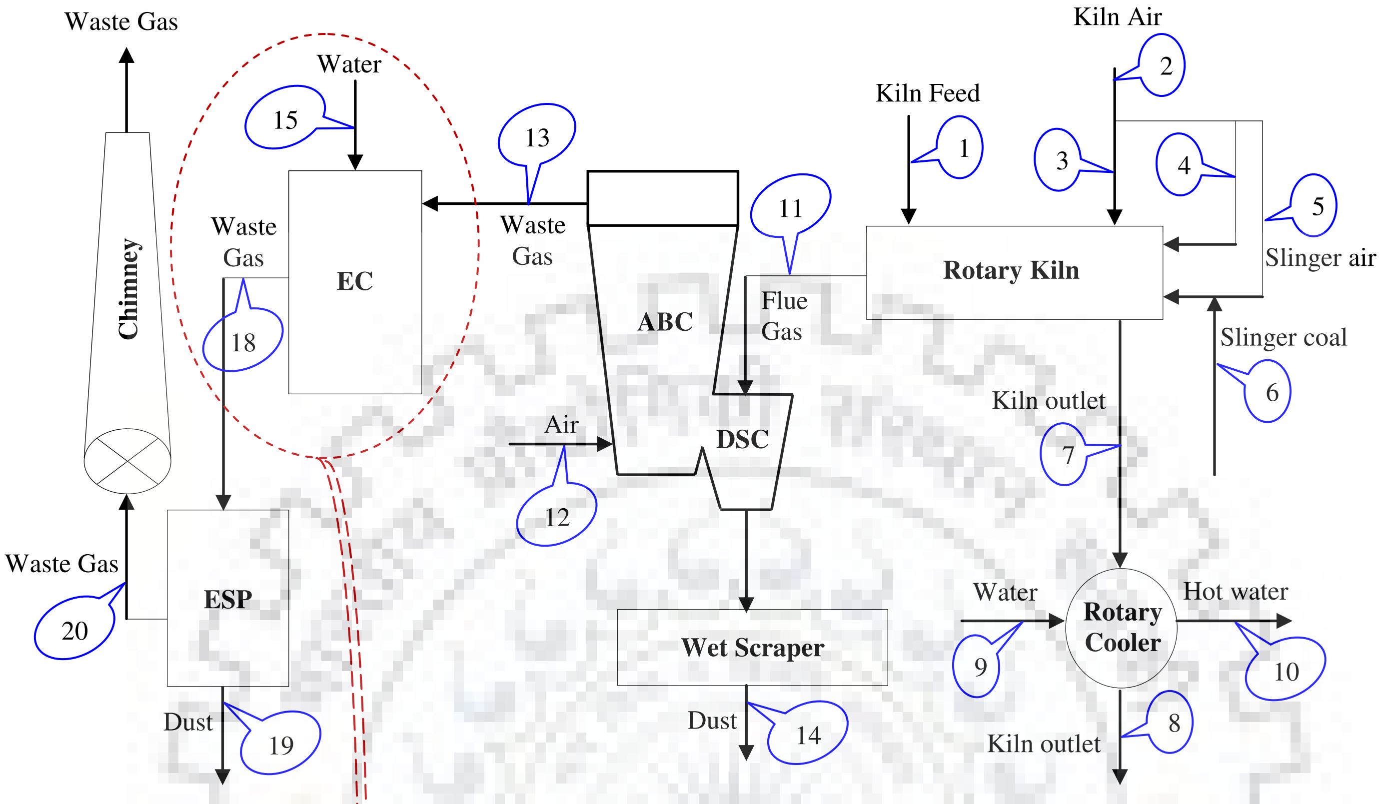


Fig. 3.1(a) Process flow diagram of a typical coal based Sponge iron process

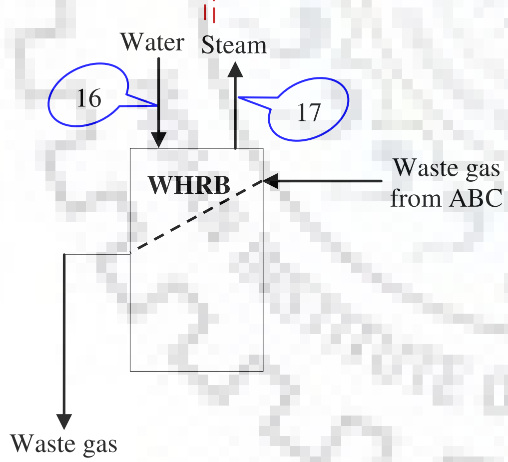


Fig. 3.1(b) Schematic of enclosed section of Fig. 3.1(a) as an available in Process-3

Table 3.1 Stream-wise data of Process-1, Process-2 and Process-3 for PFD shown in Fig. 3.1

Stream No.	Stream	Mass flow (t/h)			T (^{o}C)		
		Process-1 (Capacity: 100 t/d)	Process-2 (Capacity: 62.5 t/d)	Process-3 (Capacity: 100 t/d)	Process-1 (Capacity : 100 t/d)	Process-2 (Capacity : 62.5 t/d)	Process-3 (Capacity: 100 t/d)
1	Kiln feed	9.36	8.73	11.12	30	30	30
2	Kiln air	26.79	34.53	39.25	30	30	30
3	Secondary air	19.3	24.15	28.13	30	30	30
4	Air through central burner pipe	4.61	6.15	6.93	30	30	30
5	Slinger air	2.88	4.23	4.19	30	30	30
6	Slinger coal	2.2	2.6	2.00	30	30	30
7	Kiln outlet	5.40	4.14	5.65	1020	1020	1020
8	Kiln outlet	5.40	4.14	5.65	110	110	110
9	Water	213.63	193.48	251.38	30	30	30
10	Hot water	205.08	185.74	241.32	34.7	34.7	34.7
11	Flue gas	30.26	39.98	44.43	900	750	900
12	ABC Air	3.12	4.79	3.51	30	30	30
13	ABC Waste gas	32.68	44.27	47.24	1050	850	950
14	Dust	0.70	0.5	0.70	80	80	80
15	Water	12.54	16.86	-	30	30	-
16	Water	-	-	13.95	-	-	30
17	Steam	-	-	13.95	-	-	500
18	ESP Waste gas	45.22	61.13	47.24	250	240	250
19	Dust	1.55	0.4	1.30	220	220	220
20	Waste gas	43.67	60.73	45.94	220	220	220

3.1.1 Process description

In the conventional process, shown in Fig. 3.1, kiln feed (Stream No. 1), consisting of iron ore, feed coal and dolomite, is charged into the rotary kiln (RK). A separate conveyor collects different size fractions of coal, also called as slinger coal (Stream No. 6), for injection into the kiln from discharge end with the help of pressurized air (Stream No. 5). The coal injected from the discharge end is mainly used as reducing agent. Only the finer fraction as well as part of volatiles is burnt, which contributes to heat generation. With the help of blowers air, known as secondary air (Stream No. 3), is injected along the length of the rotary kiln at various places. Further, air (Stream No. 4) is injected to the kiln at the discharge end through central burner pipe, which during normal operation serves as process air inlet.

Principally, a rotary kiln for the reduction of iron oxides in solid state through a continuous gas/solid reaction consists of a refractory lined cylinder rotating on its axis, which is slightly inclined at 2.5^{O} to the horizontal towards the discharge end. Three piers are used to support kiln, which rotates with the help of girth gear. To avoid ingress of air mechanical sealing is done at both ends of the rotary kiln. The kiln feed entered to it at a higher level gravitates slowly to discharge end, due to inclination and rotation of the kiln. Time of residence for preheating the kiln feed and also chemical reactions essentially depend on inclination, speed of rotation and particle size of raw materials. Due to lower rotational speed of the kiln longer retention period of material in the kiln and higher filling degree can be obtained. Consequently, there is more time for feed material to undergo reduction reactions in the kiln at same temperature profile, which increases %metallization and thus, the product quality. At the same time, however, production is reduced. Thus, in selecting the rotational speed these two conditions i.e. quality and capacity should be taken into consideration.

Around half of the length of kiln from feed end represents the preheating zone. In this zone volatiles present in coal is combusted and is used to raise temperature of feed to the required temperature for reduction reactions, which occur in second half of the kiln and known as reduction zone. Gases flow counter-current to feed in the kiln. Due to radiation of waste gas more than 80% of heat is transferred to the material. Temperature of material increases with growing retention period and distance from the feed end of the kiln. At the same time, volatiles of feed coal are released into the kiln atmosphere and major part is burnt. The remaining part bums directly within the material charge where these are used for direct heating. Combustion of coal, in first half of the kiln, takes place in the presence of air that releases CO and CO_2 gases. These exothermic reactions increase temperature of feed to the required temperature for reduction reactions. In reduction zone major portion of oxygen contained in iron ore, is removed leaving metallic iron and a few iron oxide behind. Slinger coal, injected from discharge end of the kiln, reacts with CO_2 to produce CO, which acts as the reducing agent to produce Fe from iron ore.

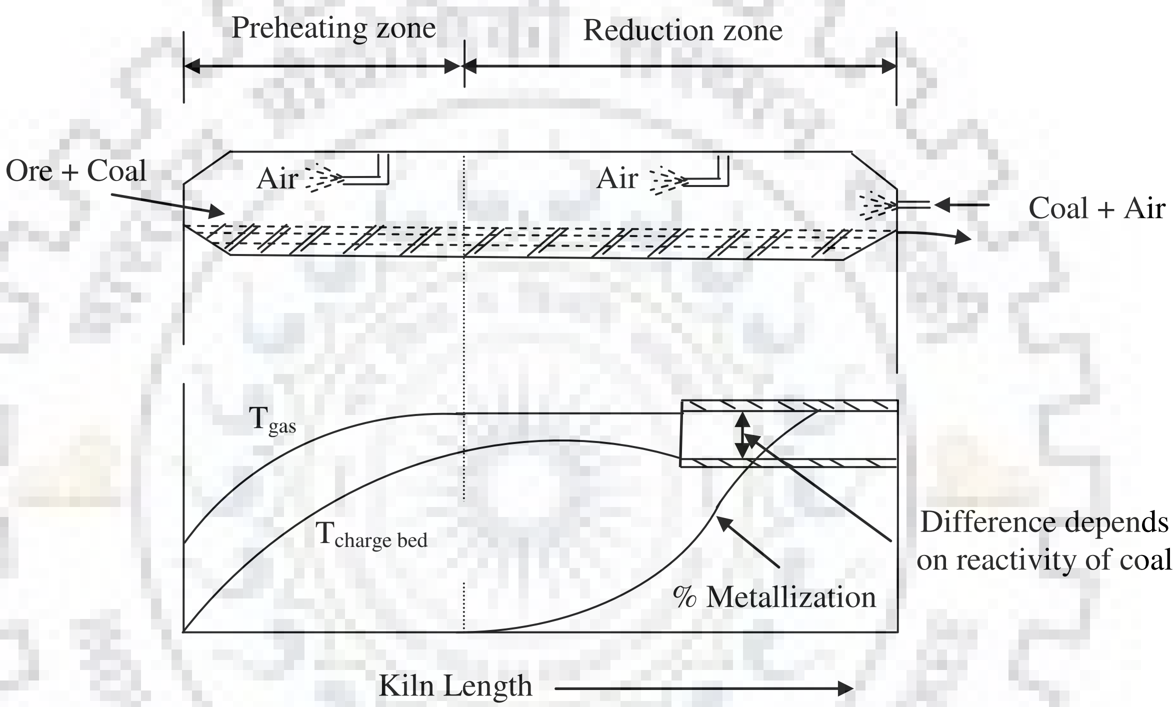


Fig. 3.2 Principle of counter current rotary kiln operation

A precondition for smooth operation of the process is that at no point solid bed temperature, in the kiln, should not exceed 1050°C. Therefore, a controlled air injection is required along the length of the kiln. Injecting too much amount of air leads to reoxidation of the reduced metal and heat generation, which may result in exceeding temperature of the material beyond 1050°C. Temperature profile of free gas T_{gas} and bed material $T_{charge\ bed}$ along the length of the kiln is shown in Fig. 3.2. The difference between these two temperatures is responsible for heating bed charge and metallization of product. Fig. 3.2 also shows the plot of % metallization along the length of reduction zone. It shows that reduction of iron ore starts in this zone and required metallization of the product is achieved at the discharge end of the kiln. Operating temperature range of kiln is from 900^{O}C to 1020^{O}C. The reduced product from kiln (Stream No. 7) i.e. sponge iron with spent lime, char and ash etc. enters into rotary cooler (RC) through transfer chute. Temperature of products is cooled to 110^{O}C while spraying water (Stream No. 9) over the cooler. Further, kiln outlet is sent by means of a vibrating feeder to product separation plant. Using electromagnetic separators products are separated into iron, char and other non- magnetic impurities.

As the operation is based on the counter current principle gases exit the kiln from feed end at very high temperature as shown in Table 3.1. These gases contain some amount of CO and CO_2, which are harmful to human and thus, are treated before releasing to open atmosphere. For this purpose waste gas passes through various equipment before releasing in to atmosphere. Firstly, waste gas exiting the kiln enters the dust settling chamber (DSC). It reduces waste gas velocity, removes large dust particles by gravity, retards pressure fluctuation and achieves uniformity of waste gases with temperature and concentration of combustible. At the end of DSC waste gases change its direction of flow and move upward into combustion area of after burning chamber (ABC) where air (Stream No. 12) is fed for secondary burning. When feed to the kiln has to be stopped, in some unstable conditions during plant operations, an emergency cap located at the top of ABC is operated to release waste gas into atmosphere. Further, gas (Stream No. 13) leaves ABC at high temperature and then cools to around 250^{O}C in the evaporating cooler (EC) using water (Stream No. 15). This is to be done to bring temperature of waste gas to a workable limit for downstream equipment. Further, waste gas (Stream No. 18) passes through electrostatic precipitator (ESP). Here, dust particles (Stream No. 19) are separated and clean gas (Stream No. 20) is released to the atmosphere via chimney (CH). An induced draft (ID) fan is mounted just before the chimney in the plant to maintain flow of waste gas in the plant. In Process-3, unlike Process-1 and Process-2, gas that exits from ABC is passed through a waste gas recovery boiler (WHRB) instead of EC as shown in Fig. 3.1(b).

3.1.2 Role of major equipment

PFD of sponge iron production process, shown in Fig. 3.1, consists of a number of equipment namely rotary kiln (RK), rotary cooler (RC), dust settling chamber (DSC), burning chamber (ABC), evaporating cooler (EC), electrostatic precipitator (ESP), wet scrapper (WS) and chimney (CH), etc. Details of major equipment in the process are discussed in subsequent paragraphs.

Rotary Kiln (RK)

This is the main equipment in the plant which is cylindrical in shape with refractory lining at inner periphery. This is the hearth of the sponge iron plant where reduction as well as combustion reactions is carried out. It plays important role in the conversion of raw material mainly (iron ore, coal, and dolomite) into sponge iron. It works on the principle that as kiln rotates feed moves slowly to the discharge end, due to inclination and rotation of the kiln.

Kiln, which rotates at 4.3 rpm by girth gear, is made up of special alloy of boiler plate and is of 25 mm thick. It is supported by three roller stations and each one of them is equipped with two supported rollers with axles and bearings. For lubrication of the roller bearings, oil pumps are provided and for cooling the oil, water is also used. The kiln is connected with the transfer chute, which is a closed rectangular duct. The primary function of it is to transfer the material coming out from the discharge end of the kiln to rotary cooler. The inside layer is covered with the refractory brick to control heat transfer of the kiln discharge material at this junction. It is sealed in such a manner that puffs cannot come out and air cannot be sucked inside even if the system is working at the discharge mode. Basic components of rotary kiln are shell, refractory lining, support tyres and rollers, drive gear and internal heat exchangers. These components are discussed in details below.

The purpose of tyres and rollers is used to support the kiln and allow it to rotate with least friction. Usually the tyre is mounted loosely on the kiln shell. A small gap is used to allow a differential expansion in between tyre and shell, which is designed to 0.2 % of shell diameter at usual operating temperature. Tyres maintain thermal movement and ride on pairs of steel rollers and set about half a kiln-diameter apart. Bearings of the roller are capable to withstand large live loads involved.

In the drive gear, girth gear is used as spur gear type and it is excellent at moderate speed and noisy at high speed. It is meshing with two pinions, which are placed on both sides of the gear. It has three rollers, made up of stainless steel having high wear resistance property, which support the kiln. The kiln alignment is checked through pulling or pushing the support roller.

The whole rotary kiln is provided with a refractory lining having a thickness of 250 mm, which is fixed by holding rings in axial direction. In the preheating zone from feed end it is lined with super duty fireclay bricks 45% Al_2O_3 and 71% Al_2O_3 castable of 300 mm thickness. Zone at air nozzles are filled with 71% of Al_2O_3 castable. The subsequent reduction zone is lined with high alumina bricks of 80% Al_2O_3 of 250 mm thick, which are resistant against CO and abrasion.

Rotary Cooler (RC)

An electrically driven movable inlet chute is located below kiln firing hood. It transfers hot sponge iron into the rotary cooler, which is a slightly inclined steel cylinder usually having around half of the length of rotary kiln. It is used to cool kiln outlet from 1020^{O}C to operable temperature of around 110^{O}C by spraying water over shell surface. Consequently, water is heated from 30^{O}C to 34.7^{O}C. It is fixed by two rollers, which are rotated with the help of teeth type girth gear. At top surface of the rotary cooler, mixed cold water containing rust resistant chemical flows, which helps in cooling down the kiln outlet

After Burning Chamber (ABC)

After passing through DSC where change in momentum causes dust particles to settle down, flue gas enters into ABC. It is used for secondary burning and also for dust removal. As the name indicates, the hydrocarbon gases, mainly CO, is subjected to combustion to convert into CO_2 for environmental protection. Over ABC an emergency stack is located, which can be opened/closed by means of hydraulically actuated stack cap. In some unstable conditions during plant operations this emergency stack cap is immediately opened and the waste gases are released through it to atmosphere. Temperature of outlet of ABC in Process-1, Process-2 and Process-3 are 1050^{O}C, 850^{O}C and 950^{O}C, respectively, as shown in Table 3.1.

Evaporating Cooler (EC)

Eight to ten water guns are placed along the circumference of EC, at equal height. Temperature of the waste gas is reduced by water exiting through these guns to around 250^{O}C, which is usually the temperature limit for ESP, the downstream equipment.

Electrostatic Precipitator (ESP)

ESP is one of the most effective dust collectors in the sponge iron industry. It uses electrostatic forces to separate dust particles from waste gases. It is acceptable among other collectors due to low pressure drop, high collection efficiency, low sensitivity at high temperature and low maintenance. It has a direct-current high voltage discharge electrode, which is placed between grounded collecting electrodes. As dirty gas enters at the bottom of the ESP and flows upward through these electrodes. Waste gas around these electrodes gets ionized and area is filled with positive and negative ions. Dust particles in waste gas are charged to a saturation level immediately after entering the space. These particles migrate towards collecting electrodes after charging. Movement of particles are opposed by viscous drag and a resultant velocity, called migration drift velocity, is attained by these particles. Dust is then precipitated on the plate electrodes and hence, deposit of dust is formed on the plate. When the layer is sufficiently thick and agglomerated, it is dislodged from plates by rapping and dust cakes slide along plates down into storage hoppers.

Chimney (CH)

This is connected to exit of the ESP. An ID fan is placed before chimney, which pushes up flue gases out through vent by creating negative pressure in the furnace. It is made up of special mild steel alloy. Its dimension is about 50 m height and 3.5 m internal diameter for the present process. Its peripheral surface is surrounded by inclined curved plate and fixed at an angle so that it withstands at the high wind pressure. Its outer peripheral design is such that high wind velocity, which exerts pressure on it, does not affect the chimney and passes through these curved vanes safely.

3.1.3 Reaction mechanism

Following reduction as well as combustion reactions take place, inside the kiln, to convert iron ore into sponge iron:

Reduction reactions:

There are two reduction reactions. In first reaction iron oxide proceeds only to ferrous oxide whereas, in second reaction ferrous oxide reacts with CO to produce metallic iron (Fe), which is also called as sponge iron. Both reactions are exothermic in nature.

$$Fe_2O_3 + CO = 2FeO + CO_2$$

$$FeO + CO = 2Fe + CO_2$$

Combustion reactions:

Most of CO_2 from above two reactions reacts with excess solid in the kiln and is converted to CO according to the following combustion reaction known as Boudouard reaction.

$$CO_2 + C = 2CO$$

Coals with higher reactivity are preferred as they provide rapid conversion of CO_2 to CO and thereby maintaining reducing conditions in the kiln metallization zone. The highly endothermic reaction of coal with CO_2 prevents bed from overheating and attaining high temperature that could lead to melting or sticking of the charge. High coal reactivity decreases the reduction zone bed temperature though increases the relative capacity. Desired bed and gas temperature in the free board can be achieved with high reactivity fuels even with very high throughput rates.

Along with the above endothermic reaction, following exothermic reactions also take place inside the kiln.

$$2CO + O_2 = 2CO_2$$

$$C + O_2 = CO_2$$

$$2C + O_2 = 2CO$$

$$2H_2 + O_2 = 2H_2\,O$$

3.2 PROCESS INFORMATION

For the current energy integration study, the information of sponge iron processes are divided into two different segments namely analysis of different streams and physico-thermal properties of raw materials. Similarly, necessary cost data also provided as capital and operational investments are required to incorporate in design modification proposed in the present work.

3.2.1 Analysis of different streams

Analyses of different streams are required to know constituents present in these as component mass balance depends on these components of streams. These analyses obtained from industrial site are presented below.

3.2.1.1 Iron ore

Iron ore is the source of metallic iron (Fe), which is also called Fe (Total) or sponge iron. The main constituents of iron ore are given in Table 3.2.

Table 3.2 Constituents of iron ore

S. No.	Constituent	%
1	Fe(total)	63
2	Gangue	8
3	Moisture	2

3.2.1.2 Coal and char

Feed coal and slinger coal used in process contain volatile matter, fixed carbon, moisture, nitrogen, hydrogen, ash, oxygen and sulphur. Char produced along with sponge iron contains volatile matter, fixed carbon and ash. The compositions of coal and char are shown in Table 3.3.

Table 3.3 Constituents of coal and char

S. No.	Constituents	Feed coal (%)	Slinger coal (%)	Char
1	Fixed carbon (FC)	40.95	43.10	28.50
2	Volatile matter (VM)	32.05	31.71	1.90
3	Moisture	3.26	3.90	-
4	Ash	23.74	21.29	69.60
5	H_2	3.96	4.04	-
6	O_2	6.59	6.67	-
7	N_2	1.15	1.19	-
8	S	0.70	0.70	-
	Total C	60.60	62.21	

3.2.1.3 Dolomite

Dolomite/Limestone is mainly used as a desulphurising agent in the process, which removes sulphur from feed mixture during the reduction process. It is mixed in small proportion along with other raw materials before charging into the kiln. The main constituents of dolomite are given in Table 3.4.

Table 3.4 Components of dolomite

Constituents	% Composition
CaO	28
MgO	20
SiO_2	8

3.2.1.4 Sponge Iron

Composition of sponge iron is shown in Table 3.5 where Fe (Total) is 92.61%

Table 3.5 Constituents of sponge iron

S. No.	Constituent	%
1	Fe (M)	83.50
2	FeO	11.71
3	Gangue	4.62
4	Carbon	0.17
5	Fe (Total)	92.61

3.2.1.5 Fly ash

The constituents present in the fly ash (a part of Stream No. 11) and its percent composition is shown in Table 3.6.

Table 3.6 Composition of fly ash

Constituents	% Composition
Fe_2O_3	42.40
C	13.48
Volatiles	5.89
CaO	2.29
SiO_2+Al_2O_3	35.94

3.2.1.6 Waste gas

The constituents present in the waste gas (Stream No. 19) and its percentage composition is shown in Table 3.7.

Table 3.7 Analysis of waste gas

Constituent	%
N_2	59.79
CO_2	33.61
CO	1.32
H_2	0.17
H_2O	3.72
O_2	0.80
CH_4	0.59

3.2.2 Physico-thermal properties of raw materials

In the present work mainly two physico-thermal properties of the streams are considered such as specific heat, used for heat as well as energy balance, and density, which is required for mass balance. These properties of various components is collected and tabulated in Table 3.8. Densities of different components are shown in Table 3.9. In the present work the property data of iron ore, coal, dolomite and sponge iron are constant and taken from literature [64,65] whereas properties of air and waste gas [176] are calculated at respective average temperatures.

Table 3.8 Specific heat of streams

Stream no.	Stream	Component	Specific heat, (J/kg^{o}C)		Mass flow rate, t/h (in Process-1)	Mass fraction	Specific heat of mixture, (kJ/kg^{o}C)
			Value	Reference			
1	Kiln feed	Iron ore	960	[65]	6.00	0.64	1103.26
		Feed coal	1380	[175]	3.16	0.34	
		Dolomite	1028.38	[64]	0.20	0.02	
2	Kiln air	Air	1032.60	[64]		1	1032.60
7	Kiln outlet	Sponge iron	594.93	[175]	4.17	0.772	698.86
		Char	1035.30	[175]	0.55	0.103	
		Ash	1067.14	[175]	0.61	0.013	
		Lime	1038.47	[175]	0.07	0.112	
9	Rotary cooler	Water	4187	[64]		-	4187
20	Waste gas	ESP exit gas	1140	[175]	101.55	1	1140
6	Coal (Inlet to kiln)	Slinger coal	1380	[126,175]	7.73	1	1380

Table 3.9 Density of streams

S. No.	Component	Density, kg/m^3	Reference
1	Water	1000	[64]
2	Air	1.12	[175]
3	Waste gas	1.2	[175]
4	Water vapor	0.58	[175]

3.2.3 Cost figures

In the present work various design and operational modifications are proposed for energy conservation in individual plants as well as total site of sponge iron cluster. Efficacies of these modifications are compared based on economic analysis, which requires determination of capital and operating cost.

3.2.3.1 Operating cost

The operating cost includes costs of iron ore, dolomite, coal, water, etc. as reported in Table 3.10 [164].

Table 3.10 Operating cost of the different materials

Material	Cost	
	Value	Unit
Iron ore	2600	Rs/t
Coal (feed/slinger coal)	2500	Rs/t
Water	60	Rs/kilolitre
Electricity cost	3.5	Rs/kWh

3.2.3.2 Capital cost

The capital investment required for implementation of heat integration modification necessitates installation of gas-gas heat exchanger, gas-solid heat exchanger and gas carrying duct. Cost of shell and tube heat exchanger for gas-gas heat transfer is taken from the work of Shenoy [150] and shown as:

$$\text{Cost of shell and tube heat exchanger} = 13{,}68{,}000 + 34{,}200\,(A)^{0.81} \qquad (3.1)$$

Cost of duct = (Manufacturing cost + Welding cost + Insulation cost + Installation cost) of duct

However, the costing methods of gas-solid heat exchangers, ducts and FD fans are considered from the work of Prasad [133] and shown in Appendix C. Also capital cost of boiler turbine is considered as Rs. 46,800/kW and maintenance cost as Rs 0.2/kWh [59].

CHAPTER 4

DEVELOPMENT OF ENERGY CONSERVATION STRATEGIES FOR HEAT AND TOTAL SITE INTEGRATION

This chapter discusses detailed material and energy balance carried out in all three processes of sponge iron cluster. Process flow diagrams (PFDs) of all three processes of cluster along with material and energy balanced data are developed and potential areas, where energy is lost and where it is required, are identified in this chapter. Considering these potential areas, four design and operational modifications are formulated for energy conservation in three sponge iron industries. To represent a particular design modification, required stream data from PFDs are extracted. Along with this, a modified model for coal consumption, to support energy integration studies, is also developed and discussed in this chapter.

4.1 MATERIAL AND ENERGY BALANCE FOR SPONGE IRON PROCESSES

The present work is based on plant data and this has to satisfy the material and energy balances before using it for heat integration. To cut down the complexity, following assumptions are made for carrying out material and energy balances around sponge iron processes as well as for generating energy integration modifications.

4.1.1 Assumptions

1. Variation in physical properties of iron ore, coal, char, ash, and dolomite with temperature is negligible. The reason behind this assumption is:

In the sponge iron process the variation in temperature of above materials is in the range from 30^{O}C to 1020^{O}C and it is reported that density and specific heat of iron ore, coal, char, ash, and dolomite vary considerably with change in temperature. However, in the present work these variations are taken as negligible. The reason behind this assumption is that, the present research deals with comparative studies amongst different design modifications in terms of energy conservation where common variations such as variation of physical and thermal properties due to change in temperature cancel out.

2. Vapor generated in rotary cooler is neglected.

In the actual process, kiln outlet stream is being cooled in the rotary cooler using water as cold stream. This exchange generates less amount of vapor (4% volume of actual water consumption) at considerably lower temperature of 34.7°C. Due to less amount as well as low temperature of vapor, it is not contributing significantly towards heat integration. Thus, the vapor stream is neglected from the calculation.

3. As shown in Fig. 3.1, DSC is attached below ABC. For simplifying the analysis these two are considered as single unit. The flue gas with fines of iron ore, exiting kiln, first enters to DSC where dust is collected and remaining part of it enters to ABC for secondary combustion. Thus, these two equipment are considered as single unit.

4. The ambient temperature is considered as 30°C. The present industrial cluster of sponge iron processes is located in a state of India where in summer the maximum temperature reaches to 45°C and in winter it is around 15°C. Thus, the average ambient temperature for the process is around 30°C.

5. The effect of weather condition is negligible on the quality of raw materials. Due to change of the weather conditions the moisture content of the iron ore and coal varies. However, the variation is not significant and so the effect of moisture in these raw materials is neglected.

4.1.2 Material and energy balances

The material and energy balances of all three processes of sponge iron cluster, shown schematically in Fig. 3.1, are carried out using following steps. Detailed calculations are shown in Appendix A. For material and energy balances specific heat and density of air and waste gas are found at respective average temperatures.

1. Component mass balance is performed around rotary kiln based on plant data that is presented in Fig. 3.1. For this purpose, components such as iron, gangue, carbon and ash are considered for mass balance as these are the main components of feed material, which is composed of iron ore, coal and dolomite. The reduction and combustion reactions involved in the calculation are given in the Section 3.1.3.

2. Mass balance is done around various major equipment such as RK, DSC+ABC, EC, ESP and CH.

3. Overall mass balance is performed around the whole plant considering all inputs and outputs to the process.

4. Energy balance is carried out around rotary kiln. For this purpose, heat of reactions, radiation loss from kiln, vaporization of moisture of coal and ore are accounted. Along with this, sensible heat lost through sponge iron, char, dust, flue gas and volatile matter are also considered as shown in Appendix A.

5. Coal consumptions, for all three processes, are determined using the model developed in the Section 4.5. Air requirement is computed based on oxygen demand of the process and air to coal ratio is calculated based on plant existing conditions.

6. Based on predicted values of coal and air, steps 1 to 5 are revised.

7. Mass and Energy contents of all input and output streams are determined and reported.

4.1.3 Process flow diagram with material and energy balanced data

PFDs of all three processes of cluster along with material and energy balanced data are shown in Fig. 4.1, Fig. 4.2 and Fig. 4.3 respectively. These are discussed briefly in this section.

4.1.3.1 PFD for Process-1

Process-1 is a typical coal based sponge iron plant of capacity 100 tpd being operated in an industrial cluster of India. PFD of Process-1 with material and energy balance data is shown in Fig. 4.1 where kiln feed of 9.36 t/h, which consists of iron ore, feed coal and dolomite, enters to the rotary kiln at 30^{o}C. Usually, coal available in feed material consumes completely while reaching to the discharge end. Thus, to carry out reduction reactions the additional demand of 2.2 t/h coal is supplied with the help of 2.33 t/h pressurized air from discharge end side at 30^{o}C as shown in Fig. 4.1. Throughout the length of the kiln, 17.98 t/h air, known as secondary air, is injected through air pipes connected to blowers, which are mounted on the periphery of the kiln. Further, 4.49 t/h air, known as primary air, is also injected at 30^{o}C from the discharge end of the kiln through central burner pipe, which serves as process air inlet during normal operation. The feed material and air moves counter currently inside the kiln.

In the reduction zone iron ore reduction takes place and metallization increases along the kiln length. Thus, at the discharge end of the rotary kiln required metallization is achieved. The reduced product of 5.4 t/h i.e. sponge iron with char, spent lime and ash etc. exits the kiln and enters into the rotary cooler at 1020^{O}C through transfer chute. It is a cylindrical vessel tilted at 2.5^{O}. Cooling water of 213.63 t/h is sprayed on the shell to reduce temperature of product to 110^{O}C. Products are then sent to screens and magnetic separators for separation of the products. Waste gas of 30.97 t/h flows opposite to feed material and exits the kiln at 900^{O}C. Further, it enters DSC, which is located beneath to ABC, as shown in Fig. 4.1, Waste gas then moves to ABC where air at 2.92 t/h and 30^{O}C is supplied for further combustion. This gas leaves ABC at 33.18 t/h and 1050^{O}C as shown in Fig. 4.1. It is then cooled from 1050^{O}C to 250^{O}C in EC where 12.54 t/h water is sprayed on the hot gas using 8 to 10 water guns. Further, 45.72 t/h gas passes through ESP and leaves dust at 1.55 t/h. The required temperature of waste gas is below 250^{O}C to enter into ESP. Finally, clean gas of 44.17 t/h is released to the atmosphere at 220^{O}C via chimney (CH). Dust that exits DSC is collected through wet scrapper as shown in Fig. 4.1.

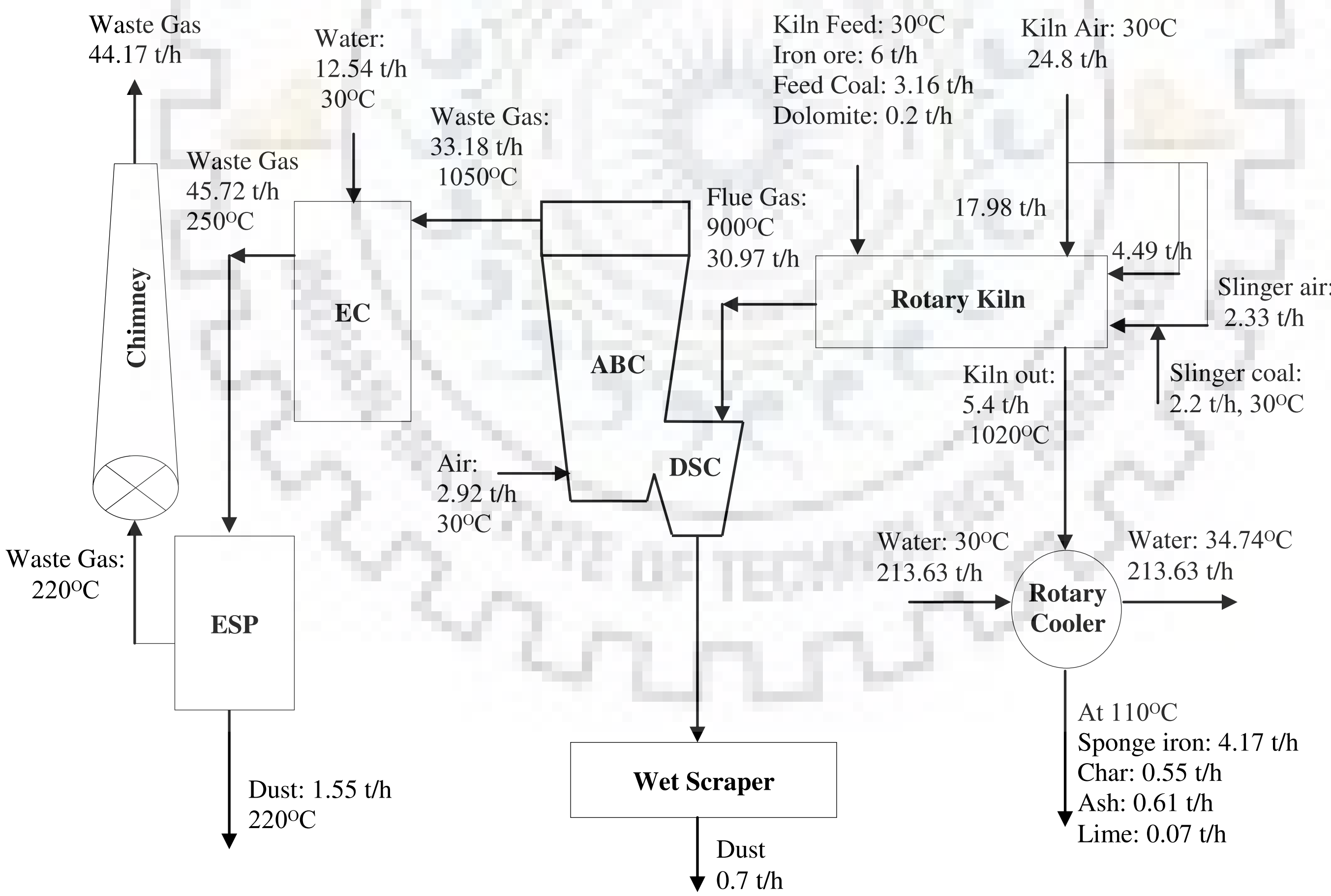


Fig. 4.1 Process flow diagram of Process-1

4.1.3.2 PFD for Process-2

PFD of existing Process-2 is shown in Fig. 4.2. This process is similar to Process-1 where kiln feed of 8.73 t/h is entered to RK at 30°C. Total 33.23 t/h of air is sent to RK for combustion as well as reduction processes. 2.6 t/h of slinger coal is supplied pneumatically from the discharge end. The reduced product of 4.14 t/h exits the kiln and enters to the cooler at 1020°C and is cooled to 110°C using 193.48 t/h of cooling water. Further, this is separated into iron, char and other non- magnetic impurities, screened into different sizes and sent to storage. Flue gas of 40.43 t/h enters to DSC at 750°C and then to ABC for secondary combustion. Further, 61.05 t/h gas passes through ESP before entering into atmosphere through CH at 220°C.

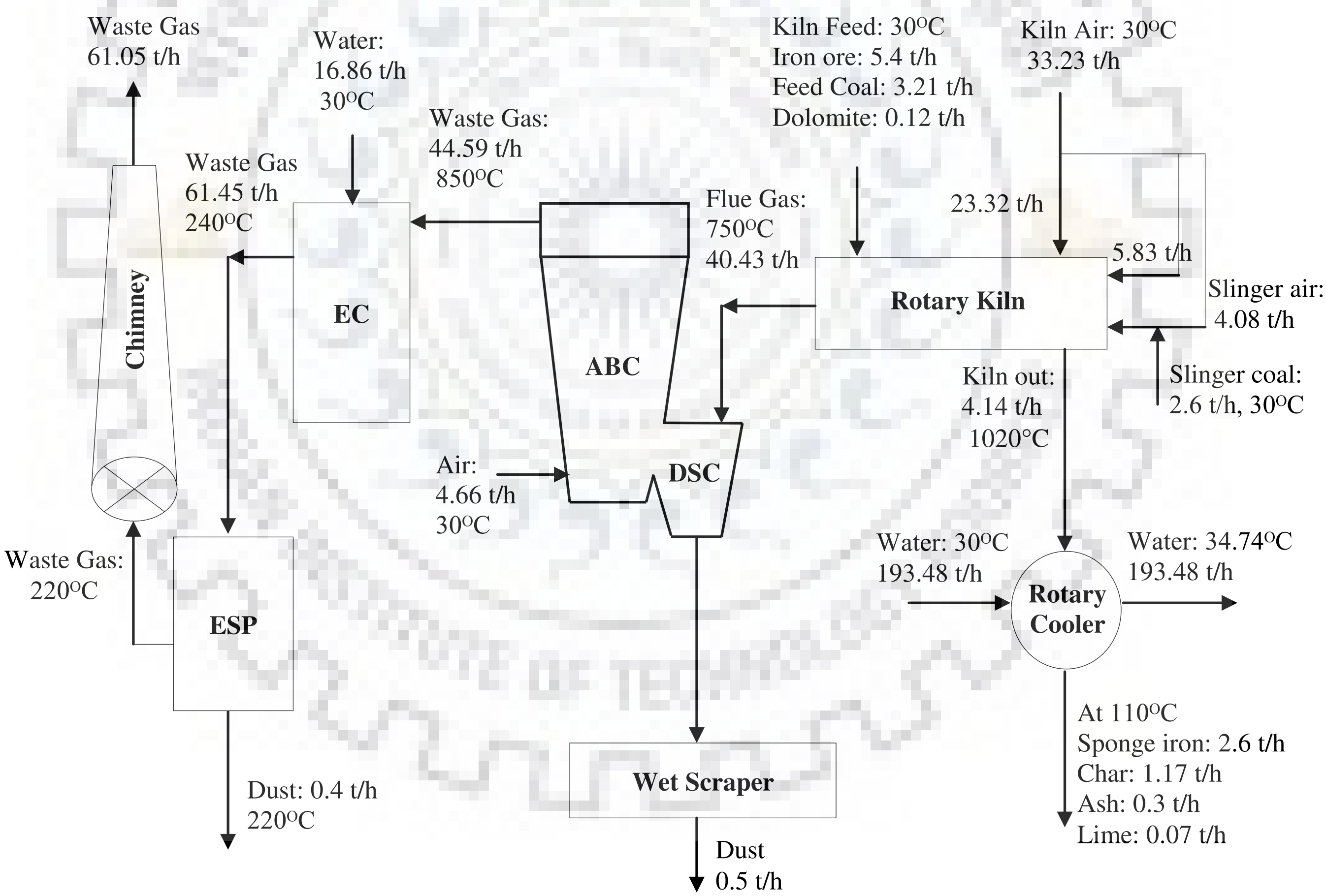


Fig. 4.2 Process flow diagram of Process-2

4.1.3.3 PFD for Process-3

Fig. 4.3 shows the existing PFD of Process-3. The process is same up to ABC as explained in Section 4.1.3.1. From ABC, waste gas enters the waste heat recovery boiler (WHRB) at 950^{o}C where it is quenched to 250^{o}C. Then it is sent to electrostatic precipitator (ESP) for removal of particulate matter. Clean gas of 46.45 t/h is released into the atmosphere at 220^{o}C via chimney.

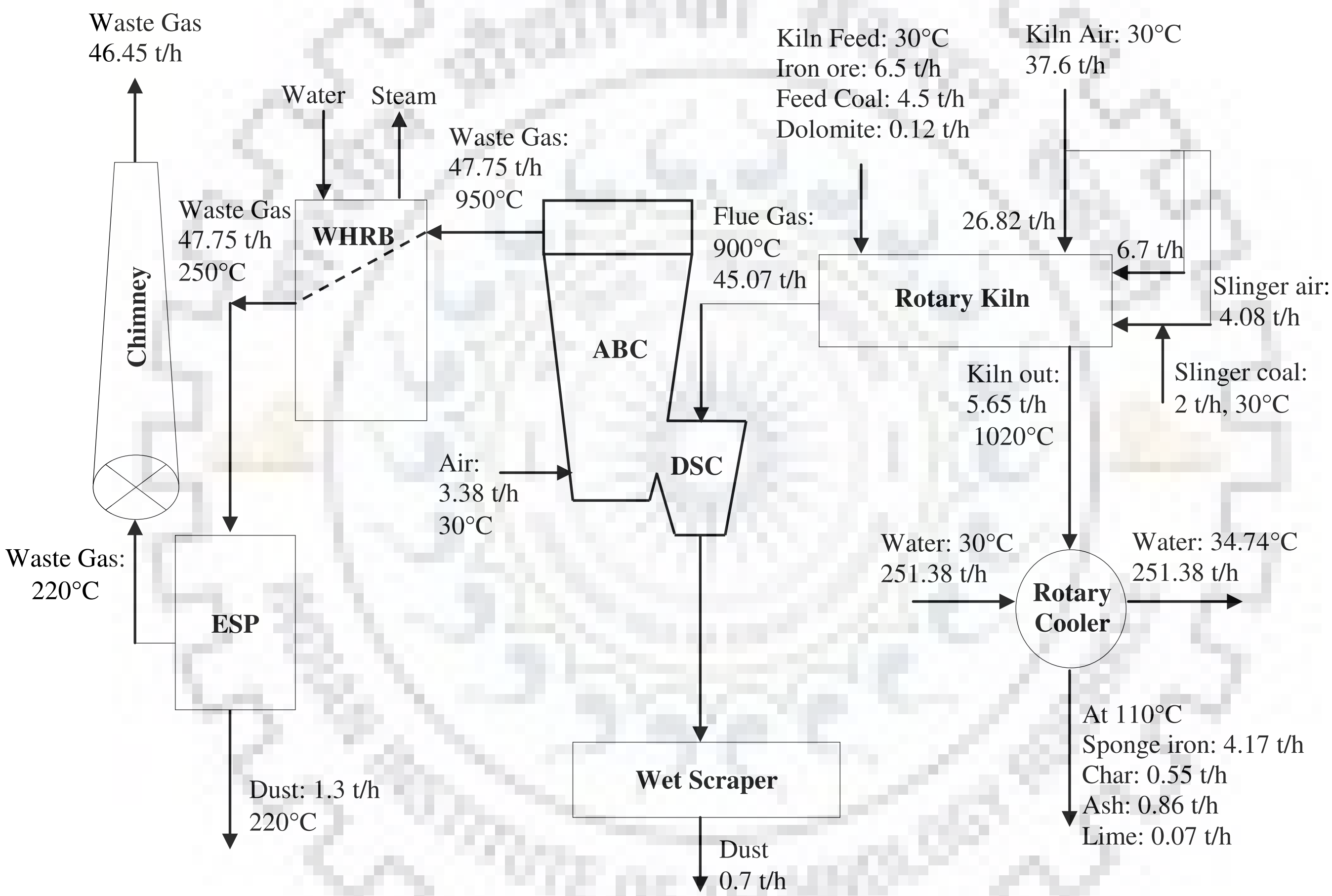


Fig. 4.3 Process flow diagram of Process-3

4.2 POTENTIAL AREAS TO RECOVER AND UTILIZE HEAT IN THE PROCESS

In coal based sponge iron processes, coal is the only source of heat and the required heat is generated when coal is burnt along with air in the rotary kiln. This heat is utilized for removing the moisture of iron ore, feed coal and slinger coal, preheating of feed up to desired temperature, metallization of iron ore, compensating radiation losses from kiln wall and decomposition of dolomite, etc. Hence, in the present process, coal and air can be considered as hot utility streams as far as process integration terminology is concerned. For the purpose of energy integration in coal based sponge iron industries one should know the potential areas in the process where heat is lost and can be recovered and where this heat can be utilized. Based on PFDs shown in Fig. 4.1 to Fig. 4.3, in sequence, these potential areas can be identified. PFD of Process-1 shown in Fig. 4.1 is reproduced as Fig. 4.4 with all possible potential areas for energy integration. Similar areas can also be found in Process-2 and Process-3. Fig. 4.4 shows total six potential areas as:

Area-1:

Waste gas exiting ESP goes to the chimney at 220^{O}C. From the chimney waste gas goes to the atmosphere, which is a loss of considerably high temperature heat.

Area-2:

Hot sponge iron is being cooled from 1020^{O}C to 110^{O}C in the rotary cooler (RC) while spraying cold water over the shell of cooler and vapor generated from it goes directly to atmosphere.

Area-3:

Waste gas coming out from ABC at very high temperature is cooled to around 250^{O}C using water in EC of Process-1 and Process-2. This is a loss of considerably high temperature heat. In Process-3 this heat is utilized for power generation.

Area-4:

Feed enters into kiln at ambient temperature, which is to be preheated up to reaction temperature i.e. 1020^{O}C before reduction process. This needs significant amount of heat to be generated from combustion of coal inside the kiln.

Area-5 and Area-6:

Similarly, slinger coal and kiln air, which enter at ambient temperature, are to be preheated up to 1020^{O}C before reactions to be started, thus, forming Area-5 and Area-6 respectively.

Thus, there are six potential areas where heat is lost and can be recovered and where this heat can be utilized [141]. These six areas are further divided into two areas, namely, heat recovering areas (heat sources) and heat utilizing areas (heat sources). Area-1, Area-2 and Area-3 are grouped as heat recovering areas (heat sources) whereas, remaining three areas, Area-4, Area-5 and Area-6, are considered as heat utilizing areas (heat sinks) where heat available in recovering areas can be transferred [141].

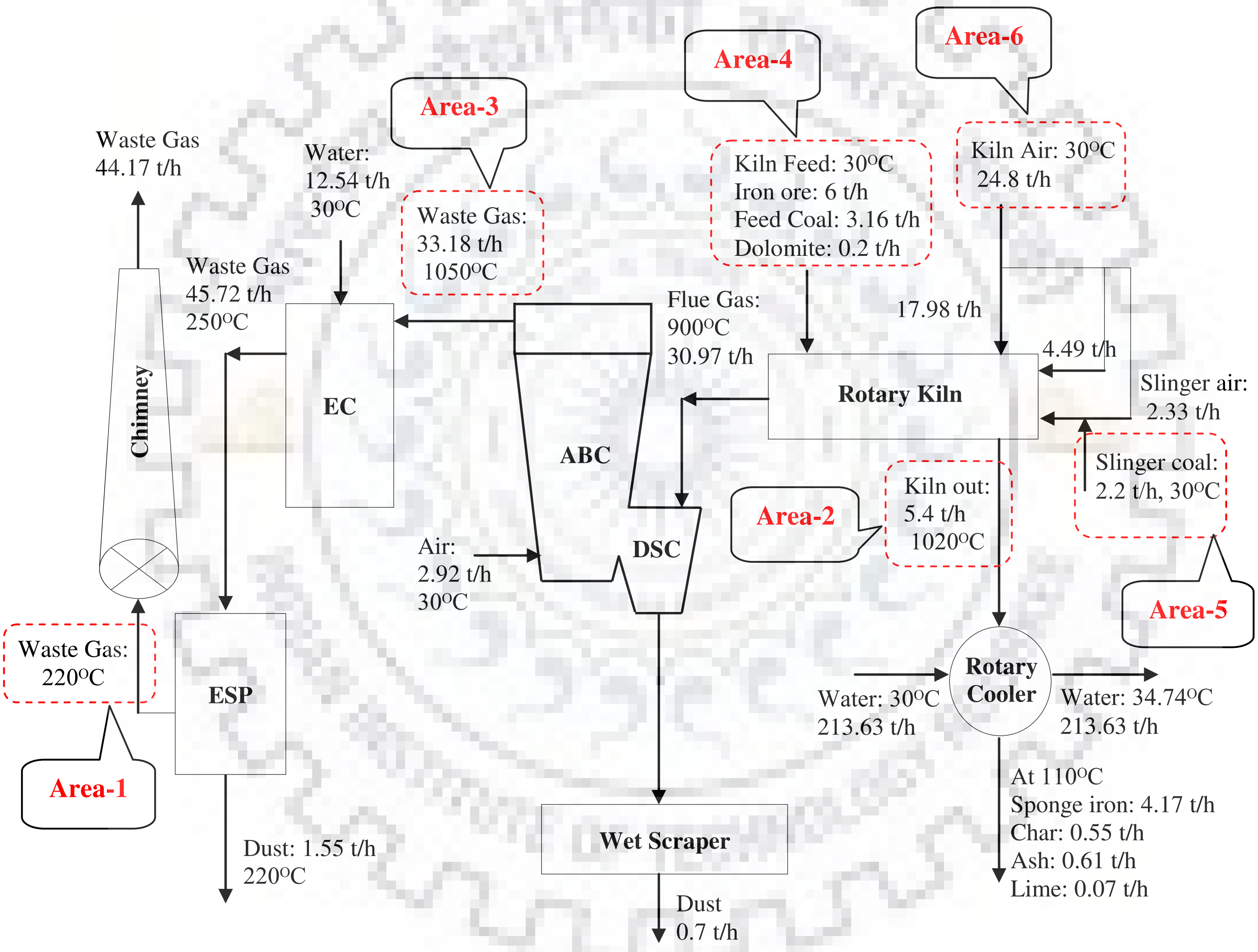


Fig. 4.4 Process flow diagram of Process-1 with potential areas for energy integration

4.3 FORMULATION OF SCHEMES FOR HEAT INTEGRATION

Two different design and operational modifications, Scheme-1 and Scheme-2, are proposed for heat integration in all three coal based sponge iron processes considering six potential areas. These schemes utilize heat of waste gas, which is otherwise wasted, in sponge iron plants. Scheme-1 and Scheme-2 are proposed based on the principles of process integration and applied in all three processes of coal based sponge iron cluster namely, Process-1, Process-2 and Process-3, individually. These schemes, proposed based on the guidelines given in Section 4.4, vary in terms of modification, investment and savings. Details of these two schemes are discussed in this section along with streams and potential areas considered for energy conservation.

4.3.1 Scheme-1

Scheme-1 is formulated while utilizing heat of Area-1 i.e. ESP exit gas. For Scheme-1 kiln feed (Area-4) and slinger coal (Area-5) are preheated from 30^{O}C to 120^{O}C using waste gas exiting ESP and kiln air (Area-6) is preheated using the remaining heat from ESP exit gas. Further, ESP exit gas (Area-1) is used to cool kiln outlet stream (Area-2) from 1020^{O}C to 110^{O}C. Thus, Scheme-1 covers all potential areas except Area-3. In the existing coal based sponge iron plants streams such as kiln feed, slinger coal and kiln air enter into the rotary kiln at ambient temperature and are preheated inside the kiln in preheating zone with the help of heat produced through combustion of coal. In the existing plants, waste gas leaves ESP at 220^{O}C, as shown in PFDs of individual processes, which is at significantly high temperature. If feed streams to the kiln are preheated outside the kiln before these enter into the kiln using heat available with ESP exit waste gas, a substantial amount of coal and water may be saved. Therefore, Scheme-1 includes solid streams such as iron ore and coal. The stream data of Scheme-1 for all three processes is shown in Table 4.1. Present scheme selects six streams, two hot and four cold streams as shown in Table 4.1, for integration in all three processes of total site.

The identified streams along with supply temperatures (T_s) and target temperatures (T_t), mass flow rates (m) and heat capacity flow rates (mC_p) are shown in Table 4.1 for these processes. In Scheme-1 waste gas exiting from ESP is cooled from 220^{O}C to 80^{O}C. In Procees-1, total 1831.13 kW heat is released from ESP exit gas out of which 333.1 kW is used to preheat kiln feed and slinger coal from 30^{O}C to 120^{O}C and remaining 1498.03 kW heat is used by kiln air to get preheated from 30^{O}C to 243.7^{O}C.

Table 4.1 Stream data for Scheme-1

Stream	Potential area	Stream Type	Stream no. used in HEN	T_s (^{O}C)	T_t (^{O}C)	Mass flow (t/h)	CP (kW/^{O}C)
Process-1							
Kiln outlet	Area-2	Hot	H1	1020	110	5.40	1.05
ESP exit gas	Area-1	Hot	H2	220	80	44.18	13.08
ESP exit gas	Area-1	Cold	C1	60	141.5	44.18	11.70
Kiln feed	Area-4	Cold	C2	30	120	9.36	2.87
Slinger coal	Area-5	Cold	C3	30	120	2.20	0.84
Kiln air	Area-6	Cold	C4	30	243.7	24.80	7.01
Process-2							
Kiln outlet	Area-2	Hot	H1	1020	110	4.14	0.76
ESP exit gas	Area-1	Hot	H2	220	80	61.05	1.07
ESP exit gas	Area-1	Cold	C1	60	109.45	61.05	0.95
Kiln feed	Area-4	Cold	C2	30	120	8.73	1.12
Slinger coal	Area-5	Cold	C3	30	120	2.60	1.38
Kiln air	Area-6	Cold	C4	30	263.78	33.23	1.02
Process-3							
Kiln outlet	Area-2	Hot	H1	1020	110	5.65	1.12
ESP exit gas	Area-1	Hot	H2	220	80	46.45	13.75
ESP exit gas	Area-1	Cold	C1	60	143.3	46.45	12.28
Kiln feed	Area-4	Cold	C2	30	120	11.12	3.49
Slinger coal	Area-5	Cold	C3	30	120	2.00	0.77
Kiln air	Area-6	Cold	C4	30	175.8	37.60	10.58

Similarly, in Procees-2 and Procees-3 kiln air is preheated to 263.8^{O}C and 175.8^{O}C respectively. Specific heat capacity of kiln outlet and kiln feed streams depend upon their composition. In the present work the property data of iron ore, coal, dolomite and sponge iron are constant and taken from literature [64,65] whereas properties of air and waste gas are calculated at respective average temperatures [176] as discussed in Section 3.2.2.

4.3.2 Scheme-2

Scheme-2 includes preheating of kiln feed (Area-4) and slinger coal (Area-5) from 30^{O}C to 300^{O}C using ABC exit gas (Area-3). Kiln outlet stream (Area-2) is cooled from 1020^{O}C to 110^{O}C using kiln air (Area-6) and further kiln air is preheated to 300^{O}C using ABC exit gas. In the existing industrial cluster, waste gas that leaves ABC is at 1050^{O}C, 850^{O}C and 950^{O}C for Process-1, Process-2 and Process-3, respectively. Scheme-2 is formulated while utilizing heat of this ABC exit gas (Area-3), which is at significantly high temperature. Thus Scheme-2 covers all potential areas except area-1. Five streams, two hot and three cold, are considered for the integration in Process-1, Process-2 and Process-3. The required stream data for these processes are shown in Table 4.2.

Table 4.2 Stream data for Scheme-2

Stream	Potential area	Stream Type	Stream no. used in HEN	T_s (^oC)	T_t (^oC)	Mass flow (t/h)	CP (kW/^oC)
Process-1							
Kiln outlet	Area-2	Hot	H1	1020	110	5.40	1.05
ABC exit gas	Area-3	Hot	H2	1050	870	33.18	10.82
Kiln feed	Area-4	Cold	C1	30	300	9.36	2.87
Slinger coal	Area-5	Cold	C2	30	300	2.20	0.84
Kiln air	Area-6	Cold	C3	30	300	24.80	7.03
Process-2							
Kiln outlet	Area-2	Hot	H1	1020	110	4.14	0.87
ABC exit gas	Area-3	Hot	H2	850	672.3	44.59	15.47
Kiln feed	Area-4	Cold	C1	30	300	8.73	2.71
Slinger coal	Area-5	Cold	C2	30	300	2.60	1.00
Kiln air	Area-6	Cold	C3	30	300	33.23	9.42
Process-3							
Kiln outlet	Area-2	Hot	H1	1020	110	5.65	1.12
ABC exit gas	Area-3	Hot	H2	950	763.5	47.75	16.12
Kiln feed	Area-4	Cold	C1	30	300	11.12	3.49
Slinger coal	Area-5	Cold	C2	30	300	2.00	0.77
Kiln air	Area-6	Cold	C3	30	300	37.60	10.66

4.4 FORMULATION OF STRATEGIES FOR TOTAL SITE INTEGRATION

Along with Scheme-1 and Scheme-2, two more design and operational modifications, Strategy-1 and Strategy-2, are proposed for total site integration of sponge iron cluster shown in Fig. 4.5. These strategies are proposed based on following guidelines to conserve energy in total site of plants of similar type where conventional methods are not applicable:

(a) Based on PFDs of plants recovering areas should be identified where heat is lost. In the similar line, utilizing areas should also be found in plants where heat available in recovering areas can be transferred.

(b) Using recovering areas different hot streams are identified whereas, cold streams are found based on utilizing areas. Based on this guideline hot and cold streams are also identified for Scheme-1 and Scheme-2.

(c) Based on combination of recovering (hot streams) and utilizing (cold streams) areas a strategy can be defined. For maximum heat recovery one should consider all possible recovering and utilizing areas while defining a strategy.

(d) For defined strategy all associated hot and cold streams of plants are collected with its supply and target temperatures, mass flow rates and specific heat capacities, etc. and named these as stream data. This guideline is also followed to generate stream data of Scheme-1 and Scheme-2.

(e) For a strategy, %heat recovery can be preliminarily determined using stream data to find the efficacy of the strategy.

These guidelines are also followed to identify Scheme-1 and Scheme-2.

In the present work, total six recovering as well utilizing areas, Area-1 to Area-6, are identified as discussed in Section 4.2. In both strategies all three processes, Process-1, Process-2 and Process-3, are considered simultaneously as single process where heat integration can be taken place between different processes. The proposed strategies should be able to recover maximum amount of energy from these areas. Considering this aspect, Strategy-1 is proposed in the present work. However, Strategy-2 is identified to carry out comparative analysis only. These strategies vary in terms of modification, investments and savings. Details of these two strategies are discussed in this section along with streams and potential areas considered for energy conservation.

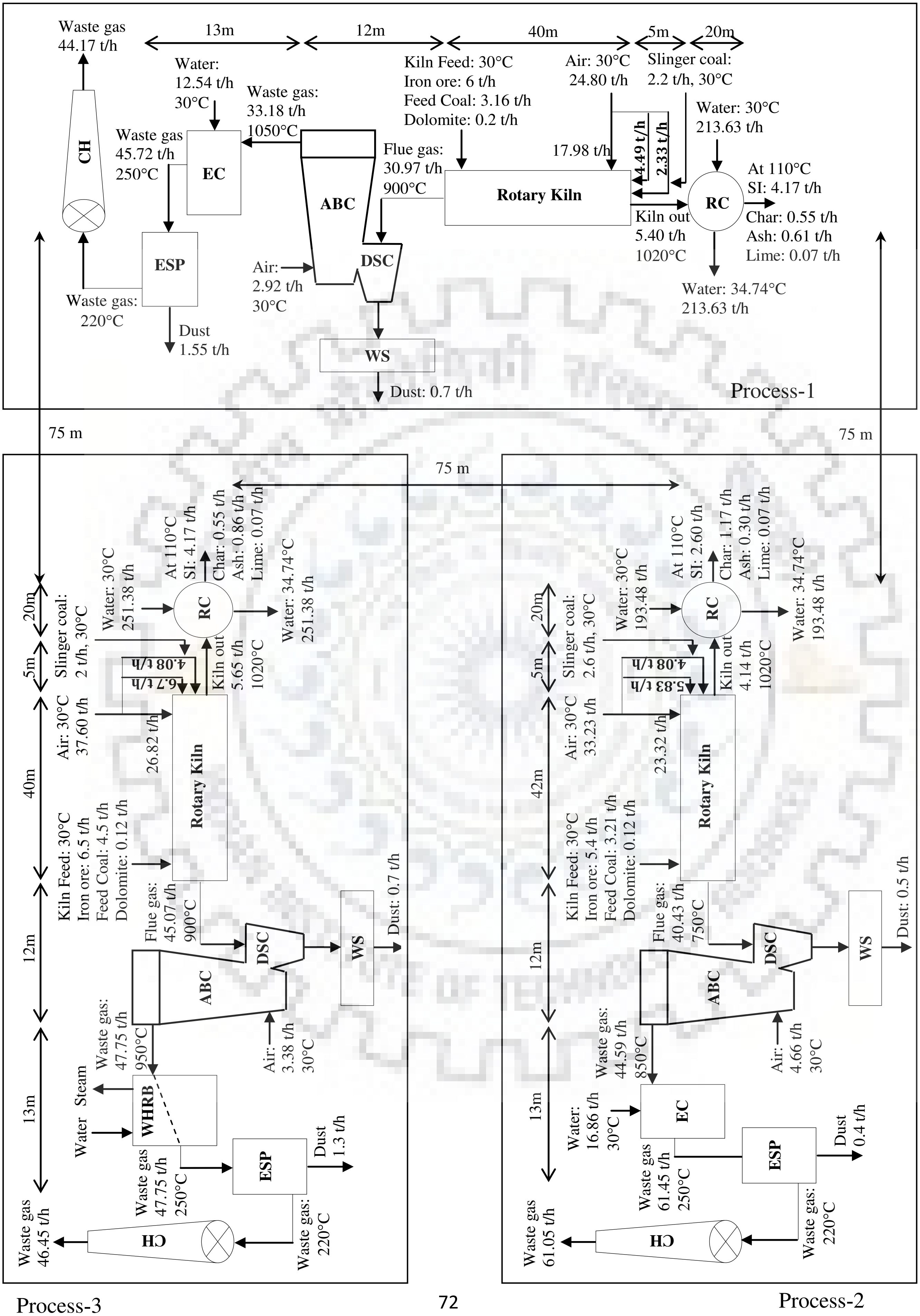


Fig. 4.5 Total Site layout of Sponge Iron Cluster

4.4.1 Strategy-1

Based on the guidelines given in Section 4.4 Strategy-1 is proposed. In Strategy-1, waste gas exits from ABC of Process-1 splits into three branches and is used for preheating the kiln feed and slinger coal of all three processes, individually, from 30^{O}C to 300^{O}C. Kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air of its respective process. Further, kiln air is preheated using waste gas to maximum possible temperature (421^{O}C in this case) such that ΔT_{min} equal to 50^{O}C is maintained. After preheating kiln feed and slinger coal of all three processes three branches of the waste gas are again mixed to enter into ESP of Process-1. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation. Once power is generated waste gas stream further splits into two parts. One part enters into ESP of Process-2 whereas other into ESP of Process-3.

The stream data of Strategy-1, which consists of 5 hot and 9 cold streams, is shown in Table 4.3. In fact, this table includes stream data of all three processes together and thus, all three processes are considered as single process in Strategy-1. The number shown as a subscript in stream data of Table 4.3 denotes the specific process in the total site. For example, kiln outlet of Process-1 is denoted as Kiln outlet$_1$ and kiln outlet of Process-2 and Process-3 are denoted as Kiln outlet$_2$ and Kiln outlet$_3$, respectively. On the other hand, combined ABC exit waste gas streams of Process-2 and Process-3 is represented as ABC exit gas$_{23}$. Similarly, different streams are denoted as shown in Table 4.3. In Strategy-1 total 13850.4 kW heat is required to preheat kiln inlet streams i.e. kiln feed, slinger coal and kiln air of all plants, to target temperatures as can be seen through Table 4.3. Here, 2,770.6 kW heat is provided by cooling kiln outlet streams from 1020^{O}C to 110^{O}C and remaining 11,079.8 kW is supplied through ABC exit gas$_1$ stream. Consequently, target temperature of ABC exit gas$_1$ stream is found as 43.2^{O}C as shown in Table 4.3. In the similar line target temperatures of streams are found and reported in Table 4.3. Thus, Strategy-1 includes all three recovering and utilizing areas as shown in Table 4.3.

Table 4.3 Stream data for Strategy-1

Stream	Potential area	Stream Type	Stream no. used in HEN	T_s (°C)	T_t (°C)	Mass flow (t/h)	CP (kW/°C)
Kiln outlet$_1$	2	Hot	H1	1020	110	5.40	1.05
Kiln outlet$_2$	2	Hot	H2	1020	110	4.14	0.87
Kiln outlet$_3$	2	Hot	H3	1020	110	5.65	1.12
ABC exit gas$_{23}$	1 and 3	Hot	H4	901	120	92.34	30.24
ABC exit gas$_1$	1 and 3	Hot	H5	1050	43.2	33.18	11.01
Kiln air$_1$	6	Cold	C1	30	421	24.80	7.10
Kiln air$_2$	6	Cold	C2	30	421	33.23	9.51
Kiln air$_3$	6	Cold	C3	30	421	37.60	10.76
Slinger coal$_1$	5	Cold	C4	30	300	2.20	0.84
Slinger coal$_2$	5	Cold	C5	30	300	2.60	1.00
Slinger coal$_3$	5	Cold	C6	30	300	2.00	0.77
Kin feed$_1$	4	Cold	C7	30	300	9.36	2.87
Kin feed$_2$	4	Cold	C8	30	300	8.73	2.71
Kin feed$_3$	4	Cold	C9	30	300	11.12	3.49

4.4.2 Strategy-2

Under Strategy-2 waste gas streams exiting ESPs of Process-1, Process-2 and Process-3 are mixed. This mixed stream is used as hot stream to preheat kiln feed and slinger coal of all the three processes. For this purpose total kiln feed, used in all three processes, is preheated at Process-1 using mixed stream of waste gases and same stream of waste gases is also used to preheat total slinger coal of three processes at Process-1 only. Then the preheated kiln feed and slinger coal is carried from Process-1 to their respective processes. Further, kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air of its respective process. After exchanging heat of waste gas with kiln feed and slinger coal of total site, the mixed waste gas stream is again splits into three different streams to move to chimney of three processes. It is noted from PFD of total site, shown in Fig.4.5, that, amongst these three processes only Process-3 has a unit of waste heat recovery boiler (WHRB) that recovers heat from waste gas exiting ABC and rest of the processes cool the waste gas exiting ABC using water in EC. Thus, Strategy-2 proposes WHRBs, for power generation, in Process-1 and Process-2 using waste gas exiting from ABC of their respective processes.

The stream data of Strategy-2, which includes 6 hot and 5 cold streams, is shown in Table 4.4. This Strategy, similar to Strategy-1, also considers stream data of all three processes together as single process. The number shown as a subscript in stream data of Table 4.4 denotes the specific process in the total site. For example, kiln outlet of Process-1 is denoted as Kiln outlet$_{1}$. The combined ESP exit waste gas streams of three processes is denoted as ESP exit gas$_{123}$ while Kiln feed$_{123}$ and Slinger coal$_{123}$ are used as nomenclature for kiln feed and slinger coal required for total site. Other streams, used in Strategy-2, are denoted as similar to Strategy-1 and shown in Table 4.4. In Strategy-2, ESP exit gas$_{123}$ is cooled from 220^{O}C to 189.2^{O}C, as shown in Table 4.4, releasing 1400.7 kW of heat for preheating Kiln feed$_{123}$ and Slinger coal$_{123}$ from 30^{O}C to 150^{O}C. Kiln outlet$_{1}$ stream is cooled from 1020^{O}C to 110^{O}C releasing 953.1 kW heat to preheat Kiln air$_{1}$ from 30^{O}C to 166.6^{O}C. Similarly, target temperatures of other streams are computed and reported in Table 4.4. Thus, Strategy-2 considers all three recovering and utilizing areas as shown in Table 4.4.

Table 4.4 Stream data for Strategy-2

Stream	Potential area	Stream Type	Stream no. used in HEN	T_s (^oC)	T_t (^oC)	Mass flow (t/h)	CP (kW/^oC)
Kiln $outlet_1$	2	Hot	H1	1020	110	5.40	1.05
Kiln $outlet_2$	2	Hot	H2	1020	110	4.14	0.87
Kiln $outlet_3$	2	Hot	H3	1020	110	5.65	1.12
ESP exit gas_{123}	1	Hot	H4	220	189.2	151.67	45.54
ABC exit gas_1	3	Hot	H5	1050	250	33.18	11.27
ABC exit gas_2	3	Hot	H6	850	240	44.59	14.79
Kiln $feed_{123}$	4	Cold	C1	30	150	29.21	9.07
Slinger $coal_{123}$	5	Cold	C2	30	150	6.80	2.61
Kiln air_1	6	Cold	C3	30	166.6	24.80	6.98
Kiln air_2	6	Cold	C4	30	115.2	33.23	9.33
Kiln air_3	6	Cold	C5	30	126.8	37.60	10.56

Both Scheme-1 and Scheme-2, for energy integration, are correlated to the key features of three processes and summarized in Table 4.5. Similarly Strategy-1 and Strategy-2, for total site integration, are also correlated to the key features of total site and summarized in Table 4.6.

4.4.3 Distances between various equipment in total site

Total site PFD of three coal based sponge iron processes is shown in Fig. 4.5. These plants are located near to each other with distances mentioned in Fig. 4.5. Distance between different plants is around 75 m i.e. between Process-1 and Process-2, Process-2 and Process-3, Process-3 and Process-1. Distance between different equipment employed in Process-1, Process-2 and Process-3 are shown in total site PFD (Fig. 4.5). Distances between equipment among different processes are shown in Table 4.7. Distance between Kiln inlet of Process-1 to ESP exit of Process-2 is 140 m whereas it is 79 m from Kiln inlet of Process-1 to ESP exit of Process-3. Similarly, other important distances that are used in calculation are shown in Table 4.7. The number shown as a subscript in

Table 4.7 denotes the specific process in the total site, which is similar to stream data of Strategy-1 and Strategy-2 shown in Table 4.3 and Table 4.4, respectively.

Table 4.5 Key features of Scheme-1 and Scheme-2

<table>
<tr><th rowspan="2">Scheme</th><th rowspan="2">Scheme details</th><th colspan="2">Potential areas covered</th></tr>
<tr><th>Heat sources</th><th>Heat sinks</th></tr>
<tr><td rowspan="4">Scheme-1</td><td>ESP exit gas (from 220^{o}C to 80^{o}C) → Kiln feed (from 30^{o}C to 120^{o}C)</td><td rowspan="4">1 and 2</td><td rowspan="4">4, 5 and 6</td></tr>
<tr><td>ESP exit gas (from 220^{o}C to 80^{o}C) → Slinger coal (from 30^{o}C to 120^{o}C)</td></tr>
<tr><td>ESP exit gas (from 220^{o}C to 80^{o}C) → Kiln air (from 30^{o}C to maximum possible temp.)</td></tr>
<tr><td>Kiln outlet (from 1020^{o}C to 110^{o}C) → ESP exit gas (from 60^{o}C to maximum possible temp.)</td></tr>
<tr><td rowspan="4">Scheme-2</td><td>Kiln outlet (from 1020^{o}C to 110^{o}C) → Kiln air (from 30^{o}C to maximum possible temp.)</td><td rowspan="4">2 and 3</td><td rowspan="4">4, 5 and 6</td></tr>
<tr><td>ABC exit gas → Kiln feed (from 30^{o}C to 300^{o}C)</td></tr>
<tr><td>ABC exit gas → Slinger coal (from 30^{o}C to 300^{o}C)</td></tr>
<tr><td>ABC exit gas → Kiln air (up to 300^{o}C)</td></tr>
</table>

Table 4.6 Key features of Strategy-1 and Strategy-2

Strategy	Scheme details	Potential areas covered	
		Heat sources	Heat sinks
Strategy -1	ABC exit gas_1 (from 1050^oC to 43.2^oC) → Kiln $feed_{123}$ (from 30^oC to 300^oC)	1,2 and 3	4, 5 and 6
	ABC exit gas_1 (from 1050^oC to 43.2^oC) → Slinger $coal_{123}$ (from 30^oC to 300^oC)		
	ABC exit gas_1 (from 1050^oC to 43.2^oC) → Kiln air_1 (from 30^oC to 421^oC)		
	ABC exit gas_1 (from 1050^oC to 43.2^oC) → Kiln air_2 (from 30^oC to 421^oC)		
	ABC exit gas_1 (from 1050^oC to 43.2^oC) → Kiln air_3 (from 30^oC to 421^oC)		
	ESP exit gas (from 220^oC to 80^oC) → Kiln air (from 30^oC to maximum possible temp.)		
	ABC exit gas_{23} (from 901^oC to 120^oC) → Power generation		
Strategy -2	ESP exit gas_{123} (from 220^oC to 189.2^oC) → Kiln $feed_{123}$ (from 30^oC to 150^oC)	1, 2 and 3	4, 5 and 6
	ESP exit gas_{123} (from 220^oC to 189.2^oC) → Slinger $coal_{123}$ (from 30^oC to 150^oC)		
	Kiln $outlet_1$ (from 1020^oC to 110^oC) → Kiln air_1 (from 30^oC to maximum possible temp.)		
	Kiln $outlet_2$ (from 1020^oC to 110^oC) → Kiln air_2 (from 30^oC to maximum possible temp.)		
	Kiln $outlet_3$ (from 1020^oC to 110^oC) → Kiln air_3 (from 30^oC to maximum possible temp.)		
	ABC exit gas_1(from 1050^oC to 250^oC) → Power generation		
	ABC exit gas_2 (from 850^oC to 240^oC) → Power generation		

Table 4.7 Distance between different equipment among different plants

From	To	Distance (m)
Kiln $inlet_1$	ESP $exit_2$	140
Kiln $inlet_1$	ESP $exit_3$	79
ABC $exit_1$	Kiln $outlet_2$	165
ABC $exit_1$	Kiln $outlet_3$	93
Kiln $outlet_1$	ESP $exit_3$	100
Kiln $inlet_2$	Kiln $inlet_3$	156
Kiln $outlet_2$	Kiln $outlet_3$	103
ABC $exit_2$	Kiln $outlet_1$	101
ABC $exit_2$	Kiln $outlet_3$	126
Kiln $inlet_3$	ESP $exit_1$	79
Kiln $inlet_3$	ESP $exit_2$	168
ABC $exit_3$	Kiln $inlet_1$	76
ABC $exit_3$	Kiln $inlet_2$	167

4.4.4 Designing of HEN for total site integration

In coal based sponge iron plants coal is used as utility and it is also one of the feed stream. In this case usual pinch design rules are not applicable in designing HEN for proposed design modifications. Thus, following guidelines are proposed for such cases, which are based on actual practice:

1. All streams should be considered as part of a single plant

2. ΔT_{min} should not be violated for any match

3. Solid streams should exchange heat with gas and liquid streams but not with solid.

4. If options are available, exchanger should be placed between near most streams.

Based on these guidelines HEN for total site can be designed.

4.4.5 Modification of total site PFD

As distances between plants are larger than that within a plant, connecting various streams of different plants for interchanging heat is more tedious in total site. Thus, in modification of total site PFD, following guidelines are proposed for movement of streams within total site:

1. Liquid stream can be transported through pipes of different diameter. Pressure drop in each pipe should be computed to propose the required number of pumps.

2. Gas stream can be transported through ducts where pressure drop in each duct is to be found to propose number of forced draft (FD) fans.

3. Solid stream should be transported through conveyors either open or closed.

4. For exchange of heat between gas/liquid and solid, preference should be given to transport gas and liquid streams over solid.

Based on these guidelines the modified PFD of total site can be designed.

4.5 MODEL DEVELOPMENT FOR COAL CONSUMPTION

In coal based sponge iron plants, coal, as the only source of energy, is combusted to provide the energy required for the reduction of iron ore to sponge iron. Coal consumption is decided by the energy demand of the process, which depends on heat requirement of incoming air and feed materials for preheating, heat involved in the reactions of reduction process, heat lost through the rotary kiln wall, preheating required by coal itself and latent heat required for evaporation of moisture of feed material. Based on these factors heat requirements inside the kiln were computed by Prasad et al. [131]. However, author did not consider heat required to preheat dolomite that enters the kiln along with iron ore and feed coal, heat released in its decomposition and also heat required to vaporize the coal volatiles. These factors, although minor, not only change the amount of coal consumption but also amount of waste gas released from sponge iron plants. Considering all these factors a modified model is developed for computation of coal utility factor and coal consumption as:

Hot utility requirement (Q_{hu})

It is found using pinch analysis [96] for the given stream data. This is supplied by combustion of coal as it is the only source of energy. HINT software is used, in the present work, to compute the hot utility. In this process coal is the only source of energy, which is provided when coal is burnt along with air.

Process heat (Q_p)

The heat of reaction is computed for all combustion as well as reduction reactions using heat balance of the process on hour basis and results are shown in Appendix A. Process heat is the heat required for the reduction reactions to continue. The following reduction reactions take place inside the kiln:

There are two reduction reactions. In first reaction iron oxide proceeds only to ferrous oxide Whereas in second reaction ferrous oxide reacts with CO to produce metallic iron (Fe) which is also called as sponge iron. Both reactions are exothermic in nature.

$$Fe_2O_3 + CO = 2FeO + CO_2$$

$$FeO + CO = 2Fe + CO_2$$

Sensible heat (Q_{ore}, Q_a, Q_c and Q_d)

The iron ore, air, dolomite and coal itself are to be brought to the reaction temperature conditions inside the rotary kiln. Constant specific heat of each stream is considered to predict the heat required for preheating these streams. The expressions are:

$$Q_{ore} = m_{ore} Cp_{ore} \left(T_p - T_i\right) \quad (4.1)$$

$$Q_a = m_a Cp_a \left(T_p - T_i\right) \quad (4.2)$$

$$Q_c = m_c Cp_c \left(T_p - T_i\right) \quad (4.3)$$

$$Q_d = m_d Cp_d \left(T_p - T_i\right) \quad (4.4)$$

Rotary kiln losses (Q_{loss})

There is a heat loss from surface of rotary kiln, which can be quantized by the product of surface area of the kiln and loss coefficient. In this process there are also losses at the inlet and outlet of the kiln, at post combustion chamber and at the inlet area of the cooler. Jena et al. [77] and Biswas et al. [17] have discussed about these losses and suggested the total loss to be twice of that caused by surface of rotary kiln. Thus,

$$Q_{loss} = (2\pi DL)h_r \quad (4.5)$$

Heat required for removing moisture of feed material (Q_m)

Coal and iron ore that are used in the process can absorb some moisture depending upon ambient conditions. Since these are not treated before entering the kiln, the heat required to evaporate this moisture from these is also supplied by the combustion of coal. In Indian conditions iron ore absorbs around 2% moisture [2] and average moisture content in coal by weight is taken from the composition of feed coal and slinger coal used in the process. Average moisture content of coal used in Process-1 is 3.52% whereas in Process-2 and Process-3 it is 3.55% and 3.46%, respectively.

$$Q_m = (0.0352 m_c + 0.02 m_s)\lambda; for\ Process - 1 \quad (4.6)$$

$$Q_m = (0.0355 m_c + 0.02 m_s)\lambda; for\ Process - 2 \quad (4.7)$$

$$Q_m = (0.0346 m_c + 0.02 m_s)\lambda; for\ Process - 3 \quad (4.8)$$

Heat of dolomite decomposition (Q_{de})

Dolomite decomposes into oxides inside the rotary kiln and heat is released during this process. This heat of decomposition is found using heat balance of the decomposition process on hour basis as shown in Appendix A.

Heat of vaporization of coal volatile matter ($Q_{v\text{-}Coal}$ and $Q_{v\text{-}Char}$)

Heat is required to vaporize coal volatiles inside the rotary kiln. Amount of volatiles vaporized is computed after subtracting volatiles present in the char coming out with sponge iron at the kiln outlet, from the total volatiles present in the coal. Considering the heat of coal devolatization to be 122 kCal/kg [77] $Q_{v\text{-}Coal}$ and $Q_{v\text{-}Char}$ are calculated on hour basis.

Depending upon the plant equipment conditions and coal used only a fraction of total energy, released from coal combustion, is used inside the rotary kiln. Defining this fraction as coal utility factor (U_f) and considering the Net heating Value (NHV) of coal to be 22,930.5 kJ/kg [77] the energy balance is carried out around the rotary kiln. The final expression for utility factor is as follows.

$$Q_{hu} + Q_p + Q_{ore} + Q_a + Q_c + Q_d + Q_{loss} + Q_m + Q_{de} + Q_{v-coal} + Q_{v-char} = m_c(NHV * U_f) \tag{4.9}$$

Once coal utility factor (U_f) is predicted from Equation 4.9, for each process at the existing conditions, the same Equation 4.9 can be used to find coal consumption for Scheme-1, Scheme-2, Strategy-1 and Startegy-2. In design calculations for single plant integration, through Scheme-1 and Scheme-2, existing kiln air to coal ratio of 4.63, 5.72 and 5.79 are used for Process-1, Process-2 and Process-3, respectively. For calculations of total site integration all parameters of individual processes should be combined together and consider for computing U_f from Equation 4.9. For total site kiln air to coal ratio is predicted as 5.41, which is to be used for further calculations in total site integration.

CHAPTER 5

DEVELOPMENT OF SOLUTION METHODOLOGY

In the present work four design and operational modifications are proposed for energy conservation in all three sponge iron processes. First two design modifications namely Scheme-1 and Scheme-2 are proposed based on the principles of process integration and applied in individual processes. Remaining two modifications, Strategy-1 and Strategy-2, are suggested for total site integration. This chapter deals with the development of methodology to solve these proposed design modifications. This includes calculating amount of coal and air required by the processes while using these modifications and also carrying out economic analysis to select best possible design modification to existing sponge iron processes. These schemes and strategies for energy conservation in coal based sponge iron cluster along with the model for computation of coal consumption are formulated in Chapter 4. The process flow diagrams (PFDs) of all three processes of sponge iron cluster along with operating conditions are also given in Chapter 4.

5.1 STEP WISE PROCEDURE FOR ENERGY INTEGRATION

The energy integration schemes for sponge iron processes are solved using HINT software and Microsoft Excel-2007. Detailed solution methodology is covered in the following steps:

1. Data from sponge iron processes including respective PFDs are collected from the industrial site.
2. The physical properties of iron ore, coal, dolomite, air, sponge iron and waste gas streams are collected from different sources. These are detailed in Section 3.2.2.
3. Material and energy balances are performed for sponge iron processes which are detailed under Section 4.1.
4. Air requirement based on oxygen demand of the process is computed and amount of coal is computed using Equation 4.9 developed in Section 4.5. Air to coal ratio is computed based on plant existing conditions.
5. PFD of sponge iron processes are prepared based on material and energy balanced data and various energy conservation schemes are defined. These are detailed under Section 4.3.

6. According to scheme, defined in step-5, stream data for hot and cold streams are extracted and stream tables are prepared. These are detailed under Section 4.3.

7. Minimum hot and cold utility requirements are targeted for each scheme using pinch analysis. Coal and there by air requirements are computed.

8. Using the revised values of coal and air, computed in Step-7, stream table is prepared again and hot and cold utility requirements are computed using pinch analysis.

9. Step-7 and step-8 are repeated until utilities remain same.

10. For final values of hot utility HEN is designed and modified PFD is prepared.

11. Economic analysis is carried out for the modified PFD.

5.2 STEP WISE PROCEDURE FOR MODIFIED TOTAL SITE INTEGRATION

As discussed in Section 4.4, two different design and operational modifications, Strategy-1 and Strategy-2, are proposed for total site integration of sponge iron cluster. These are based on modified and generalised approach for total site integration of industrial clusters, where plants of similar process are placed. It involves very less steps as compared to conventional total site integration approach. The modified approach for total site of similar plants can be better understood while comparing it with that proposed for general problems of process integration. Conventionally TS Integration is carried out through following steps [53,80,162,165]:

Step-1: Data extraction of all individual processes in TS

Step-2: Preparation of stream tables and selection of ΔT_{min}

Step-3: Energy targeting using PTA for individual processes

Step-4: Plotting GCC for each process and remove pockets

Step-5: Construction of TS profiles

Step-6: Selection of utility data and matching utility generation and consumption targets

Step-7: Plotting Site Utility Grand Composite Curve

Step-8: Targeting shaft work and cogeneration

Step-9: Designing heat exchanger network (HEN) of TS

Step-10: Modification of TS process flow diagram (PFD)

However, steps involved in the proposed approach are [142]:

Step-1: Define a strategy for energy conservation

Guidelines to define a strategy are detailed under Section 4.4.

Step-2: Data extraction for all processes together in total site according to the strategy

Guidelines for data extraction are detailed under Section 4.4.

Step-3: Preparation of stream table considering all processes together and selection of ΔT_{min}

Guidelines for preparation of stream data are detailed under Section 4.4. Suitable ΔT_{min} for pinch analysis of stream data is selected. If heat transfer between gas and solid streams is involved large ΔT_{min} should be chosen, say $40^{O}C$-$50^{O}C$, to compensate lower heat transfer coefficient that needs large heat transfer area.

Step-4: Utility targeting of total site

Target minimum hot and cold utility requirements for each strategy using pinch analysis. Determine coal requirement using Equation 4.9, developed in Section 4.5.

Step-5: Designing HEN of total site

HEN of total site is designed following the guidelines proposed in Section 4.4.4.

Step-6: Modification of total site PFD

Modified PFD of total site is prepared following the guidelines proposed in Section 4.4.5. Detailed economic analysis is carried out for the modified PFD of total site to select best strategy.

Thus, the modified approach involves only 6 steps, which are very few as compared to conventional method of total site integration. The detailed computation methodology for the modified approach is presented in Fig. 5.1 [142]. In sponge iron plants coal is used as utility and it is also one of the feed stream. But, in some industrial sites, utility stream may or may not be one of the process stream. The flow chart, in Fig. 5.1, is applicable for both cases where utility stream such as coal, coke, char, etc. may or may not be one of the process stream. For a process, if utility stream is one of the feed, stream data for proposed strategy vary when utility consumption changes. Hence, an iterative approach should be employed, as shown in Fig. 5.1.

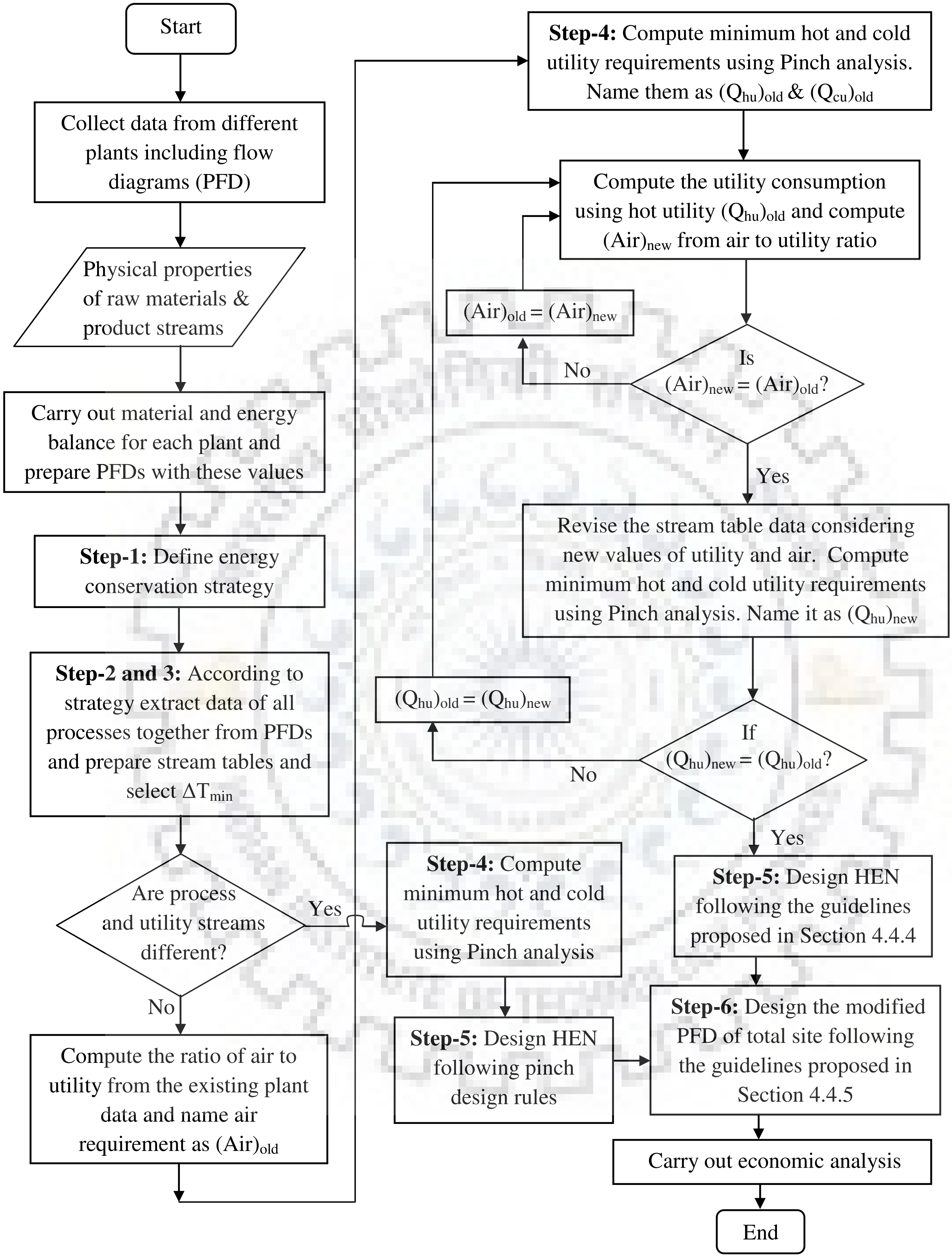


Fig. 5.1 Flow chart of the solution methodology adopted for modified total site integration

The model for computation of coal consumption that is developed under Section 4.5 is used in both energy integration and total site integration. The final expression is given as,

$$Q_{hu} + Q_p + Q_{ore} + Q_a + Q_c + Q_d + Q_{loss} + Q_m + Q_{de} + Q_{v-coal} + Q_{v-char} = m_c(NHV * U_f) \tag{4.9}$$

Here,

Q_{hu} is hot utility which is a dependent variable

Q_c and Q_a are dependent variables whereas m_c (mass flow rate of coal) is design parameter.

And all other parameters are independent variables.

5.3 STEP WISE PROCEDURE FOR POWER GENERATION

In energy conservation through Strategy-1 and Strategy-2 power generation is carried out using waste gases exiting ABC. Steps 1 to 5, shown in Section 5.2 for modified approach, should be carried out prior to power generation and thereafter following steps are to be used.

1. First, Grand Composite Curve (GCC) is drawn using HINT software considering ΔT_{min} as $50^{o}C$. GCC shows total heat of waste gas that can be used for energy integration.
2. Then draw steam generation profile in the GCC plot showing the extent of preheating, boiling and superheating on shifted temperature scale.
3. Total power generation is computed, which depends on saturated temperature, superheated temperature, superheated pressure and wetness fraction. These parameters are varied to find maximum power generation.
4. Power generation cycle is designed with all optimized operating parameters, which consists of super heater, boiler, preheater, turbine and condenser.
5. Water consumption is computed for the modified system i.e. sponge iron process with power generation cycle.

Once these steps are followed then Step 6, shown in Section 5.2 for modified approach, is carried out.

In power generation model which is used for computing power generation, the parameters such as saturated temperature, superheated temperature, superheated pressure and wetness fraction are varied to find maximum power generation.

Hence, saturated temperature, superheated temperature, superheated pressure and wetness fraction are independent variables whereas amount of power generated is a design parameter which is dependent variable.

CHAPTER 6

RESULTS AND DISCUSSION

This chapter demonstrates results predicted from theoretical studies carried out in the present work for heat and total site integration of coal based sponge iron plants. For this purpose, actual data of three sponge iron industries such as Process-1, Process-2 and Process-3 being operated with different capacities in an industrial cluster of India are considered. All these three processes are SL/RN based (Stelco-Lurgi / Republic Steel-National Lead). The process flow diagrams of sponge iron plants with operating data are shown in Chapter-4 whereas detailed process is described in Section 3.1. In each plant potential areas are identified, as discussed in Section 4.2, where heat is lost and can be conserved and also where heat is required. Based on the guidelines discussed in Section 4.4, two different schemes for heat integration, namely Scheme-1 and Scheme-2, are proposed. Further, two strategies, Strategy-1 and Strategy-2, are suggested for total site integration based on the modified approach discussed in Section 5.2. Details of two schemes and two strategies are discussed in Sections 4.3 and 4.4, respectively. These schemes and strategies vary in terms of modification, savings and investments. For heat and total site integration pinch analysis [161] is used. The complete computational methodology considered for solution of schemes and strategies along with model to estimate coal consumption are developed in Chapter-5. Selections of best scheme for heat integration and strategy for total site integration are carried out based on coal consumption, water requirement, total annual cost (TAC), net profit value (NPV), internal rate of return (IRR) and discounted payback period (DPP), waste gas generation, %heat recovery, etc. The process information like physical properties of the raw materials and cost data are shown in Section 3.2. The detailed material and energy balance around sponge iron plants, power generation and economic analysis are shown in Appendix.

6.1 ANALYSIS OF HEAT INTEGRATION OF INDIVIDUAL PROCESSES

Two different schemes, Scheme-1 and Scheme-2, are proposed for heat integration in all three coal based sponge iron processes. These are discussed and presented in Section 4.3. Best scheme, amongst these, is selected for each process.

6.1.1 Process-1 and Scheme-1

Process-1 is a typical coal based sponge iron plant of capacity 100 t/d being operated in an industrial cluster of India. Process flow diagram (PFD) of Process-1 with material and energy balanced data is shown in Fig. 4.1 and its detailed process description is given in Section 3.1. In this section detailed analysis of results of Scheme-1 for Process-1 is discussed.

Scheme-1 considers solid streams such as kiln feed (iron ore, feed coal and dolomite) and slinger coal as process streams as discussed in Section 4.3.1. In the existing sponge iron process these streams enter into the rotary kiln at ambient temperature and are preheated inside the kiln with the help of heat produced through combustion of coal. In the present process, waste gas leaves the ESP at significantly high temperature i.e. 220^{O}C, as shown in Fig. 4.1. If feed streams to the kiln are preheated outside the kiln before it enters to the kiln using heat available with waste gas exiting ESP, a substantial amount of coal may be saved. Based on it the Scheme-1 has been formulated. Thus, in Scheme-1 kiln feed and slinger coal are preheated from 30^{O}C to 120^{O}C using waste gas exiting ESP and kiln air is preheated using the remaining heat from ESP exit gas. Further, after preheating, the same ESP exit gas is used to cool kiln outlet stream from 1020^{O}C to 110^{O}C. Therefore, the stream data of Scheme-1 for Process-1 includes solid streams such as kiln feed and slinger coal and gas streams such as kiln air and waste gas, as shown in Table 4.1. It shows that the present scheme contains two hot and four cold streams.

The energy requirement for Scheme-1 is targeted using pinch analysis [161] considering ΔT_{min} as 50^{O}C. This scheme utilizes waste gas as a hot stream. As waste gas stream consists of lower heat transfer coefficient, it requires significantly larger area for heat transfer. It can be compensated to some extent while increasing temperature difference (ΔT) between waste gas and cold stream. Therefore, large ΔT is preferred for process [161] where gas streams are involved in heat exchangers. Due to this fact, ΔT_{min} is considered as 50^{O}C for heat integration through Scheme-1. For stream data of Scheme-1, shown in Table 4.1, both hot and cold utilities are found to be 70.82 kW as can be depicted through composite curves drawn in Fig. 6.1. In fact, in the present process only hot utility is used for further computation as it is provided by combustion of coal and may affect the coal consumption of the process. Further, Fig. 6.1 shows that 2642.55 kW of heat can be recovered within the process, which is called internal heat exchange.

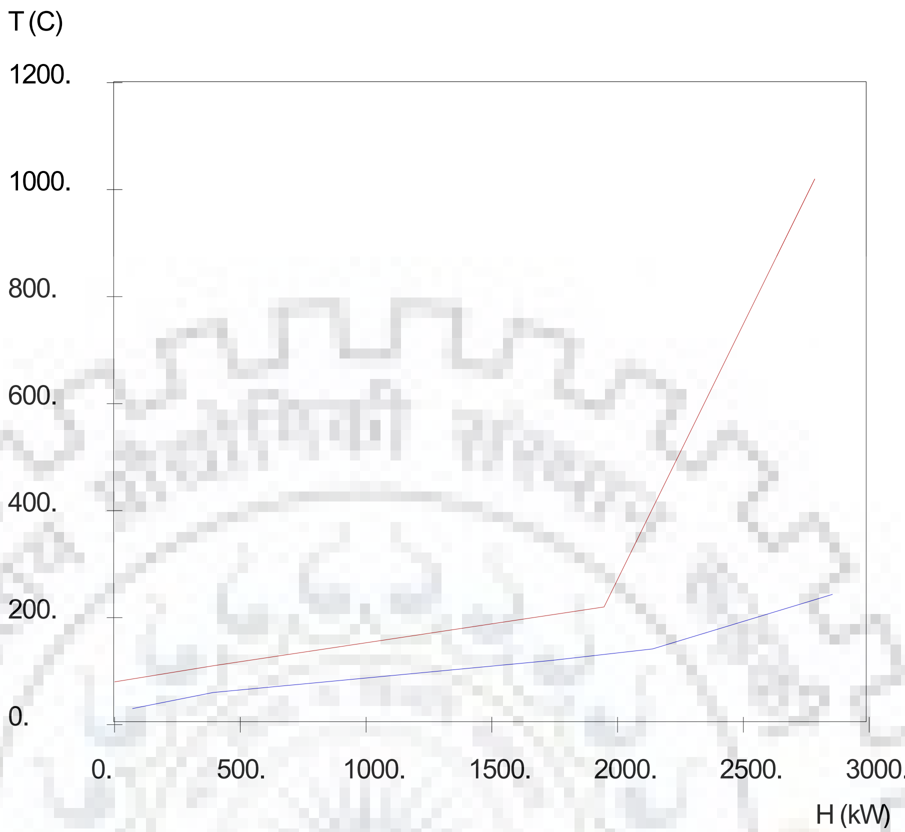


Fig. 6.1 Composite curves of Scheme-1 for Process-1

In this process coal is the only source of energy, which is provided when coal is burnt along with air. Consequently, in the present process coal and air are considered as hot utilities. In the Scheme-1 coal and air are preheated using waste gas and at the same time these are used as hot utilities. It proves the dual nature of coal and air. In the actual process coal enters the kiln at 30^{O}C and then preheated inside the kiln up to reaction temperature. Heat required for this preheating is supplied through combustion of coal in the presence of air. However, applying heat integration in Scheme-1, coal is preheated outside the kiln using waste gas up to 120^{O}C and then fed to the rotary kiln. Further, it is preheated from 120^{O}C to the reaction temperature inside the kiln using heat of combustion of coal. This in turn reduces amount of coal consumed for preheating coal inside the kiln. Corresponding requirement of air for burning of coal also reduces. Further, reduced amounts of coal and air are preheated using waste gas. Consequently, amount of coal and hence air decreases. Thus, a trial and error computational technique is required to estimate amounts of coal and air in this process, which is also shown in Fig. 5.1.

Results of different iterations to predict amounts of coal and air are summarized in Table 6.1. It shows that coal and air consumptions at Iteration-0 (existing system) are 5.36 t/h and 24.80 t/h, respectively, as shown in Fig. 4.1 also. Considering preheating of kiln feed, slinger coal and air up to 120^{O}C, 120^{O}C and 243.7^{O}C, respectively, as shown in Table 4.1, and hot utility as 70.82 kW coal consumption is computed using Equation 4.9 in iteration-1, which is found as 4.61 t/h as shown in stage-1 of iteration-1 of Table 6.1. Corresponding amount of air is computed using ratio of air to coal as maintained in iteration-0, which is 4.63. Consequently, air is found as 21.33 t/h. Thus, for first stage of iteration-1 coal and air is found as 4.61 t/h and 21.33 t/h, respectively. However, while computing coal consumption in first stage of iteration-1, through Equation 4.9, air available in iteration-0 i.e. 24.80 t/h is used. These values of air and coal are not in same ratio as 4.63. Thus, revised value of air i.e. 21.33 t/h is used in second stage of iteration-1 to compute the amount of coal, which is found to be 4.36 t/h. Corresponding amount of air using ratio of 4.63 is 20.20 t/h. Similarly, different stages of iteration-1 are performed till values of coal and air become equal in two consecutive stages. The final amounts of coal and air, in iteration-1, are 4.24 t/h and 19.64 t/h, as shown in Table 6.1, and these values maintain the ratio of 4.63. Reductions in coal and air consumption are due to heat integration through Scheme-1 to the process in which coal and air are preheated from 30^{O}C to 120^{O}C and from 30^{O}C to 243.7^{O}C, respectively. Therefore, the amount of coal required to preheat coal inside the kiln is reduced. Now, the revised values of coal and air i.e. 4.24 t/h and 19.64 t/h are considered to modify the stream data, shown in Table 4.1. For the revised stream data hot utility is found as 36.35 kW as shown in Table 6.1. These revised values are used to compute coal consumption in iteration-2. Following similar procedure for computing coal and air consumptions as carried out in iteration-1, results for other iterations are found in different stages and tabulated in Table 6.1. Final coal and air requirements are 4.41 t/h and 20.42 t/h, respectively. Therefore, 17.67% of coal and air requirements are reduced in comparison to the existing process.

There are two salient features that can be observed, for Process-1 and Scheme-1, from Table 6.1. Values of coal and air in iteration-2 are slightly increased in comparison to iteration-1. This is due to the fact that in iteration-1 kiln feed, slinger coal and air are preheated from 30^{O}C to 120^{O}C, from 30^{O}C to 120^{O}C and from 30^{O}C to 243.7^{O}C, respectively. For this purpose, waste gas of 44.18 t/h cools down from 220^{O}C to 80^{O}C. Due to this preheating using waste gas, amounts of coal and air are reduced as shown in Table 6.1. Further, revised values of coal and air are used for material balance. As a result of it, amount of waste gas exiting from ESP is reduced to 33.83 t/h, which is cooled from 220^{O}C to 80^{O}C in iteration-2. Thus, the overall heat available for preheating is reduced in iteration-2 than that was for iteration-1. Now, it preheats the kiln feed, slinger coal and air from 30°C to 120°C, from 30°C to 120°C and from 30°C to 230°C respectively. It shows that in teration-2 the temperature of air after preheating is less in comparison to iteration-1. Thus, air to be preheated inside the kiln is from 230°C to reaction temperature in iteration-2 instead of from 243.7°C to reaction temperature in iteration-1. Therefore, preheating of air requires slightly higher amount of coal in iteration-2 than that is used in iteration-1. Consequently, amount of air in iteration-2 also increases slightly to maintain air to coal ratio 4.63.

The other salient feature can be observed, for Process-1 and Scheme-1, through iteration-5 of Table 6.1. Usually, iterations to compute coal and air requirements can be stopped once values of coal and air become equal in two consecutive iterations. At this point, one can construct the final heat exchange network (HEN). However, it was found from Table 6.1 that the values of coal and air in iteration-5 are increased in comparison to iteration-4. This is due to the fact that in iteration-4 kiln feed, slinger coal and air are preheated from 30°C to 120°C, from 30°C to 120°C and from 30°C to 230.2°C respectively. For this purpose, waste gas of 33.98 t/h is used which cools down from 220°C to 80°C. However, HEN, similar to the network shown in Fig. 6.2, indicates that the waste gas can be cooled only from 220°C to 113.8°C instead of from 220°C to 80°C as air cannot be heated to more than 170°C (=220°C-80°C) to maintain ΔT_{min} as 50°C in HEN. Due to this the hot utility requirement, computed using pinch analysis [161], increased from 36.75 kW in iteration-4 to 335 kW in iteration-5. This increase in hot utility is used to heat air from 170°C to 230.2°C. Therefore, higher amount of coal is required in iteration-5 than that is used in iteration-4. Consequently, amount of air in iteration-5 also increases to maintain air to coal ratio 4.63. Results of all iterations are tabulated in Table 6.1 and the final HEN of heat integrated Process-1 found through Scheme-1 is shown in Fig. 6.2.

Table 6.1 Iteration results of Scheme-1 for Process-1

Iteration No	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	5,360.0	24,802.0
1	70.82	1	4,610.4	21,333.6
		2	4,364.5	20,195.8
		3	4,283.9	19,822.4
		4	4,257.4	19,700.0
		5	4,248.7	19,659.8
		6	4,245.9	19,646.6
		7	4,244.9	19,642.3
		8	4,244.6	19,640.9
		9	4,244.5	19,640.4
		10	4,244.5	19,640.2
		11	4,244.5	19,640.2
2	36.35		4,262.7	19,724.4
3	36.77		4,262.4	19,723.3
4	36.77		4,262.4	19,723.3
5	334.98		4,407.1	20,392.6
6	346.34		4,412.6	20,418.1
7	346.79		4,412.8	20,419.1
8	346.79		4,412.8	20,419.1

Temperatures and heat loads shown in HEN are based on values of final iteration reported in Table 6.1. Fig. 6.2 indicates that four new heat exchangers (HX) are required in the HEN in the modified design of Process-1 through Scheme-1. Fig. 6.2 shows that waste gas stream (H2) is splitted into two streams, which is done while maintaining ΔT_{min} at 50^{O}C in the process. One part of waste gas exchanges heat with slinger coal (C3) to get cooled from 220^{O}C to 198^{O}C through HX-2, which has load of 62.5 kW. It is a gas-solid exchanger where slinger coal (C3) is preheated from 30^{O}C to 120^{O}C. The same part of waste gas is further cooled to 114.4^{O}C in HX-4, which has load of 236.8 kW. It is a gas-solid exchanger where kiln feed (C2) is preheated from 30^{O}C to

$120^{O}C$. The second part of waste gas is cooled from $220^{O}C$ to $114.4^{O}C$ in a gas-gas exchanger, HX-3, having a load of 806.9 kW. It is required to preheat kiln air up to $170^{O}C$ only as air cannot be heated beyond $170^{O}C$ ($220^{O}C$-$50^{O}C$) in the heat exchanger due to the fact that maximum temperature of waste gas is $220^{O}C$ and at least $50^{O}C$ temperature difference is required for gas to gas heat transfer. Heating of air from $170^{O}C$ to $230.2^{O}C$ requires additional 346.8 kW of heat, as evident from Fig. 6.2, which should be supplied through hot utility i.e. through combustion of coal as it is the only source of heat available in the process. Finally, kiln outlet stream (H1) is cooled from $1020^{O}C$ to $110^{O}C$ using waste gas (C1) at $60^{O}C$ in a gas-solid heat exchanger, HX-1, having a heat load of 953.1 kW. Consequently, waste gas is heated up to $161.8^{O}C$ as shown in Fig. 6.2. Thus, HEN shown in Fig. 6.2 includes four heat exchangers, in which no heat exchanger violates ΔT_{min}.

However, in the Process-1 internal heat exchange, shown in the composite curve (Fig 6.1), before modification through Scheme-1 is different from the internal heat exchange, shown in HEN, after modification and these are 2642.55 kW and 2062.19 kW, respectively. This difference in internal heat exchange is due to the change in flow rates and temperatures of kiln inlet streams (kiln feed, slinger coal and kiln air) and waste gas stream. After design modification through Scheme-1 amounts of kiln inlet streams and waste gas generation reduce and hence, net heat contents of these also decrease. For example, the flow rate of kiln air is 24.8 t/h and 20.42 t/h, before and after modification, respectively. Also, before modification, the kiln air is preheated to $243.7^{O}C$ whereas in actual practice it is preheated to only $230.2^{O}C$ as shown in Fig. 6.2. Therefore, variation in internal heat exchange is shown in HEN in comparison to that in composite curve.

In Scheme-1 five ducts are required for carrying waste gas from ESP to different places where it exchanges heat before it is directed to the chimney. Initially, gas is required to bring from ESP to kiln inlet, which needs a duct of length 25m. Further, following ducts are used: Second duct of 40m length from kiln inlet to kiln outlet to preheat slinger coal, third duct of length 40m carries waste gas from kiln outlet to kiln inlet to heat the kiln feed, fourth duct of length 65m from kiln inlet to cooler exit. Finally, a duct of length 70m is employed to carry waste gas from cooler inlet to chimney. Lengths of these ducts are as per actual distances between equipment in Process-1. The modified PFD for Process-1 through Scheme-1 is shown in Fig. 6.3. Reduced amounts of coal and air cause less waste gas to discharge from the process. The waste gas generation, in existing system as shown in Fig. 4.1, is 44.17 t/h, which is reduced to 35.23 t/h in the modified process shown in Fig. 6.3. Thus, 20.24% less waste gas is generated in Process-1 which is an additional benefit of heat integration through Scheme-1. Similarly, 95.54% of water consumption is reduced in the modified system in comparison to existing PFD.

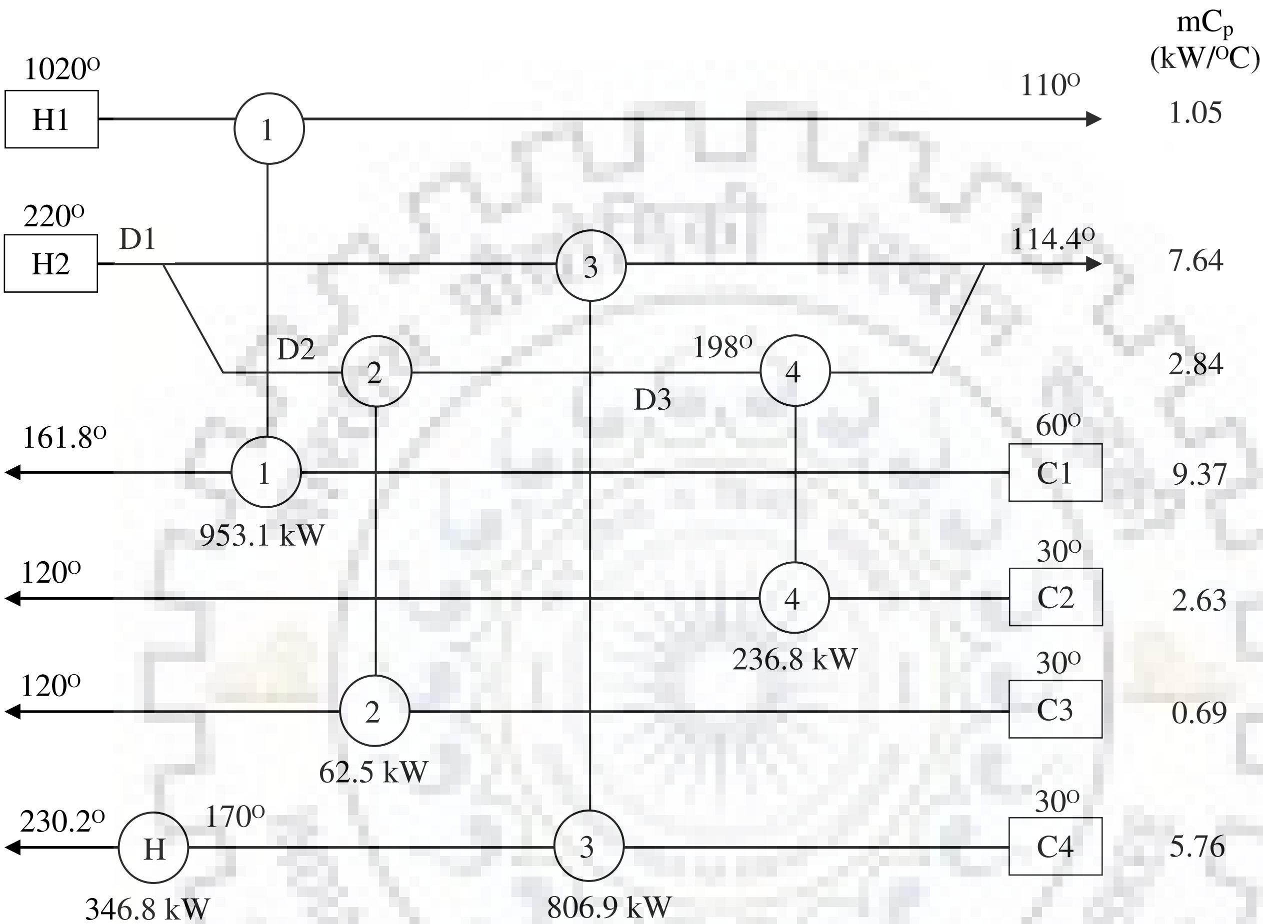


Fig. 6.2 Heat Exchanger Network of Scheme-1 for Process-1

6.1.1.1 Operational aspects of Process-1 for Scheme-1

Scheme-1 requires five numbers of ducts as discussed above. An induced draft (ID) fan is mounted just before the chimney to maintain the flow of waste gas in the existing plant. However, for heat integration through Scheme-1 waste gas should traverse from ESP to different places where it exchanges heat before it is directed back to the chimney. Hence, forced draft (FD) fans are used to carry waste gas through these ducts. Primarily, an insulated duct (D1) of length 25m and diameter 0.3m is employed to transfer waste gas, having CP of 10.48 kW/°C, from ESP to kiln inlet. While passing through this duct temperature of the waste gas drops down from 220°C to 219.7°C as shown in Fig. 6.3. The methodology for calculating temperature profile is taken from the work of Prasad [130]. Once waste gas reaches the kiln inlet at 219.7°C this duct is divided into two branches. First branch of duct carries waste gas of CP equal to 7.64 kW/°C at 219.7°C. At this temperature waste gas enters into a shell and tube gas-gas heat exchanger, HX-3, shown in Fig. 6.3. As waste gas has more fouling tendency, it is allocated to tube side and thus, kiln air is entered to shell side. Considering overall heat transfer coefficient (U) as 50 $W/m^{2}°C$ [154] heat transfer area is predicted as 241.3 m^2. After exchanging heat with kiln air waste gas temperature drops from 219.7°C to 114.1°C instead of from 220°C to 114.4°C as shown in Fig. 6.2. Second branch of 25m duct is attached with an insulated duct (D2) of 40m length and 0.3m diameter. It carries remaining waste gas with value of CP of as 2.84 kW/°C (10.48-7.64). Waste gas enters into this duct at 219.7°C and its temperature is dropped to 219°C while traversing through the duct. Further, it goes inside the gas-solid exchanger, HX-2 in Fig. 6.3, at 219°C, which is different than that shown in Fig. 6.2, where waste gas enters at 220°C. Exchanging 62.49 kW of heat in HX-2 waste gas leaves the exchanger at 196.9°C after transferring its heat to slinger coal. However, in Fig. 6.2 waste gas exits the exchanger, HX-2, at 198°C.

Waste gas exiting exchanger, HX-2, at 196.9°C moves within an insulated duct (D3) of 40 m length and 0.3m diameter. While passing through this duct gas drops its temperature to 196.2°C and then enters to the gas-solid exchanger, HX-4. In this exchanger kiln feed is preheated from 30°C to 120°C while exchanging load of 236.84 kW. The temperature of the waste gas reduces to 112.6°C after exchanging the heat with kiln feed in HX-4. As waste gas does not react with kiln feed direct contact gas-solid exchanger, HX-4, can easily be used. Once kiln feed and kiln air are preheated, the waste gas exiting exchangers HX-3 and HX-4, respectively, are mixed to acquire temperature of 113.7°C. Now, this waste gas is used to cool the kiln outlet. However, temperature of waste gas should bring down to 60°C to maintain ΔT_{min}. For this purpose an extra gas to gas heat exchanger, HX-5, is used considering atmospheric air as the cooling medium. Hence, total five heat exchangers (three gas-solid and two gas-gas) are required for heat integration through Scheme-1. After attaining 60°C in HX-5 waste gas is passed through a non-insulated duct (D4) of 65m length and 0.3m diameter dropping its temperature to 58.8°C before entering to gas-solid heat exchanger, HX-1 as shown in Fig. 6.3. Here, kiln outlet is cooled from 1020°C to 110°C while exchanging 953.1 kW heat. As a result, waste gas heats up to 160.6°C. Finally, after preheating

C2, C3 and C4 streams and cooling H1 and H2 streams as shown in Table 4.1 waste gas (C1) moves through a non-insulated duct (D5) of 70 m length and 0.3 m diameter. This duct carries waste gas to the chimney where it enters at 153.5°C. Temperature profiles of all five ducts are shown in Fig. 6.4. Mild steel is used for casing of the duct and for insulation glass wool is placed at outer surface of the duct.

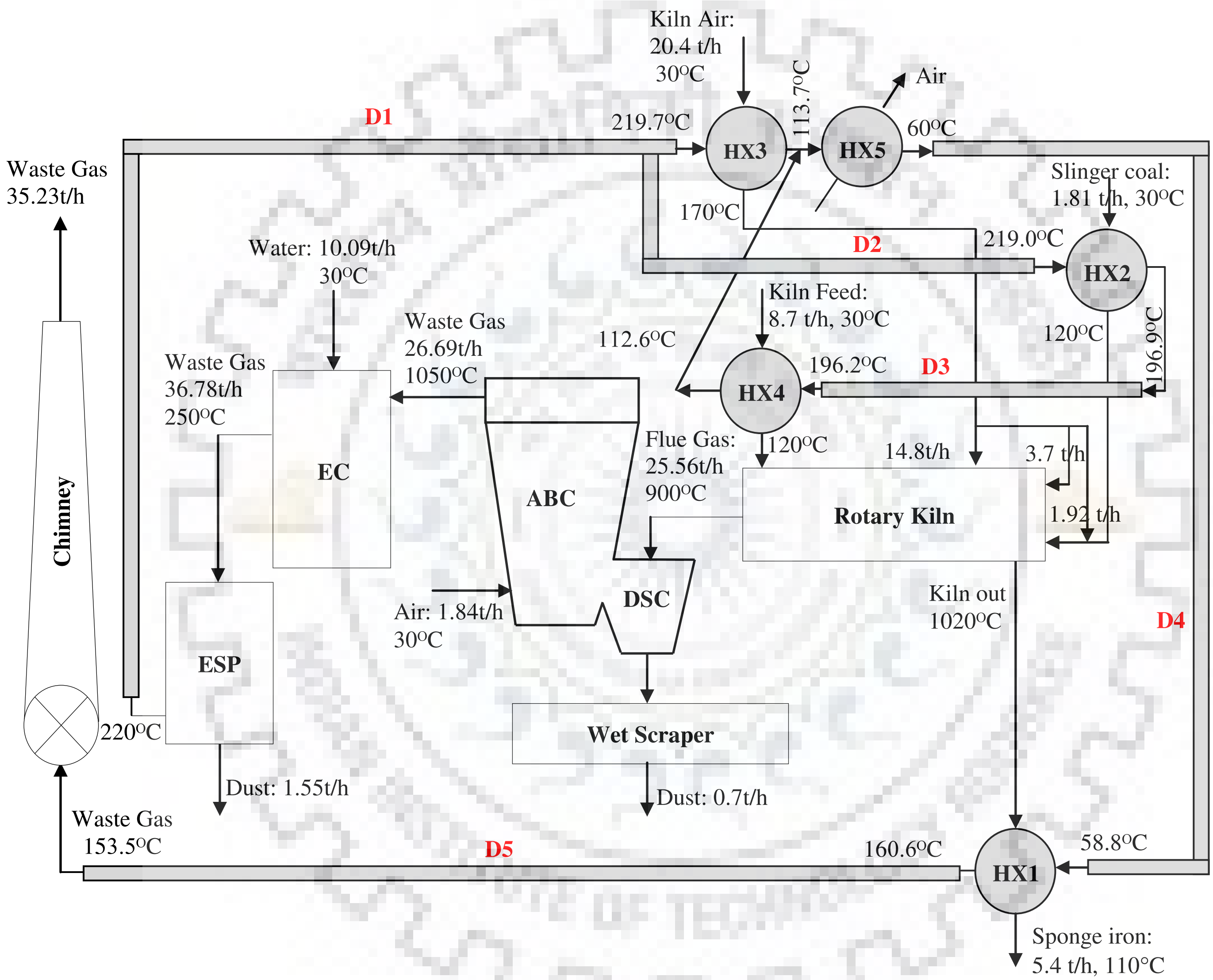


Fig. 6.3 Modified process flow diagram of Process-1 for Scheme-1

In addition, there is certain amount of pressure drop in the duct when waste gas flows from ESP to kiln inlet, from kiln inlet to kiln outlet, from kiln outlet to kiln inlet, from kiln inlet to cooler exit and finally from kiln outlet to chimney. The pressure drops found as 0.0552 atm, 0.0081 atm, 0.0076 atm, 0.0974 atm and 0.132 atm for ducts of length 25m, 40m, 40m, 65m, and 70m, respectively. Details of all five ducts, used for modification through Scheme-1, is summarized in Table 6.2. The methodology for calculating pressure drops is taken from the work of Prasad [133]. These pressure drops need to be compensated either by the process or by outside means using some drives. FD fans are used for this purpose where one FD can sustain 1.26 atm pressure drop [177]. Accordingly, five FD fans are required; one for each duct, to maintain the necessary pressure drops in these ducts for Scheme-1.

Table 6.2 Pressure and temperature drops in ducts for Process-1 Scheme-1

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	25	ESP exit gas	ESP	Kiln inlet	0.055	220	219.7
D2	40	ESP exit gas	Kiln inlet	Kiln outlet	0.008	219.7	219.0
D3	40	ESP exit gas	Kiln outlet	Kiln inlet	0.008	196.9	196.2
D4	65	ESP exit gas	Kiln inlet	RC exit	0.097	60	58.8
D5	70	ESP exit gas	RC inlet	Chimney	0.132	160.6	153.5

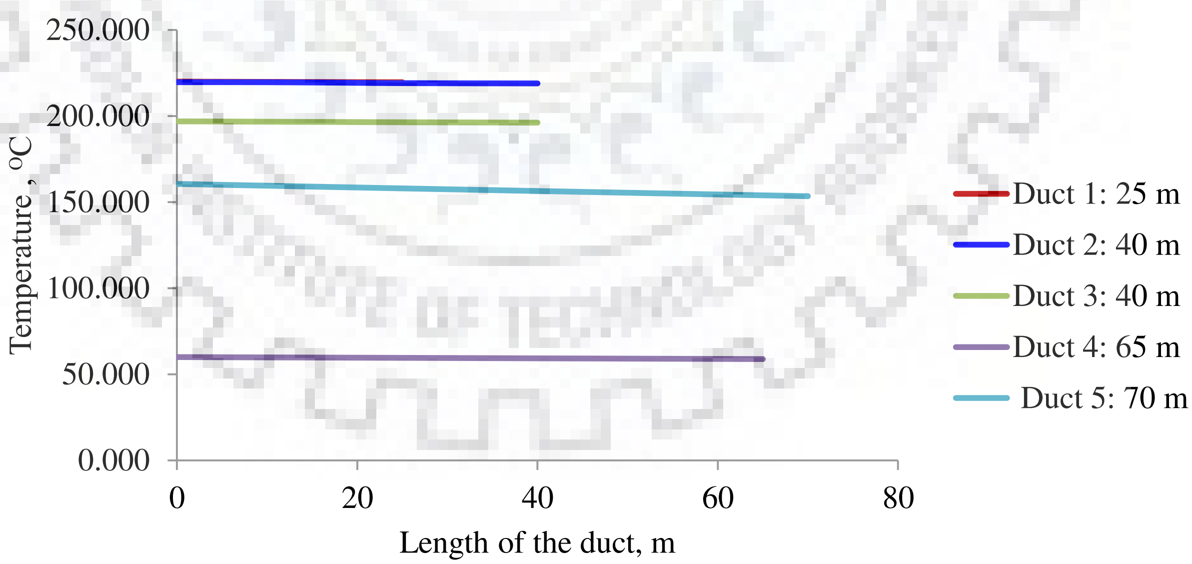


Fig. 6.4 Temperature profile of waste gas in ducts for Process-1 and Scheme-1

6.1.1.2 Economic aspects of Process-1 for Scheme-1

The economic analysis of Scheme-1 in terms of operating cost, capital investment, TAC, savings and payback period is shown in Table 6.3. In Scheme-1 five heat exchangers are required. Amongst these, three are gas-solid exchangers (HX-1, HX-2 and HX-4) and two are gas-gas shell and tube heat exchangers (HX-3 and HX-5). Capital cost of shell and tube heat exchangers, HX-3 and HX-5, are found to be Rs. 42.78 lakh and Rs. 49.23 lakh, respectively, based on Equation 3.1. Capital cost of exchangers, HX-1, HX-2 and HX-4, are estimated as Rs. 113.5 lakh, Rs. 63.97 lakh and Rs. 73.67 lakh, respectively. Capital costs of five ducts of total length 240m is found to be Rs. 25.94 lakh. For five FD fans total capital cost is computed as Rs. 336.14 lakh. Thus, total capital investment required for modification of Process-1 through Scheme-1 is 705.23 lakh. The costing methods for gas-solid heat exchangers, ducts and FD fans are considered from the work of Prasad [133]. Total coal consumption, through Scheme-1, reduces by 17.67% in comparison to the existing system, which saves Rs. 2131.2/h. Similarly, the modification reduces the water consumption by 95.54%, which saves Rs 11668.48/h. It should be noted that the operating cost does not include costs of iron ore and dolomite as these are fixed quantities for all design modifications proposed in the current work and thus, can be excluded from comparative study. Further, electricity cost for five fans is predicted as Rs 0.34 lakh/day. Thus, the net savings in terms of coal, water and electricity is found to be Rs 2.97 lakh/day. Considering 10% rate of return, for all proposed schemes and strategies net profit value (NPV), internal rate of return (IRR) and discounted payback period (DPP) are computed and reported in Table 6.3. The NPV is 5959.7 lakh with 130.7% of IRR and discounted payback period of 8.58 months. Capital recovery factor is used to compute TAC considering 10% rate of return with a plant life of 10 years and it comes out to be Rs 1156.4 lakh, which includes operating cost and annualized capital cost. Detailed economic analysis is shown in Appendix C. All cost coefficients used in economic analysis is tabulated in Table 6.3A.

The method for computing the cost of some equipment is taken from published articles on sponge iron plants so that the advantage of the proposed schemes can be compared with their findings. It is considered only for comparison purpose. As these costs are older, cost index factors are used to predict recent cost of these components. Cost index factors [173] in 1995, 2009 and 2017 are 381.1, 521.9 and 567.5, respectively. Considering these factors, total capital cost is found as 797.0 lakh instead of 705.23 lakh as reported in Table 6. The discounted payback period thus found is

9.7 months instead of 8.58 months. Similarly NPV and IRR are found to be 5867.9 lakh and 114.6% which are slightly lower than the values reported in Table 6.3. Economic analysis with cost index factors, for all design modifications considered in the current study, is summarized in Appendix C.

Table 6.3 Economic analysis of Scheme-1 for Process-1

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/year)	Savings (Lakh/year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	869.76	HX-1	113.50	1156.4	1084.68	5959.7	130.7	8.58
Water	47.73	HX-2	63.97					
Electricity	124.17	HX-3	42.78					
		HX-4	73.67					
		HX-5	49.23					
		Ducts	25.94					
		FD fans	336.14					
		Total	705.23					

Table 6.3A: Cost coefficients used in economic analysis

S. No	Component	Cost
1	Fixed cost of G-G heat exchanger (Rs)	Rs. 1368000 + 34200 $(A)^{0.81}$ Where, A is heat exchanger area [150]
2	G-S heat exchanger: Standard cost components of G-S heat exchanger (load = 4136 kW) [133]	
	Cost of rotary drier with accessories such as girth gears, support rollers, drier mounted instruments, weigh feeders, conveyors and vibrofeeders	Rs. 23 Lakh
	Cost of three phase induction motor for exchanger (Capacity: 55kW)	Rs. 22 Lakh
	Cost of gear box for exchanger	Rs. 31 Lakh
	Approximate installation cost	Rs. 7.5 Lakh
3	FD fan: Standard cost components of FD fan (Capacity = 122000 m^3/hr) [133]	
	Cost of FD fan	Rs. 65 Lakh
	Cost of three phase induction motor for FD fan (Capacity: 80kW)	Rs. 41 Lakh
	Cost of bearing of FD fan	Rs. 11.3 Lakh
	Cost of coupling used to couple fan and motor	Rs. 1.6 Lakh
	Approximate installation cost	Rs. 15500/-
	Approximate cost of electrical works including material cost	Rs. 26 Lakh
	Approximate volume of work and cost	Rs. 21 Lakh
4	Duct cost [133]	
	Cost of mild steel	Rs. 33020/t
	Standard cost for manufacturing, welding and installation of duct	Rs. 154415/-
	Cost of insulation material (glass wool)	Rs. 650/t
	Cost of insulation material (ceramic fibre)	Rs. 46000/t
	Commissioning cost	Rs. 5 Lakh
5	Cost of boiler and steam turbine [59]	Rs. 46800/kW

6.1.2 Process-1 and Scheme-2

In this section detailed analysis of results of Scheme-2 for Process-1 is discussed. Scheme-2 is discussed in detail under Section 4.3.2. In this scheme, waste gas exiting ABC at 1050^{O}C is used for heat integration. If feed streams to the kiln are preheated outside the kiln before it enters into the kiln using heat available with ABC exit gas and kiln outlet stream, a substantial amount of coal and water may be saved. Based on it Scheme-2 has been formulated where kiln feed and slinger coal are preheated from 30^{O}C to 300^{O}C using ABC exit gas and kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air. Kiln air is further preheated to 300^{O}C using ABC exit gas. Therefore, stream data for Scheme-2 includes solid streams such as kiln feed and slinger coal and gas streams such as kiln air and waste gas, as shown in Table 4.2. It involves two hot and three cold streams.

For the stream data of Scheme-2 for Process-1, shown in Table 4.2, both hot and cold utilities are found to be zero as can be evident from composite curves presented in Fig. 6.5. ΔT_{min} is considered as 50^{O}C for Scheme-2 also. It is a threshold problem as hot and cold utility requirements are zero [161]. That means the complete problem formulated through Table 4.2 is in energy balance and no additional heating and cooling utilities are required from outside. Fig. 6.5 shows that 2901.27 kW of heat can be recovered within the process, which is called internal heat exchange.

The amount of coal consumption is computed through a trial and error method, as coal works as utility and process stream simultaneously, in the similar manner as discussed in Scheme-1. The coal and air consumption at Iteration-0 (existing system) are 5.36 t/h and 24.80 t/h, respectively, as shown in Fig. 4.1 and Table 6.4. For first stage of iteration-1, coal consumption is calculated using Equation 4.9, which is found to be 4.27 t/h. The corresponding amount of air consumed is computed using ratio of air to coal, which is 4.63, and found as 19.78 t/h. However, while computing coal consumption in first stage of iteration-1, through Equation 4.9, air used is that available in iteration-0 i.e. 24.80 t/h. This air to coal is not in same ratio as 4.63. Thus, revised value of air i.e. 19.78 t/h is used in second stage of iteration-1 to compute amount of coal, which is found to be 3.95 t/h. Corresponding amount of air is 18.28 t/h. Similarly, different stages of iteration-1 are performed till values of coal and air become equal in two consecutive stages. Final amounts of coal and air, in iteration-1, are 3.81 t/h and 17.64 t/h, respectively, and these values maintain the ratio of 4.63. Following similar procedure for computing coal and air requirements, as

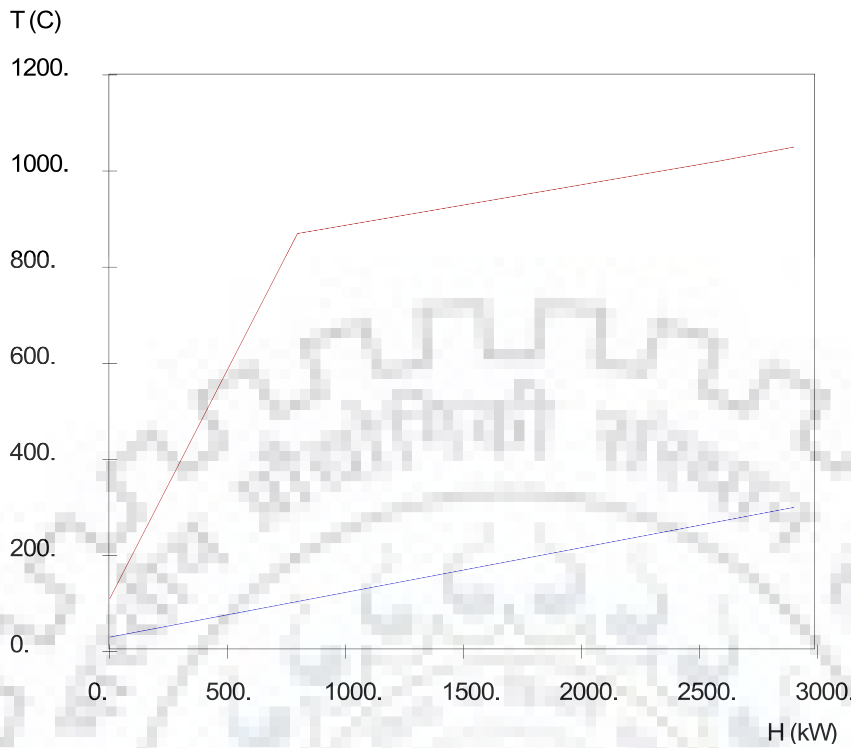


Fig 6.5 Composite curves of Scheme-2 for Process-1

Table 6.4 Iteration results of Scheme-2 for Process-1

Iteration No	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	0	5,360.0	24,802.0
1	1	4,274.3	19,778.3
	2	3,950.0	18,277.5
	3	3,853.1	17,829.1
	4	3,824.1	17,695.1
	5	3,815.5	17,655.1
	6	3,812.9	17,643.2
	7	3,812.1	17,639.6
	8	3,811.9	17,638.5
	9	3,811.8	17,638.2
	10	3,811.8	17,638.1
	11	3,811.8	17,638.1
2		3,811.8	17,638.1

carried out in iteration-1, results for other iterations are found and reported in Table 6.4. Final coal and air requirements are 3.81 t/h and 17.64 t/h respectively. In fact, 28.88 % of coal and air requirements are reduced in comparison to the existing process. HEN of Process-1 for Scheme-2 is shown in Fig. 6.6, which includes four heat exchangers that do not violate ΔT_{min}.

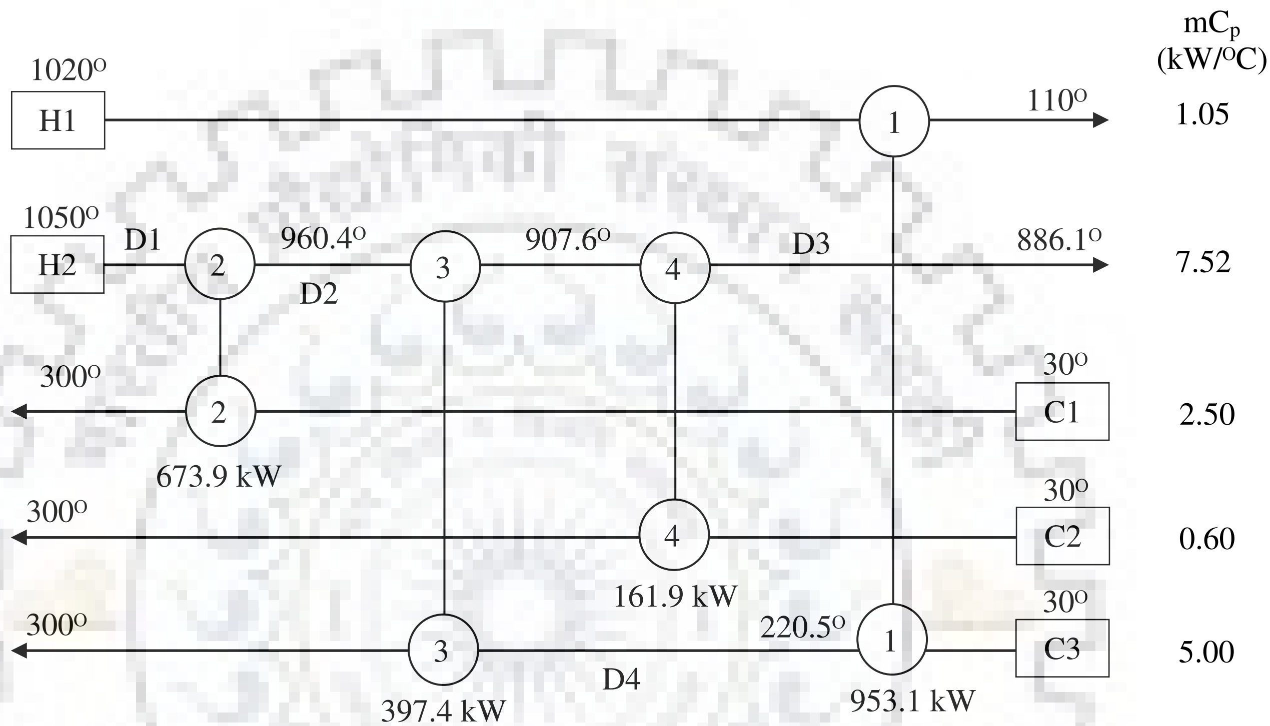


Fig. 6.6 Heat Exchanger Network of Scheme-2 for Process-1

Temperatures and heat loads shown in HEN (Fig. 6.6) are based on values of final iteration reported in Table 6.4. This Fig. 6.6 includes four new heat exchangers (HX) as additional equipment to be employed in the modified design of Process-1 through Scheme-2. The exchanger, HX-1, is a gas-solid heat exchanger, which exchanges 953.1 kW heat between kiln outlet stream (H1) and kiln air (C3). Hence, kiln outlet stream cools from 1020°C to 110°C whereas temperature of kiln air increased from 30°C to 220.5°C. Second heat exchanger, HX-2, is a gas-solid heat exchanger having a heat load of 673.9 kW. It preheats the kiln feed (C1) from 30°C to 300°C using waste gas stream (H2), which exits ABC at 1050°C. While exchanging heat in HX-2 waste gas temperature drops to 960.4°C. Further, it enters to HX-3, which is gas-gas heat exchanger with heat load of 397.4 kW. Kiln air (C3) is further preheated from 220.5°C to 300°C in this exchanger using waste gas available at 960.4°C and subsequently, its temperature drops to 907.6°C. At this temperature waste gas goes through a gas-solid heat exchanger, HX-4, of 161.9 kW heat load where slinger coal (C2) is preheated from 30^{O}C to 300^{O}C. Consequently, waste gas temperature drops to 886.1^{O}C as shown in Fig. 6.6. The HEN, shown in Fig. 6.6, indicates ideal temperatures where heat loss while traversing through ducts is not considered. This fact should be checked through design.

The internal heat exchange, shown in composite curve (Fig 6.5), before modification through Scheme-2 is different than that shown in HEN, after modification and these are 2901.2 kW and 2186.3 kW, respectively. This difference is due to the fact that after design modification through Scheme-2 amounts of kiln inlet streams and ABC exit gas decrease along with net heat content of these. For example, the flow rate of waste gas exiting ABC is 33.2 t/h and 23.1 t/h, before and after modification, respectively. Also, before modification, recovery of waste heat from ABC exit gas is targeted by cooling it from 1050^{O}C to 870^{O}C whereas in the modified design it is cooled to only 886.1^{O}C as shown in Fig. 6.6.

In Scheme-2 and Process-1 four ducts are required. Amongst these three are used for carrying waste gas from ABC to different places where it exchange heat before it is directed to ABC and one is required for carrying kiln air. First duct of length 12m carries waste gas from ABC to kiln inlet whereas, second duct of 40m length from kiln inlet to kiln outlet is moved to preheat kiln air and slinger coal. Further, third duct of length 52m is considered to carry back waste gas from kiln outlet to ABC exit. Finally, a duct of length 7m is employed to carry kiln air from cooler inlet to kiln outlet. The modified PFD for Process-1 through Scheme-2 is shown in Fig. 6.7. The waste gas generation, in existing system as shown in Fig. 4.1, is 44.17 t/h, which is reduced to 28.37 t/h in the modified process shown in Fig. 6.7. Thus, 35.78% less waste gas is generated in Process-1 through Scheme-2. Similarly 96.97% of water saving is observed in the modified system as compared to the existing PFD.

6.1.2.1 Operational aspects of Process-1 for Scheme-2

As discussed in Scheme-1, flow of waste gas in the existing plant is maintained through ID fan mounted before the chimney. However, for heat integration through Scheme-2 waste gas traverse from ABC to different places to exchange heat with kiln inlet streams using FD fans. The modified PFD shown in Fig. 6.7 indicates that an insulated duct (D1) of length 12 m and diameter 0.3m is employed to transfer waste gas, having CP of 7.52 kW/^{O}C, from ABC to kiln inlet. While passing through this duct temperature of waste gas reduces from 1050^{o}C to 1048.1^{o}C as shown in the Fig. 6.7. Further, waste gas enters into HX-2 as shown in Fig. 6.7 to exchange heat with kiln feed and drops its temperature from 1048.1^{o}C to 958.6^{o}C instead of from 1050^{o}C to 960.4^{o}C as shown in Fig. 6.6. Waste gas exiting exchanger, HX-2, at 958.6^{o}C moves through an insulated duct (D2) of 40 m length and 0.3m diameter. While passing through this duct waste gas drops its temperature to 952.9^{o}C and then it enters to exchanger, HX-3. In this exchanger the kiln air is preheated from 220.1^{o}C to 299.6^{o}C through exchanging load of 397.4 kW instead of from 220.5^{o}C to 300^{o}C as shown in Fig. 6.6.

Further, waste gas goes into the gas-solid exchanger, HX-4, shown in Fig. 6.7 at 900.1^{o}C, instead of 907.6^{o}C shown in Fig. 6.6, and leaves it at 878.6^{o}C after transferring 161.9 kW heat to slinger coal. However, in Fig. 6.6 waste gas exits the exchanger, HX-4, at 886.1^{o}C. An insulated duct (D4) of small length 7m and diameter 0.3m is required to carry kiln air that was already preheated in HX-1. In this duct kiln air drops its temperature from 220.5^{o}C to 220.1^{o}C as shown in Fig. 6.7. Finally, waste gas moves to the ABC through a non-insulated duct (D3) of 52m length and 0.3m diameter. Consequently, temperature of gas drops from 878.6^{o}C to 872.1^{o}C. The temperature profiles of all four ducts, predicted using the method proposed by Prasad [130], are shown in Fig. 6.8. For insulation ceramic wool is placed at outer surface of the ducts, which are made up of mild steel.

Gases pass through four ducts and create certain pressure drop in these, which are evaluated using the method discussed by Prasad [133] and found as 0.0334 atm, 0.1113atm, 0.1357 atm and 0.0047 atm for ducts of length 12m, 40m, 52m and 7m, respectively. Details of these ducts are tabulated in Table 6.5. These pressure drops in ducts are compensated using FD fans where one FD can sustain 1.26 atm pressure drop [177]. Accordingly, four FD fans are required, one for each duct.

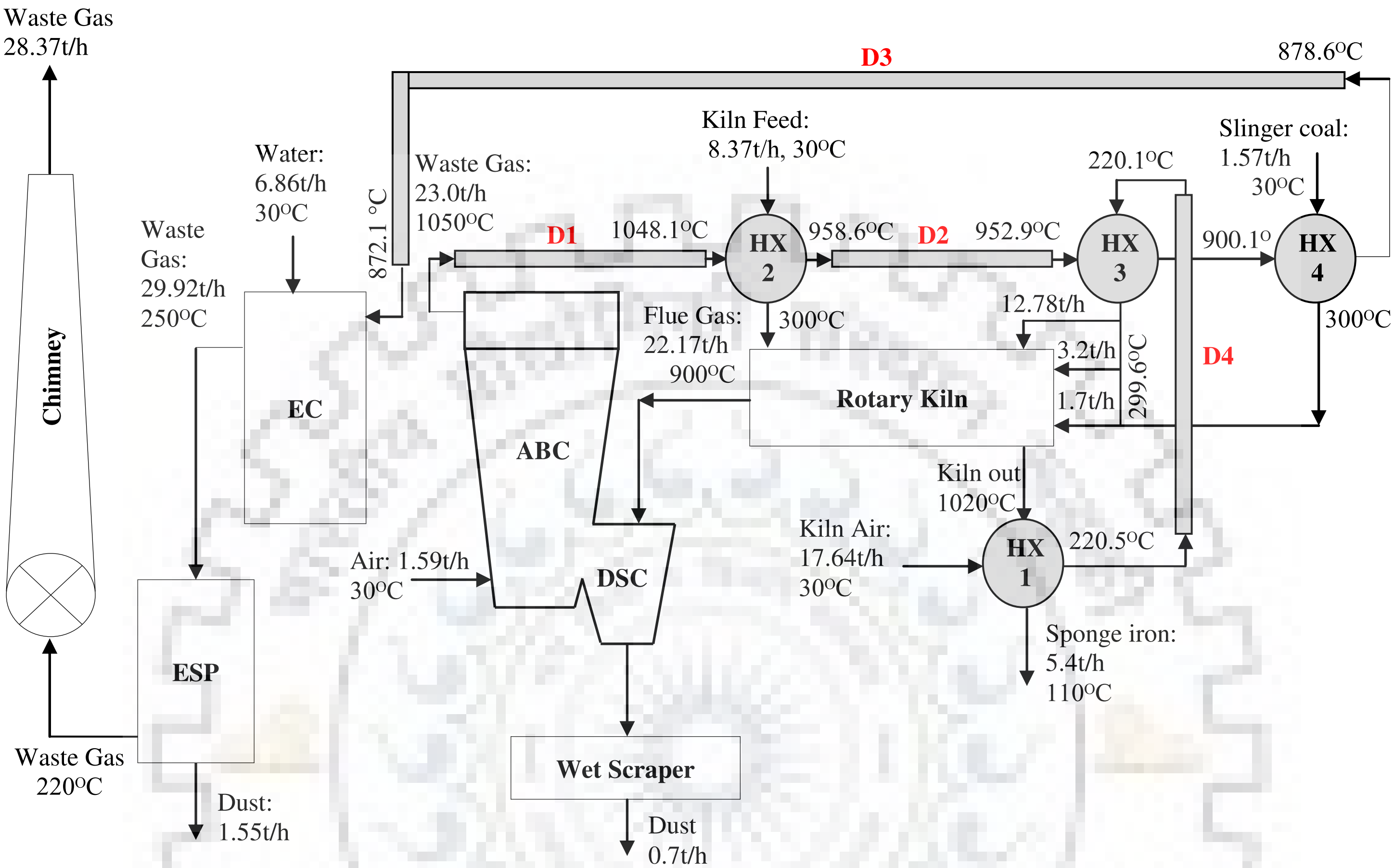


Fig. 6.7 Modified process flow diagram of Process-1 for Scheme-2

Table 6.5 Pressure and temperature drops in ducts for Process-1 Scheme-2

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	12	ABC exit gas	ABC exit	Kiln inlet	0.033	1050.0	1048.1
D2	40	ABC exit gas	Kiln inlet	Kiln outlet	0.111	958.6	952.9
D3	52	ABC exit gas	Kiln outlet	ABC exit	0.136	878.6	872.1
D4	7	Kiln air	RC inlet	Kiln outlet	0.005	220.5	220.1

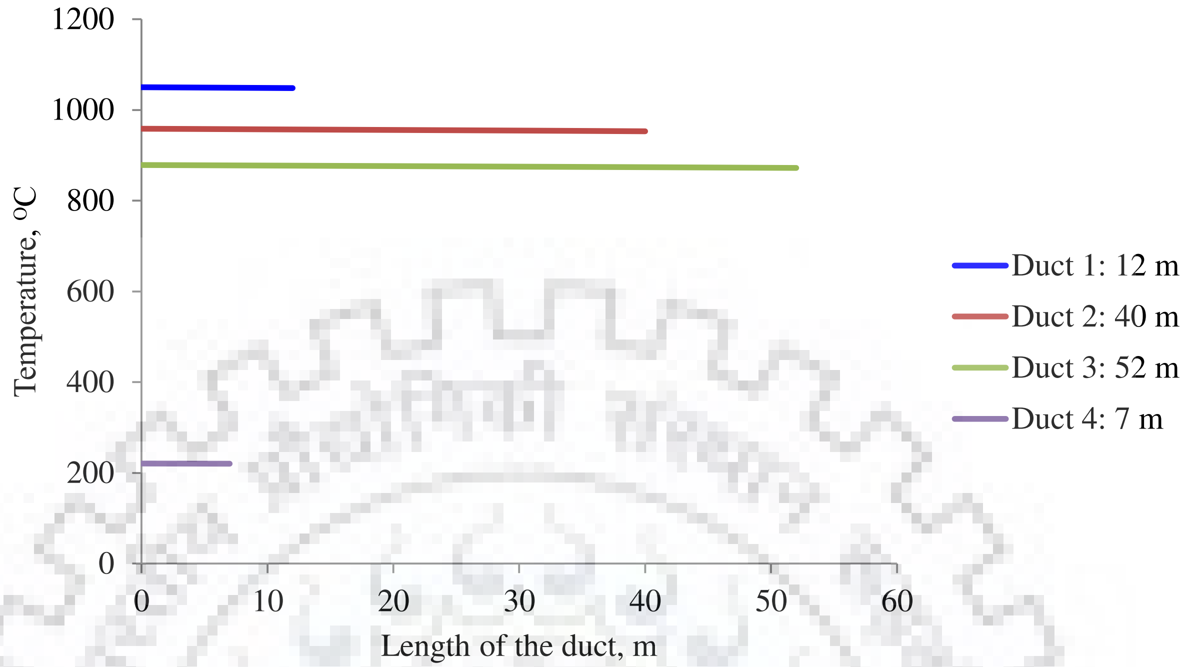


Fig. 6.8 Temperature profile of waste gas in ducts for Process-1 Scheme-2

6.1.2.2 Economic aspects of Process-1 for Scheme-2

The economic analysis of Scheme-2, similar to Scheme-1, in terms of operating cost, capital investment, total annual cost (TAC), savings and payback period are discussed here. In Scheme-2 for Process-1 four heat exchangers are required. Amongst these three are gas-solid heat exchangers (HX-1, HX-2 and HX-4) and one is gas-gas shell and tube heat exchangers (HX-3). Capital costs of these exchangers are computed as described for Process-1 and Scheme-1 and summarised in Table 6.6. Further, costs of four FD fans, four ducts of total length 111m is computed as shown the work of Prasad [133] and reported in Table 6.6. Total coal consumption, through Scheme-2, reduces by 28.88% in comparison to the existing system, which saves Rs. 3483.46/h. Similarly, the modified design reduces water consumption by 96.97%, which saves Rs 11843.06/h. Thus, the net savings in terms of coal, water and electricity is found to be Rs 3.41 lakh/day. Considering total capital cost of modification as well as savings the discounted payback period of 6.98 months is predicted. TAC of the plant comes out to be Rs 990.1 lakh considering capital recovery factor for 10% rate of return and plant life of 10 years. Detailed economic analysis is shown in Appendix C.

Table 6.6 Economic analysis of Scheme-2 for Process-1

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	751.30	HX-1	113.50	990.1	1243.26	6981.9	162.8	6.98
Water	32.44	HX-2	97.97					
Electricity	99.34	HX-3	16.23					
		HX-4	69.50					
		Ducts	28.66					
		FD fans	331.59					
		Total	657.44					

6.1.3 Process-2 and Scheme-1

Process-2 is a typical coal based sponge iron plant of capacity 62.4 t/d. PFD of Process-2 is shown in Fig. 4.2 and its detailed process description is discussed in Section 3.1. In this section detailed analysis of results of heat integration of Process-2, when Scheme-1 is applied in it, is discussed. Scheme-1 for Process-2 is formulated in the similar way as that for Process-1 and stream data is shown in Table 4.1, where two hot and four cold streams are involved.

For stream data of Scheme-1, shown in Table 4.1, hot and cold utilities are found to be 182.73 kW as shown in Fig. 6.9. ΔT_{min} is considered as 50^{o}C due to the fact discussed in Section 6.1.1. Fig. 6.9 shows that 2960.21 kW of internal heat exchange is possible within the process.

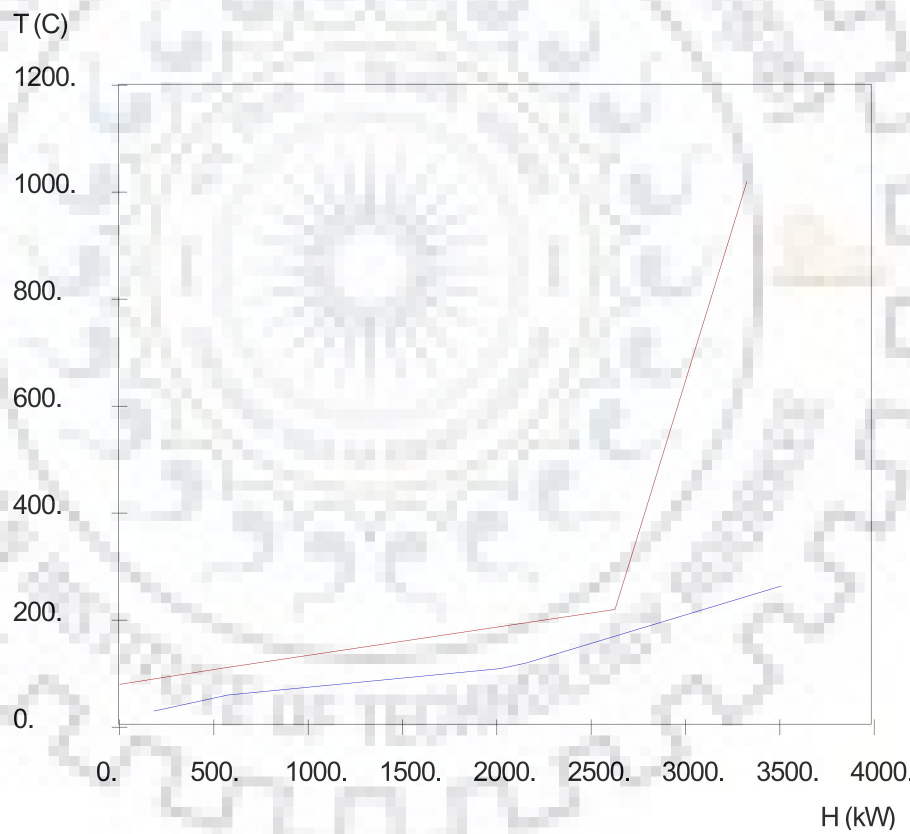


Fig. 6.9 Composite curves of Scheme-1 for Process-2

In Scheme-1 coal and air are preheated using waste gas and at the same time these are used as hot utilities. Due to such dual nature a trial and error computational is carried out as discussed under Section 6.1.1, to estimate amounts of coal and air in this process.

For the existing system shown in Fig. 4.2 coal and air consumptions are 5.81 t/h and 33.23 t/h, respectively. It is called as Iteration-0 as can be observed through Table 6.7. Considering preheating of kiln feed, slinger coal and air up to 120^{O}C, 120^{O}C and 263.8^{O}C, respectively, as shown in Table 4.1, and hot utility as 182.73 kW coal consumption is computed using Equation 4.9 in iteration-1, which is found as 4.86 t/h as shown in stage-1 of iteration-1 in Table 6.7. The corresponding amount of air consumed is computed as 27.82 t/h using ratio of air to coal as maintained in iteration-0, which is 5.72. Thus, for first stage of iteration-1 coal and air are found as 4.86 t/h and 27.82 t/h, respectively. The revised value of air i.e. 27.82 t/h is used in second stage of iteration-1 to compute the amount of coal, which is found to be 4.51 t/h. Corresponding amount of air using ratio of 5.72 is 25.78 t/h. Similarly, different stages of iteration-1 are performed till the values of coal and air become equal in two consecutive stages. The final amounts of coal and air, in iteration-1, are 4.30 t/h and 24.57 t/h, which maintain the ratio of 5.72. The reduction in coal and air consumption is due to heat integration through Scheme-1 to the process in which coal and air are preheated from 30^{O}C to 120^{O}C and from 30^{O}C to 263.8^{O}C, respectively. Therefore the amount of coal required to preheat the coal inside the kiln is reduced. Now, revised values of coal and air i.e. 4.30 t/h and 24.57 t/h are considered to modify the stream data, shown in Table 4.1. For the revised stream data hot utility is found as 85.69 kW as shown in Table 6.7. These revised values are used to compute coal consumption in iteration-2. Following similar procedure for computing coal and air consumptions as carried out in iteration-1, results for other iterations are found in different stages and tabulated in Table 6.7. Final coal and air requirements are 4.54 t/h and 25.98 t/h respectively. Accordingly, 21.83% of coal and air requirements are reduced in comparison to the existing process.

Table 6.7 Iteration results of Scheme-1 for Process-2

Iteration No	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	5,810.0	33,231.0
		1	4,863.1	27,815.0
		2	4,508.0	25,784.1
		3	4,374.9	25,022.5
		4	4,324.9	24,736.9
		5	4,306.2	24,629.8
		6	4,299.2	24,589.7
1	182.73	7	4,296.5	24,574.6
		8	4,295.6	24,569.0
		9	4,295.2	24,566.9
		10	4,295.0	24,566.1
		11	4,295.0	24,565.8
		12	4,295.0	24,565.6
		13	4,295.0	24,565.6
2	85.69		4,292.8	24,553.0
3	85.62		4,292.8	24,553.0
4	85.62		4,292.8	24,553.0
5	555.56		4,525.8	25,885.8
6	585.72		4,540.7	25,971.4
7	587.11		4,541.4	25,975.3
8	587.75		4,541.8	25,977.1
9	587.79		4,541.8	25,977.2
10	587.79		4,541.8	25,977.2

Further, it is observed that usually, iterations to compute coal and air requirements can be stopped once values of coal and air become equal in two consecutive iterations. However, Table 6.7 shows that values of coal and air in iteration-5 are increased in comparison to iteration-4. This is due to the fact that in iteration-4 kiln feed, slinger coal and air are preheated from 30^{O}C to 120^{O}C, 120^{O}C and 250.1^{O}C, respectively. For this purpose, waste gas of 43.61 t/h is used which cools down from 220^{O}C to 80^{O}C. However, HEN, similar to the network shown in Fig. 6.10, indicates that the waste gas can be cooled only from 220^{O}C to 123.5^{O}C instead of from 220^{O}C to 80^{O}C. Due to this the hot utility requirement, computed using pinch analysis [161], increased from 85.62 kW in iteration-4 to 555.56 kW in iteration-5 as shown in Table 6.7. Hence, higher amount of coal is required in iteration-5 than that is used in iteration-4. Consequently, amount of air in iteration-5 also increases to maintain air to coal ratio 5.72. This behavior of coal consumption is also observed while using Scheme-1 in Process-1 and explained in Section 6.1.1. The final HEN of heat integrated Process-2 found through Scheme-1 is shown in Fig. 6.10. Temperatures and heat loads shown in this HEN are based on values of final iteration reported in Table 6.7.

Fig. 6.10 indicates that four additional heat exchangers (HX) are required in the HEN of Process-2 through Scheme-1. Fig. 6.10 shows that waste gas stream (H2) is split into two streams to maintain ΔT_{min} at 50^{O}C in the process. One part of waste gas exchanges heat with slinger coal (C3) to get cooled from 220^{O}C to 196.7^{O}C through HX-2 of load of 70.1 kW. It is a gas-solid exchanger where slinger coal (C3) is preheated from 30^{O}C to 120^{O}C. The same part of waste gas is further cooled to 124^{O}C in HX-4, which has load of 219.2 kW. It is a gas-solid exchanger where kiln feed (C2) is preheated from 30^{O}C to 120^{O}C. The second part of waste gas is cooled from 220^{O}C to 124^{O}C in a gas-gas exchanger, HX-3, having a load of 1026.8 kW. It is required to preheat kiln air up to 170^{O}C (=220^{O}C-50^{O}C) only due to the fact that the maximum temperature of waste gas is 220^{O}C and 50^{O}C temperature difference is required for gas to gas heat transfer. The heating of air from 170^{O}C to 250.1^{O}C requires additional 587.2 kW of heat, as evident from Fig. 6.10, which is supplied through hot utility i.e. coal. Finally, kiln outlet stream (H1) is cooled from 1020^{O}C to 110^{O}C using waste gas (C1) at 60^{O}C in a gas-solid heat exchanger, HX-1, having a heat load of 795.2 kW. Consequently, waste gas is heated up to 125.1^{O}C as shown in Fig. 6.10. Thus, the HEN shown in Fig. 6.10 includes four heat exchangers, in which no heat exchanger violates ΔT_{min}.

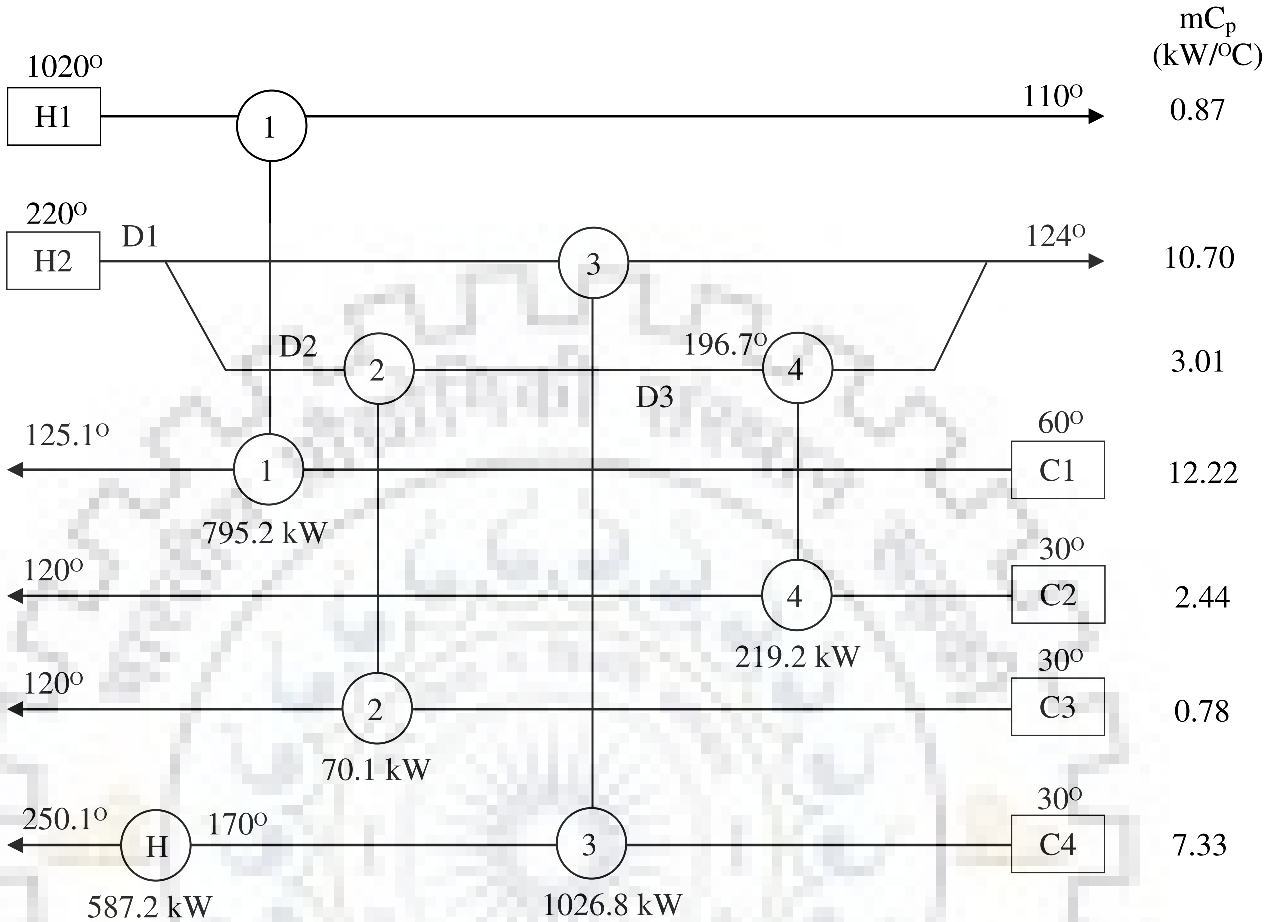


Fig. 6.10 Heat Exchanger Network of Scheme-1 for Process-2

However, in Process-2 internal heat exchange of 2960.21 kW is found through composite curve (Fig 6.9), whereas in HEN, it is only 2112.31 kW. This difference in internal heat exchange is due to variation in flow rates and temperatures of kiln inlet streams (kiln feed, slinger coal and kiln air) and waste gas stream. For example, flow rate of kiln air is 33.23 t/h and 25.98 t/h, before and after modification, respectively. Also, before modification, kiln air is preheated to 263.8°C whereas in HEN shown in Fig. 6.10 it is preheated to only 250.1°C.

In Process-2 and Scheme-1 five ducts are required as: First duct of length 25m is required to bring waste gas from ESP to kiln inlet Second duct of 42m length from kiln inlet to kiln outlet to preheat slinger coal, third duct of length 42m carries waste gas from kiln outlet to kiln inlet to heat the kiln feed, fourth duct of length 67m from kiln inlet to cooler exit. Finally, fifth duct of length 72m is employed to carry waste gas from cooler inlet to chimney. These ducts are clearly shown in modified PFD for Process-2 through Scheme-1 drawn in Fig. 6.11. Further, it indicates that 46.09 t/h of waste gas is discharged, which is 24.5% less as compared to the existing system. Moreover, 93.94% saving in water consumption is also observed in the modified design of Process-2 and Scheme-1.

6.1.3.1 Operational aspects of Process-2 for Scheme-1

Scheme-1 requires five numbers of ducts as discussed in above paragraph. Primarily, an insulated duct (D1) of length 25m and diameter 0.3m is employed to transfer waste gas, having CP of 13.71 kW/^{O}C, from ESP to kiln inlet. While passing through this duct temperature of the waste gas drops down from 220^{O}C to 219.7^{O}C as shown in Fig. 6.11. The temperature profile inside the dust is predicted following the work of Prasad [130]. Once waste gas reaches the kiln inlet at 219.7^{O}C this duct is divided into two branches. First branch of the duct carries waste gas of CP equal to 10.7 kW/^{O}C at 219.7^{O}C where it enters to exchanger, HX-3, shown in Fig. 6.11. Heat transfer area of HX-3 is predicted as 286.6 m^2 considering value of U as 50 W/m^{2}^{O}C [154]. After exchanging heat with kiln air waste gas temperature reduces from 219.7^{O}C to 123.7^{O}C instead of from 220^{O}C to 124^{O}C as shown in Fig. 6.10. Second branch of 25m duct is attached with an insulated duct (D2) of 42m length and 0.3m diameter where waste gas with CP equal to 3.01 kW/^{O}C (=13.71-10.7) enters at 219.7^{O}C and exits at 219.1^{O}C. Further, it enters into the gas-solid exchanger, HX-2 in Fig. 6.11, at 219.1^{O}C, which is in contrast to value shown in Fig. 6.10, where waste gas enters at 220^{O}C. Exchanging 70.1 kW of heat in HX-2 waste gas leaves the exchanger at 195.8^{O}C after transferring its heat to the slinger coal. However, in Fig. 6.10 waste gas exits the exchanger, HX-2, at 196.7^{O}C.

Waste gas exiting exchanger, HX-2, at 195.8^{O}C moves within an insulated duct (D3) of 42 m length and 0.3m diameter. While passing through this duct gas drops its temperature to 195.1^{O}C and then it enters to HX-4. In this exchanger the kiln feed is preheated from 30^{O}C to 120^{O}C through exchanging load of 219.2 kW. Temperature of waste gas reduces to 122.3^{O}C after exchanging heat with kiln feed in HX-4. Once the kiln feed and kiln air are preheated, the waste gas exiting from exchangers HX-3 and HX-4, respectively, are mixed to acquire temperature of 123.4^{O}C. Now, this waste gas is used to cool the kiln outlet. But before that, the temperature of waste gas should bring down to 60^{O}C to maintain the ΔT_{min}. For this purpose an extra gas to gas

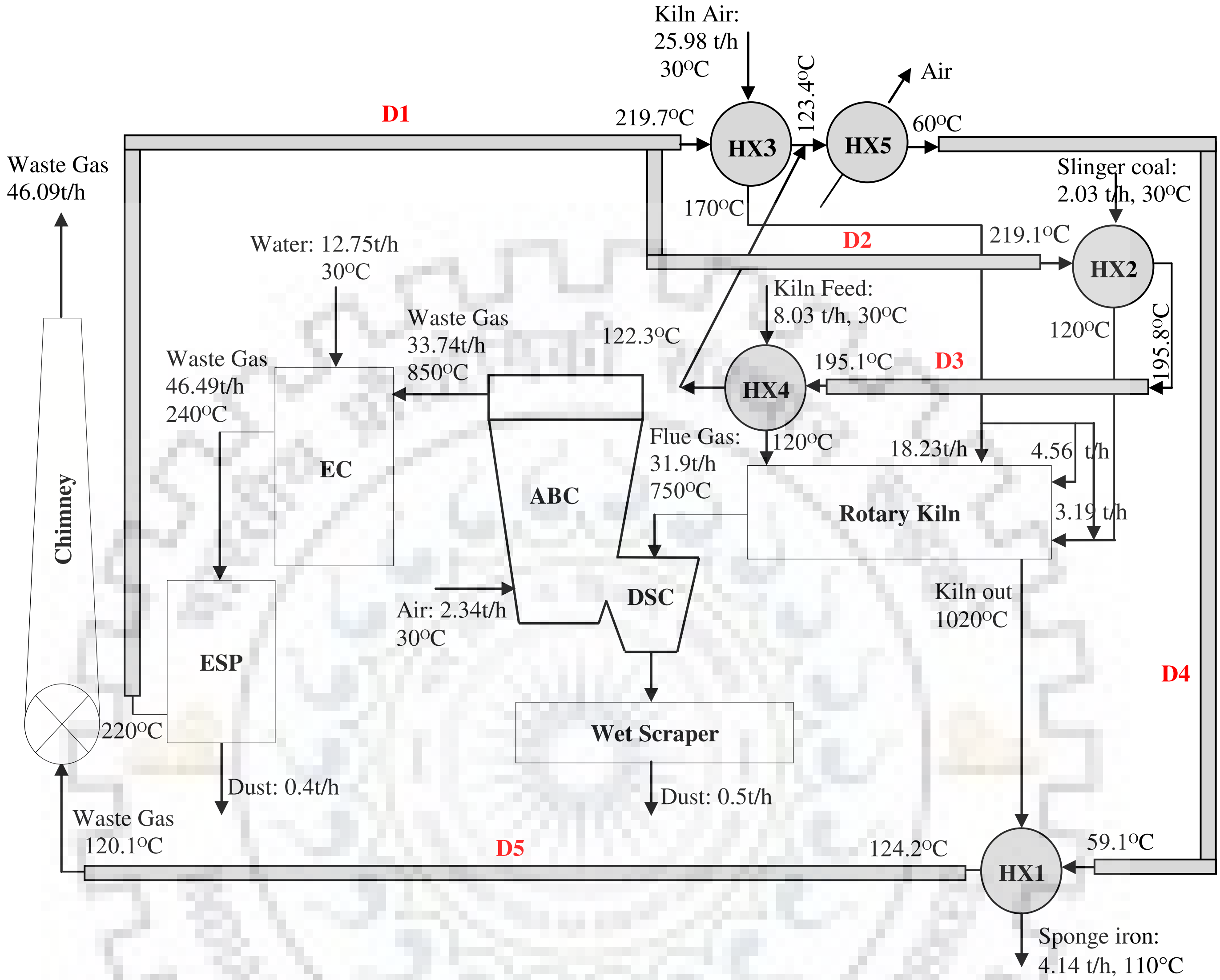


Fig. 6.11 Modified process flow diagram of Process-2 for Scheme-1

Heat exchanger, HX-5, is used taking atmospheric air as the cooling medium. Hence, total of five heat exchangers (three gas-solid and two gas-gas) are required for heat integration in Process-2 through Scheme-1. At 60°C waste gas is passed through a non-insulated duct (D4) of 67m length and 0.3m diameter dropping its temperature to 59.1°C. Then waste gas enters a gas-solid heat exchanger, HX-1, shown in Fig. 6.11, for cooling kiln outlet from 1020°C to 110°C while exchanging heat of 795.2 kW. Consequently, waste gas heats up to 124.2°C. Finally, it moves through non-insulated duct (D5) of 72m length and 0.3m diameter, which directs waste gas to the chimney. The temperature profiles of all five ducts are shown in Fig. 6.12. The methodology for computing temperature profile is taken from the work of Prasad [130]. For all dusts mild steel is used as material of construction and glass wool as insulating material, which is placed at outer surface of the duct.

Further, there is a certain pressure drop while waste gas moves through these five ducts, which is estimated using method discussed by Prasad [133] and tabulated in Table 6.8. These pressure drops have to be compensated through FD fans. It is observed that one FD can sustain 1.26 atm pressure drop [177]. Accordingly, five FD fans are required, one for each duct, for modification suggested in Process-2 through Scheme-1.

Table 6.8 Pressure and temperature drops in ducts for Process-2 Scheme-1

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	25	ESP exit gas	ESP	Kiln inlet	0.091	220	219.7
D2	42	ESP exit gas	Kiln inlet	Kiln outlet	0.009	219.7	219.1
D3	42	ESP exit gas	Kiln outlet	Kiln inlet	0.009	195.8	195.1
D4	67	ESP exit gas	Kiln inlet	RC exit	0.165	60	59.1
D5	72	ESP exit gas	RC inlet	Chimney	0.212	124.2	120.1

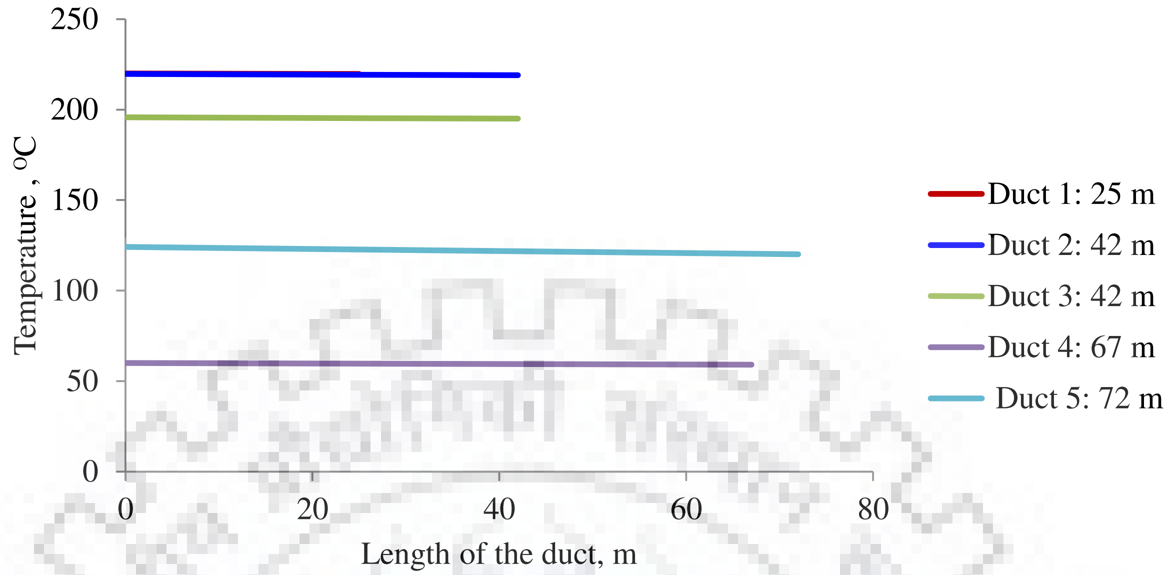


Fig. 6.12 Temperature profile of waste gas in ducts for Process-2 Scheme-1

6.1.3.2 Economic aspects of Process-2 for Scheme-1

The economic analysis of Scheme-1 in terms of operating cost, capital investment, total annual cost (TAC), savings and payback period is discussed here. Capital costs for three gas-solid heat exchangers, two gas-gas shell and tube heat exchangers, five ducts of total length 248m, five FD fans are found as 726.96 lakh as shown in Table 6.9. The costing method of gas-solid heat exchangers, ducts and FD fans is considered from the work of Prasad [133] and shown in Appendix C. Savings through coal and water consumptions are predicted as Rs. 2853.5/h and Rs 10669.39/h, respectively. Further, electricity cost for five fans is predicted as Rs 0.34 lakh/day. The net savings in terms of coal, water and electricity is found to be Rs 2.97 lakh/day. Based on total capital cost of new equipment and savings the discounted payback period of 9.05 months is predicted. If capital recovery factor is considered with a plant life of 10 years and 10% rate of return, TAC of the plant comes out to be Rs 1198.0 lakh. The total economic analysis for Process-2 through Scheme-1 is shown in Table 6.9.

Table 6.9 Economic analysis of Scheme-1 for Process-2

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	895.18	HX-1	104.72	1198.0	1060.43	5788.9	123.5	9.05
Water	60.33	HX-2	64.4					
Electricity	124.17	HX-3	47.13					
		HX-4	72.69					
		HX-5	60.02					
		Ducts	25.22					
		FD fans	352.78					
		Total	726.96					

6.1.4 Process-2 and Scheme-2

Scheme-2 is discussed in detail under Section 4.3.2 and stream data for this scheme are shown in Table 4.2. In this scheme, waste gas exiting ABC at 850°C is used for heat integration. It preheats kiln feed and slinger coal from 30°C to 300°C. Kiln outlet stream is cooled from 1020°C to 110°C using kiln air, which is further preheated to 300°C using ABC exit gas. Therefore, Scheme-2 includes two hot and three cold streams.

For stream data of Scheme-2 for Process-2, shown in Table 4.2, hot and cold utilities are found as zero. It is a threshold problem as the hot and cold utility requirements are zero [161]. It can be depicted through composite curves of the scheme presented in Fig. 6.13. It shows that 3543.8 kW of heat can be recovered within the process, which is called internal heat exchange.

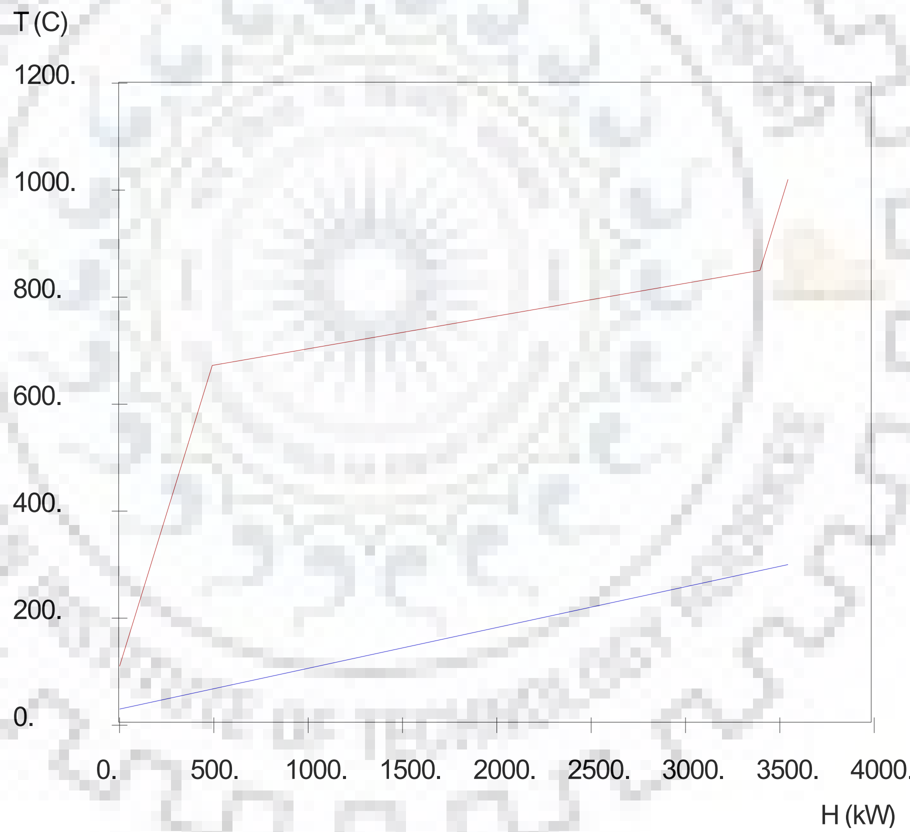


Fig. 6.13 Composite curves for Process-2 Scheme-2

Coal and air consumptions at Iteration-0 (existing system) are 5.81 t/h and 33.23 t/h, respectively, as shown in Fig. 4.2 and Table 6.10. Considering preheating of kiln feed, slinger coal and air up to 300°C as shown in Table 4.2, and hot utility as 0 kW coal consumption is computed using

Equation 4.9 in iteration-1, which is found as 4.53 t/h as shown in stage-1 of iteration-1. Corresponding amount of air is found to be 25.92 t/h using ratio of air to coal as 5.72. This revised value of air i.e. 25.92 t/h is used in second stage of iteration-1 to compute the amount of coal, which is found to be 4.08 t/h. As a result, air requirement is obtained as 23.36 t/h. In the similar line different stages of iteration-1 and shown in Table 6.10. It shows that in iteration-1 final amounts of coal and air are 3.84 t/h and 21.97 t/h, respectively. Following similar procedure results for other iterations are found and tabulated in Table 6.10. Final coal and air requirements are 3.84 t/h and 21.97 t/h, respectively. Therefore, in Process-2 and Scheme-2, 33.88 % reduction of coal is found in comparison to the existing process.

Table 6.10 Iteration results of Scheme-2 for Process-2

Iteration No	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	5,810.0	33,231.0
1	0	1	4,531.7	25,919.6
		2	4,083.4	23,355.6
		3	3,926.2	22,456.4
		4	3,871.1	22,141.1
		5	3,851.7	22,030.5
		6	3,845.0	21,991.7
		7	3,842.6	21,978.1
		8	3,841.8	21,973.4
		9	3,841.5	21,971.7
		10	3,841.4	21,971.1
		11	3,841.3	21,970.9
		12	3,841.3	21,970.8
		13	3,841.3	21,970.8
2	0		3,841.3	21,970.8

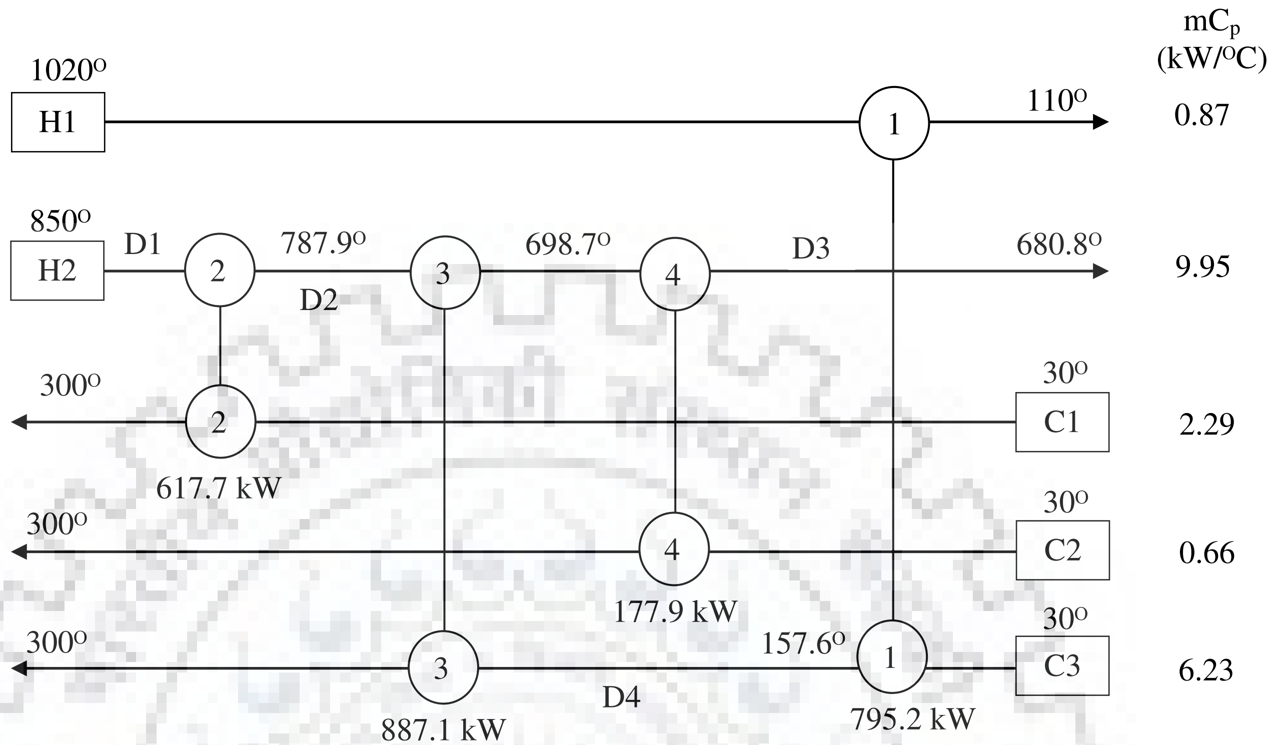


Fig. 6.14 Heat Exchanger Network of Scheme-2 for Process-2

The HEN of Process-2 for Scheme-2 is shown in Fig. 6.14, which involves four new heat exchangers. Temperatures and heat loads shown in Fig. 6.14 are based on results of final iteration reported in Table 6.10. The exchanger, HX-1, is a gas-solid heat exchanger, which exchanges 795.2 kW heat between kiln outlet stream (H1) and kiln air (C3). As a result, kiln outlet stream cools from 1020°C to 110°C whereas kiln air heats up from 30°C to 157.6°C. Second heat exchanger, HX-2, is a gas-solid heat exchanger having a heat load of 617.7 kW. It preheats kiln feed (C1) from 30°C to 300°C using waste gas stream (H2), which exits ABC at 850°C. While exchanging heat in HX-2 waste gas temperature drops to 787.9°C. Further, it enters to HX-3, which is gas-gas heat exchanger with heat load of 887.1 kW. Kiln air (C3) is further preheated from 157.6°C to 300°C in this exchanger using waste gas available at 787.9°C and subsequently, its temperature drops to 698.7°C. At this temperature waste gas goes through a gas-solid heat exchanger, HX-4, of 177.9 kW heat load where slinger coal (C2) is preheated from 30°C to 300°C. Consequently, waste gas temperature drops to 680.8°C as shown in the Fig. 6.14. It shows that no heat exchanger violates ΔT_{min}.

For Process-2 internal heat exchange, shown in the composite curve (Fig 6.13), before modification through Scheme-2 i.e. 3543.8 kW is different than that shown in HEN, after modification, which is 2477.9 kW. This difference in internal heat exchange is due to the change in flow rates and temperatures of kiln inlet streams. For example, before and after modification flow rate of waste gas exiting ABC is 44.6 t/h and 28.7 t/h, respectively. Along with this, before modification, recovery of waste heat from ABC exit gas is targeted while cooling it from 850^{O}C to 672.3^{O}C whereas in HEN it is cooled to only 680.8^{O}C as shown in Fig. 6.14.

In the Scheme-2 Process-2 three ducts are required for carrying waste gas from ABC to different places where it exchanges heat before it is directed back to ABC and one duct is required for carrying kiln air, hence, total of four ducts. Initially, waste gas is required to bring from ABC to kiln inlet, which needs a duct of length 12m. Further, following ducts are used: Second duct of 42m length from kiln inlet to kiln outlet to preheat kiln air & slinger coal and third duct of length 54m to carry back waste gas from kiln outlet to ABC exit. Finally, a duct of length 7m is employed to carry kiln air from cooler inlet to kiln outlet. The modified PFD for Process-2 through Scheme-2 is shown in Fig. 6.15. The reduced amounts of coal and air cause less waste gas to discharge from the process. The waste gas generation, in existing system as shown in Fig. 4.2, is 61.05 t/h, which is reduced to 33.99 t/h in the modified design shown in Fig. 6.15. Thus, 44.32% less waste gas is generated in Process-2 which is an additional benefit of heat integration through Scheme-2. Similarly, 97.28% of water consumption is reduced in the modified system in comparison to the existing PFD.

6.1.4.1 Operational aspects of Process-2 for Scheme-2

Initially, an insulated duct (D1) of length 12 m and diameter 0.3m is used to transfer waste gas, with CP of 9.95 kW/^{O}C, from ABC to kiln inlet. Consequently, temperature of waste gas decreases from 850^{O}C to 848.8^{O}C as shown in the Fig. 6.15. Once the waste gas reaches the kiln inlet at 848.8^{O}C it enters into the gas-solid heat exchanger, HX-2, shown in Fig. 6.15. After exchanging heat with kiln feed waste gas temperature drops from 848.8^{O}C to 786.7^{O}C instead of from 850^{O}C to 787.9^{O}C as shown in Fig. 6.14. Waste gas exiting exchanger, HX-2, at 786.7^{O}C moves through an insulated duct (D2) of 42 m length and 0.3m diameter and further, drops its temperature to 783.1^{O}C. Thus, waste gas enters at 783.1^{O}C to the gas-gas exchanger, HX-3. In this exchanger kiln air is preheated from 157.4^{O}C to 299.8^{O}C through exchanging load of 887.1 kW instead of from 220.5^{O}C to 300^{O}C as shown in Fig. 6.14.

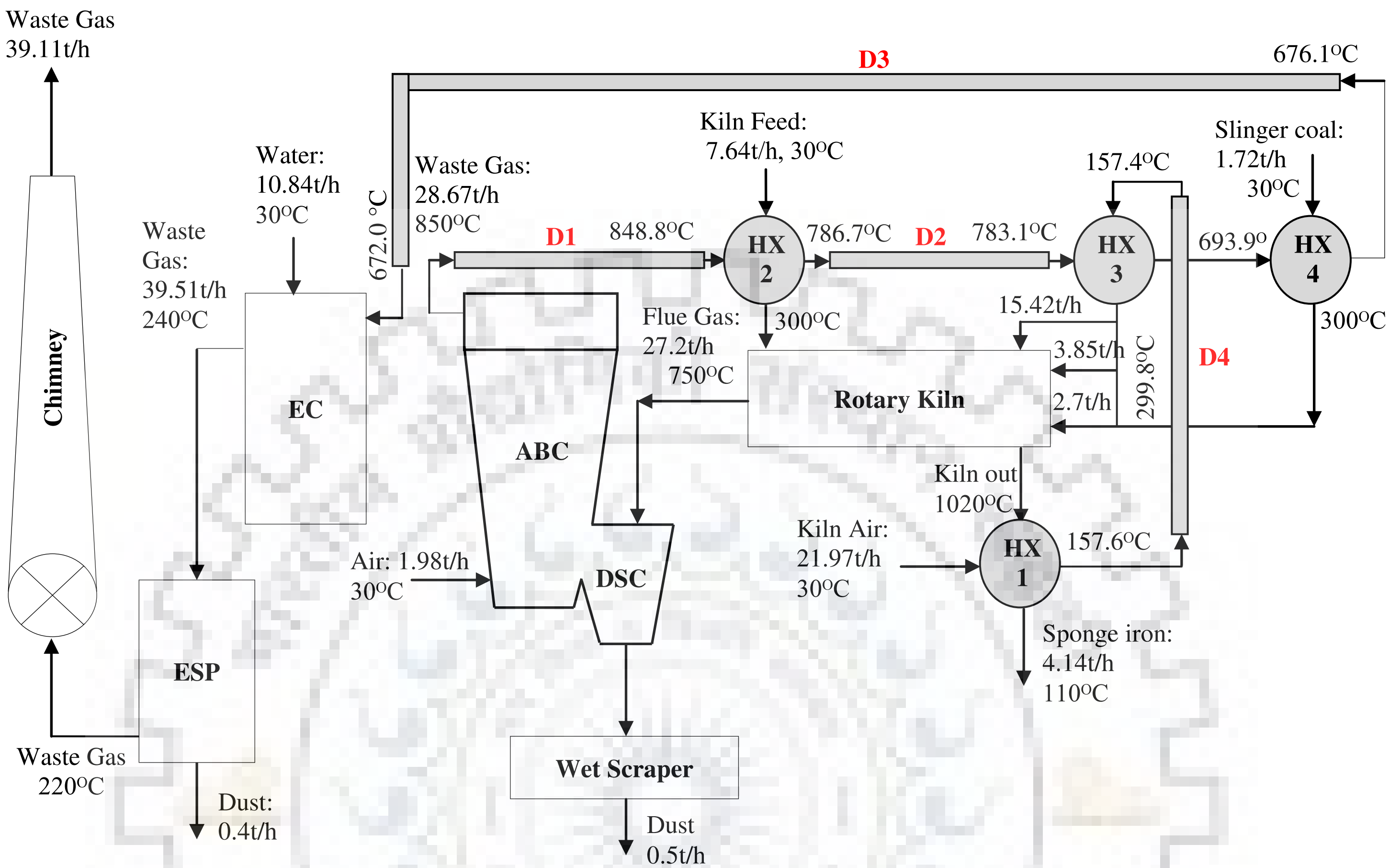


Fig. 6.15 Modified process flow diagram of Process-2 for Scheme-2

Further, waste gas enters into gas-solid exchanger, HX-4, shown in Fig. 6.15 at 693.9°C. However, this temperature is 698.7°C in Fig. 6.14. Exchanging 177.9 kW of heat in HX-4 waste gas leaves the exchanger at 676.1°C while transferring its heat to the slinger coal. However, in Fig. 6.14 waste gas exits the exchanger, HX-4, at 680.8°C. An insulated duct (D4) of length 7m and diameter 0.3m is required to carry kiln air that was already preheated in gas-solid heat exchanger, HX-1, from cooler inlet to kiln outlet. While passing through this duct kiln air drops its temperature from 157.6°C to 157.4°C as shown in Fig. 6.15. Finally, a non-insulated duct (D3) of 54m length and 0.3m diameter is used to carry waste gas at 676.1°C to ABC where temperature of gas drops to 672°C. The temperature profiles of all four ducts are shown in Fig. 6.16, which are drawn using the method proposed by Prasad [130]. Mild steel is used for casing of the duct and for insulation ceramic wool is placed at outer surface of the duct.

As gases pass through these four ducts certain pressure drops are created in these, which are estimated and tabulated in Table 6.11. The methodology for calculating pressure drops, similar to Scheme-1, is taken from the work of Prasad [133]. FD fans are used to compensate the pressure drop created in ducts. In fact, one FD can sustain 1.26 atm pressure drop [177] and thus, total four FD fans are required, one for each duct for Scheme-2.

Table 6.11 Pressure and temperature drops in ducts for Process-2 Scheme-2

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	12	ABC exit gas	ABC exit	Kiln inlet	0.047	850.0	848.8
D2	42	ABC exit gas	Kiln inlet	Kiln outlet	0.153	786.7	783.1
D3	54	ABC exit gas	Kiln outlet	ABC exit	0.165	676.1	672.0
D4	7	Kiln air	RC inlet	Kiln outlet	0.006	157.6	157.4

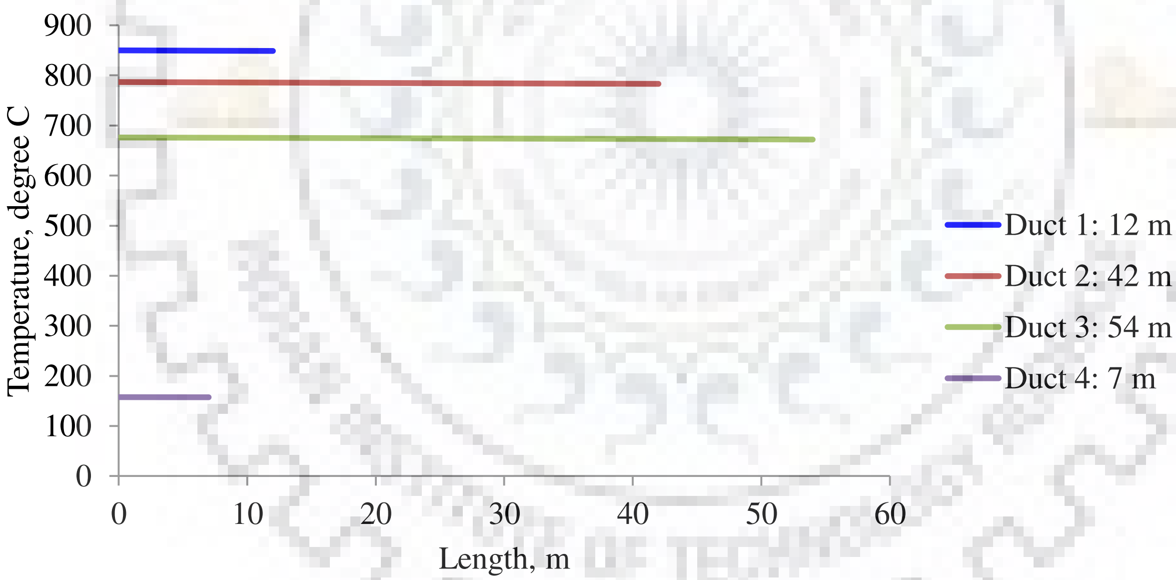


Fig. 6.16 Temperature profile of waste gas in ducts for Process-2 Scheme-2

6.1.4.2 Economic aspects of Process-2 for Scheme-2

The economic analysis of Scheme-2, similar to Scheme-1, in terms of operating cost, capital investment, total annual cost (TAC), savings and payback period is carried out and results are reported in Table 6.12. Total capital investment of Rs. 657.5 lakh is found for four heat exchangers, four ducts of total length 115m and four FD fans. The total coal consumption, through Scheme-2, reduces by 33.88% in comparison to the existing system, which saves Rs. 4429.6/h. Similarly, Rs 11049.4/h is saved through reduction of water consumption by 97.28%. Further, electricity cost for four fans is predicted as Rs 0.27 lakh/day. Thus, net savings in terms of coal, water and electricity is found to be Rs 3.44 lakh/day. For this modification discounted payback period of 6.91 months is predicted. Total economic analysis for Process-2 and Scheme-2 is computed as described in Appendix C.

Table 6.12 Economic analysis of Scheme-2 for Process-2

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	757.12	HX-1	104.72	990.5	1256.62	7063.9	164.7	6.91
Water	27.04	HX-2	94.85					
Electricity	99.34	HX-3	19.74					
		HX-4	70.39					
		Ducts	28.75					
		FD fans	339.05					
		Total	657.50					

6.1.5 Process-3 and Scheme-1

In this section detailed analysis of results of heat integration in Process-3 through Scheme-1 is discussed. Process-3 is a typical coal based sponge iron plant of 100 t/d capacity. PFD of Process-3 with material and energy balanced data is shown in Fig. 4.3 and its detailed process description is given in Section 3.1. Scheme-1 for Process-3 is formulated in the similar lines as carried out for Process-1 and Process-2 and corresponding stream data are reported in Table 4.1. It shows that the present scheme contains two hot and four cold streams. Hot and cold utility requirements are found to be zero using pinch analysis [161] and considering ΔT_{min} as $50^{o}C$. It can be depicted through composite curves shown in Fig. 6.17, which indicates that 2948.4 kW of internal heat exchange is possible within the process.

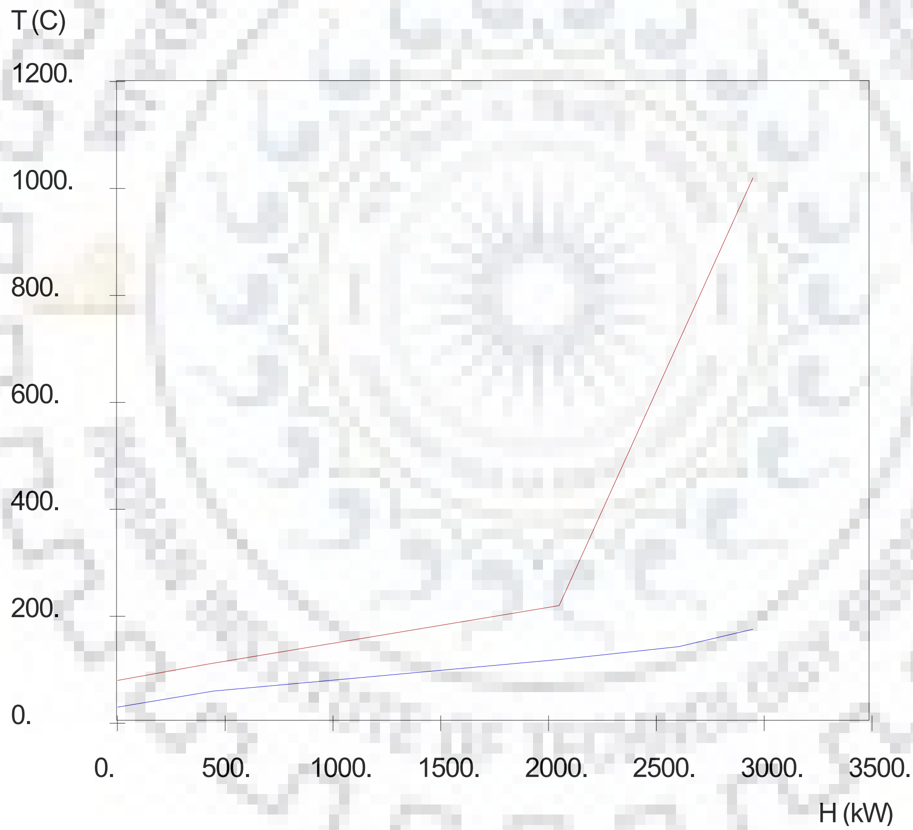


Fig. 6.17 Composite curves of Scheme-1 for Process-3

A trial and error computational technique is required, as discussed under Section 6.1.1, to estimate the amount of coal and air in this process, which is also shown in Fig. 5.1. Results of these iterations are shown in Table 6.13, which are explained in similar manner as carried out for

Process-1 and Process-2 in Section 6.1.1 and 6.1.3, respectively. Table 6.13 shows that final coal and air requirements are 4.95 t/h and 28.64 t/h respectively. Therefore, 23.85% of coal and air requirements are reduced in comparison to the existing process.

Table 6.13 Iteration results of Scheme-1 for Process-3

Iteration No	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	6,500.0	37,603.5
1	0	1	5,608.7	32,446.9
		2	5,221.1	30,205.0
		3	5,052.7	29,230.4
		4	4,979.4	28,806.6
		5	4,947.6	28,622.4
		6	4,933.7	28,542.3
		7	4,927.7	28,507.4
		8	4,925.1	28,492.3
		9	4,923.9	28,485.7
		10	4,923.4	28,482.9
		11	4,923.2	28,481.6
		12	4,923.1	28,481.1
		13	4,923.1	28,480.8
		14	4,923.1	28,480.7
		15	4,923.1	28,480.7
2	0		4,950.6	28,640.2
3	0		4,950.0	28,636.5
4	0		4,950.0	28,636.6
5	0		4,950.0	28,636.6

One salient feature can be observed, for Process-3 Scheme-1, through iteration-2 of Table 6.13. Values of coal and air in iteration-2 are slightly increased in comparison to iteration-1. This is due to the fact that in iteration-1 kiln feed, slinger coal and air are preheated from $30^{o}C$ to $120^{o}C$, 120°C and 175.8°C, respectively. For this purpose, waste gas of 46.45 t/h cools down from 220°C to 80°C. Due to this preheating using waste gas, amounts of coal and air are reduced as shown in Table 6.13. Further, revised values of coal and air are used for material balance. Consequently, amount of waste gas exiting from ESP is also reduced to 34.93 t/h, which is cooled from 220°C to 80°C in iteration-2. Thus, overall heat available for preheating is reduced in iteration-2 than that was for iteration-1. Now, it preheats kiln feed, slinger coal and air from 30°C to 120°C, 120°C and 169.7°C, respectively. It shows that in iteration-2 temperature of air after preheating is less in comparison to iteration-1. Thus, air to be preheated inside the kiln is from 169.7°C to reaction temperature in iteration-2 instead of from 175.8°C to reaction temperature in iteration-1. Therefore, preheating of air requires slightly higher amount of coal in iteration-2 than that is used in iteration-1. Consequently, amount of air in iteration-2 also increases slightly to maintain air to coal ratio 5.79.

The final HEN of heat integrated Process-3 found through Scheme-1 is shown in Fig. 6.18 where temperatures and heat loads are based on values of final iteration reported in Table 6.13. Fig. 6.18 indicates that four new HX are required in the HEN as additional equipment in the modified design of Process-3 through Scheme-1. In this HEN HX-1, HX-2 and HX-4 are gas-solid heat exchangers whereas; HX-3 is gas-gas exchanger. The network, Fig. 6.18, is similar to that shown in Fig. 6.2 and Fig. 6.10 and can be explained in similar lines. It is observed that no heat exchanger violates ΔT_{min} in HEN shown in Fig. 6.18.

However, in Process-3 the internal heat exchange, shown in composite curve (Fig. 6.17) and HEN (Fig. 6.18) are 2948.4 kW and 2478.6 kW, respectively. This difference in internal heat exchange is due to the change in flow rates and temperatures of kiln inlet streams (kiln feed, slinger coal and kiln air) and waste gas stream as explained in Section 6.1.1 and 6.1.3.

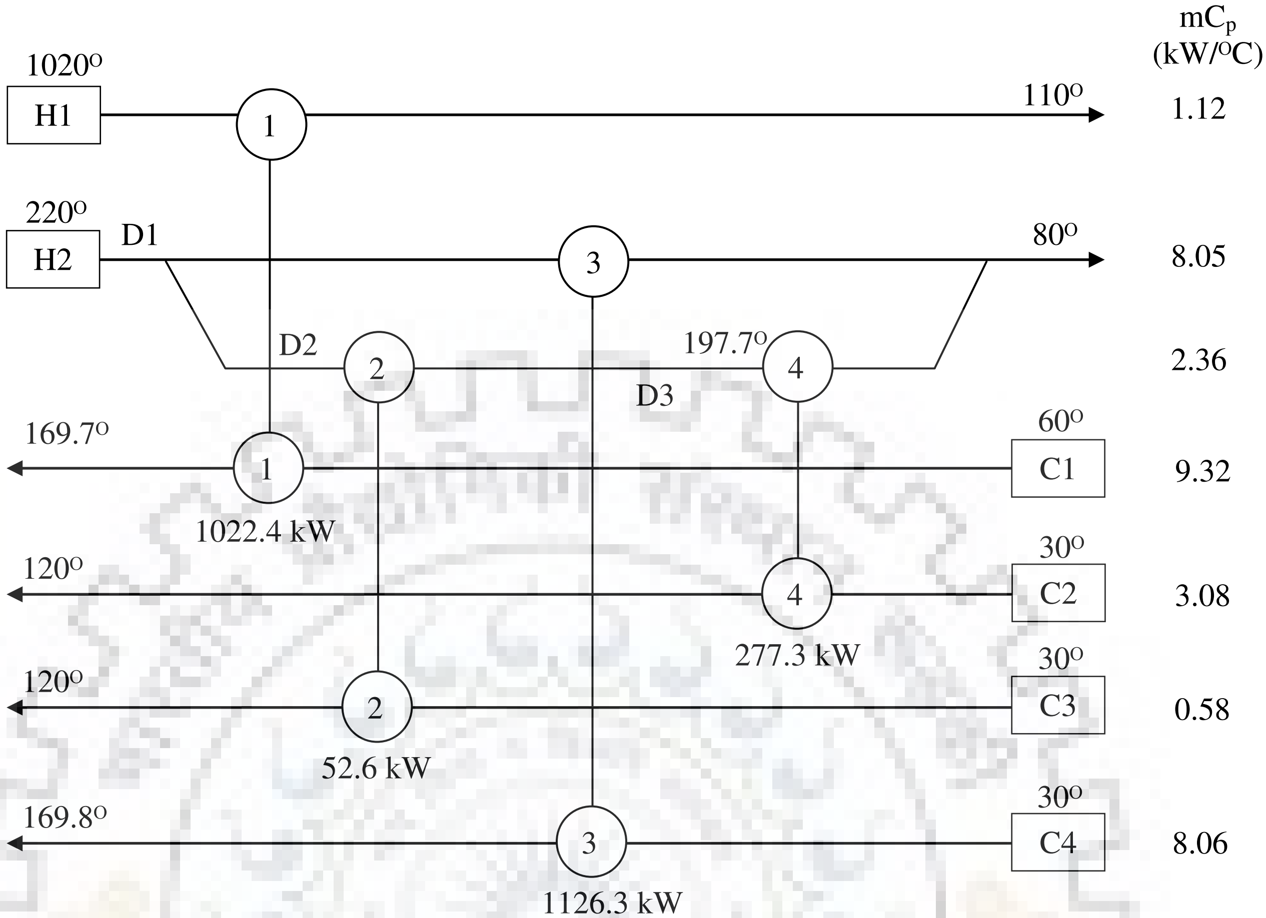


Fig. 6.18 Heat Exchanger Network of Scheme-1 for Process-3

Following ducts are used in Scheme-1 for carrying waste gas from ESP to different places where it exchanges heat before it is directed to the chimney. First duct of length 25m from ESP to kiln inlet, second duct of 40m length from kiln inlet to kiln outlet to preheat slinger coal, third duct of length 40m carries waste gas from kiln outlet to kiln inlet to heat the kiln feed, fourth duct of length 65m from kiln inlet to cooler exit. Finally, a duct of length 70m is employed to carry waste gas from cooler inlet to chimney. The modified PFD for Process-3 through Scheme-1 is shown in Fig. 6.19. The reduced amounts of coal and air cause less waste gas to discharge from the process. Due to suggested modification in Process-3 through Scheme-1, 24.4% less waste gas is generated which is an additional benefit of heat integration. Similarly, 96% savings in water consumption is also found.

6.1.5.1 Operational aspects of Process-3 for Scheme-1

Scheme-1 requires five numbers of ducts as discussed above. Initially, an insulated duct (D1) of length 25m and diameter 0.3m is employed to transfer waste gas, having CP of 10.41 kW/^{O}C, from ESP to kiln inlet. While passing through this duct temperature of waste gas drops down from 220^{O}C to 219.7^{O}C as shown in Fig. 6.19. The methodology for calculating temperature profile is taken from the work of Prasad [130]. Once waste gas reaches the kiln inlet at 219.7^{O}C this duct is divided into two branches. First branch of the duct carries waste gas of CP equal to 8.05 kW/^{O}C at 219.7^{O}C. At this temperature waste gas enters into a shell and tube gas-gas heat exchanger, HX-3, shown in Fig. 6.19. Considering U value as 50 W/m^{2O}C [154] heat transfer area of HX-3 is predicted as 454.2 m^2. After exchanging heat with kiln air waste gas temperature drops from 219.7^{O}C to 79.7^{O}C instead of from 220^{O}C to 80^{O}C as shown in Fig. 6.18. Second branch of 25m duct is attached with an insulated duct (D2) of 40m length and 0.3m diameter. It carries the remaining value of CP of waste gas i.e. (10.41-8.05) 2.36 kW/^{O}C. Waste gas enters into this duct at 219.7^{O}C and its temperature is dropped to 218.8^{O}C while traversing through the duct. Further, it enters into gas-solid exchanger, HX-2 in Fig. 6.19, at 218.8^{O}C, which is in contrast to value shown in Fig. 6.18, where waste gas enters at 220^{O}C. Exchanging 52.6 kW of heat in HX-2 waste gas leaves the exchanger at 196.5^{O}C after transferring its heat to the slinger coal. However, in Fig. 6.18 waste gas exits the exchanger, HX-2, at 197.7^{O}C.

Waste gas exiting exchanger, HX-2, at 196.5^{O}C passes through an insulated duct (D3) of 40 m length and 0.3m diameter. While passing through this duct gas drops its temperature to 195.6^{O}C. Thus, waste gas enters at 195.6^{O}C to the gas-solid exchanger, HX-4 where kiln feed is preheated from 30^{O}C to 120^{O}C through exchanging load of 277.3 kW. The temperature of the waste gas reduces to 77.9^{O}C after exchanging the heat with kiln feed in HX-4. Once the kiln feed and kiln air are preheated, the waste gas exiting from exchangers HX-3 and HX-4, respectively, are mixed to acquire temperature of 79.3^{O}C. Further, temperature of waste gas should bring down to 60^{O}C to maintain the ΔT_{min} before it is used to cool kiln outlet. For this purpose an extra gas to gas heat exchanger, HX-5, is used considering air as cooling medium. Hence, total of five heat exchangers (three gas-solid and two gas-gas) are required for heat integration through Scheme-1. After attaining 60^{O}C in HX-5 waste gas is passed through a non-insulated duct (D4) of 65m length and 0.3m diameter dropping its temperature to 58.8^{O}C. Then waste gas enters a gas-solid heat exchanger, HX-1, shown in Fig. 6.19, at 58.8^{O}C for cooling the kiln outlet from 1020^{O}C to 110^{O}C while exchanging heat of 1022.4 kW. After this exchange of heat the waste gas heats up to 169.8°C. Finally, after preheating C2, C3 and C4 streams and cooling H1 and H2 streams as shown in Table 4.1 the waste gas (C1) moves through an non-insulated duct (D5) of 70m length and 0.3m diameter. It carries waste gas to the chimney and while passing through this duct the temperature of gas drops to 160.2°C. Thus, waste gas enters the chimney at 160.2°C. The temperature profiles of all five ducts are shown in Fig. 6.20. Mild steel is used for casing of the duct and for insulation glass wool is placed at outer surface of the duct.

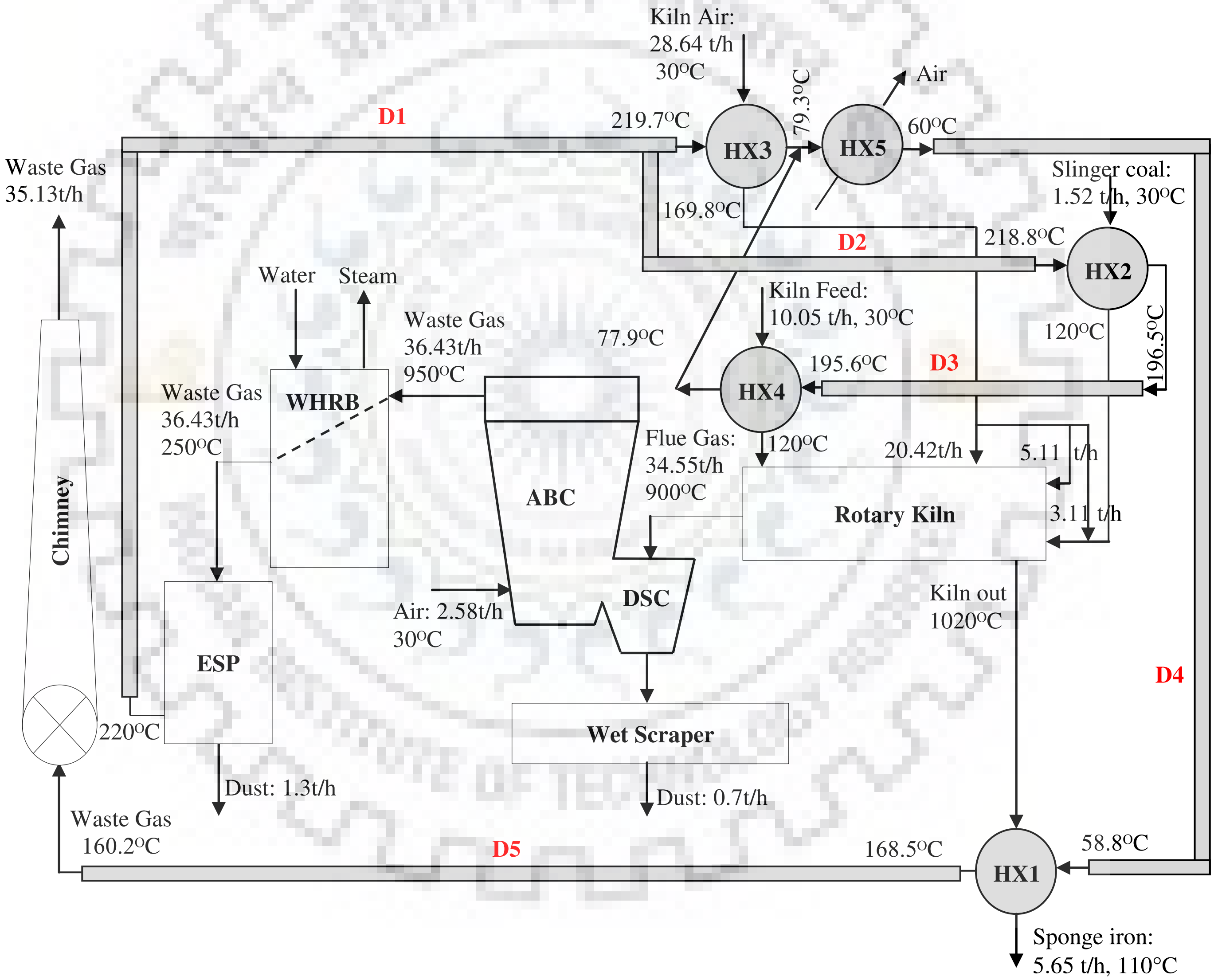


Fig. 6.19 Modified process flow diagram of Process-3 for Scheme-1

Further, as in Process-1 and Process-2, there are certain pressure drops when waste gas moves through these five ducts, which are estimated using method discussed by Prasad [133] and tabulated in Table 6.14. These pressure drops are compensated using five FD fans, one for each duct, as one FD can sustain 1.26 atm pressure drop [177].

Table 6.14 Pressure and temperature drops in ducts for Process-3 Scheme-1

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	25	ESP exit gas	ESP	Kiln inlet	0.055	220	219.7
D2	40	ESP exit gas	Kiln inlet	Kiln outlet	0.006	219.7	218.8
D3	40	ESP exit gas	Kiln outlet	Kiln inlet	0.006	196.5	195.6
D4	65	ESP exit gas	Kiln inlet	RC exit	0.097	60	58.8
D5	70	ESP exit gas	RC inlet	Chimney	0.135	168.5	160.2

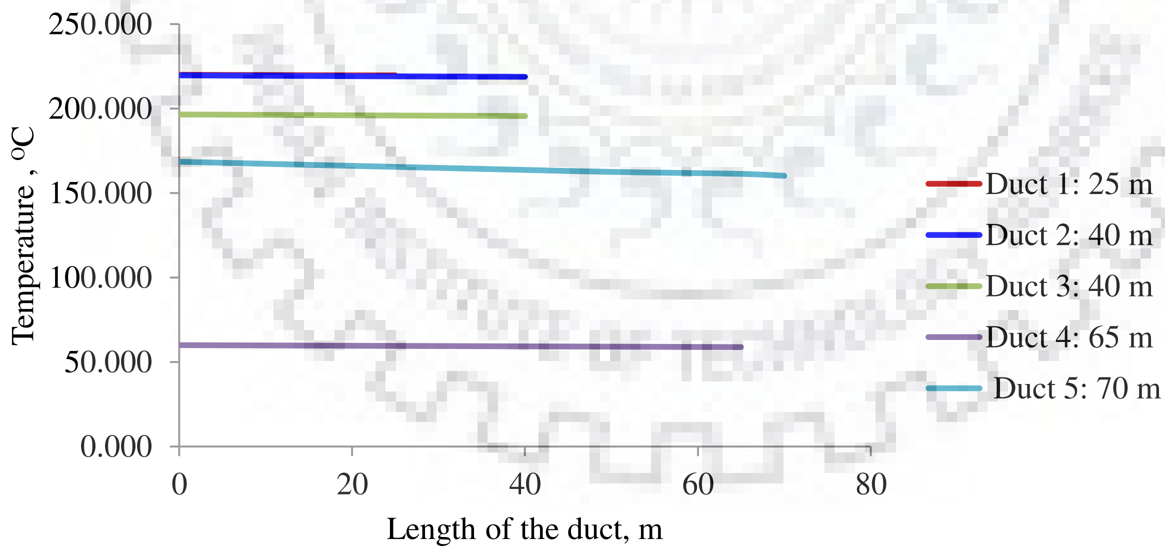


Fig. 6.20 Temperature profile of waste gas in ducts for Process-3 Scheme-1

6.1.5.2 Economic aspects of Process-3 for Scheme-1

The economic analysis of Process-3 and Scheme-1 in terms of operating cost, capital investment, TAC, savings and payback period is carried out as discussed in Sections 6.1.1.2 and 6.1.3.2. Results of economic analysis are summarized in Table 6.15, which shows 7.48 months of discounted payback period for the suggested modification in Process-3 using Scheme-1.

Table 6.15 Economic analysis of Scheme-1 for Process-3

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	975.65	HX-1	117.35	1265.8	1253.52	6991.8	151.3	7.48
Water	50.35	HX-2	63.42					
Electricity	124.17	HX-3	62.25					
		HX-4	75.92					
		HX-5	31.89					
		Ducts	25.14					
		FD fans	334.54					
		Total	710.51					

6.1.6 Process-3 and Scheme-2

Scheme-2 is discussed in detail under Section 4.3.2. For this scheme, kiln feed and slinger coal are preheated from 30^{O}C to 300^{O}C using ABC exit gas at 950^{O}C and kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air which is further preheated, similar to Process-1 and Process-2, to 300^{O}C by ABC exit gas. Therefore, stream data for Scheme-2 shown in Table 4.2 contains two hot and three cold streams and corresponding hot and cold utilities are found to be zero as computed for Process-1 and Process-2. This can be depicted through composite curves shown in Fig. 6.21 which also indicates 4029.21 kW as possible internal heat exchange between hot and cold streams.

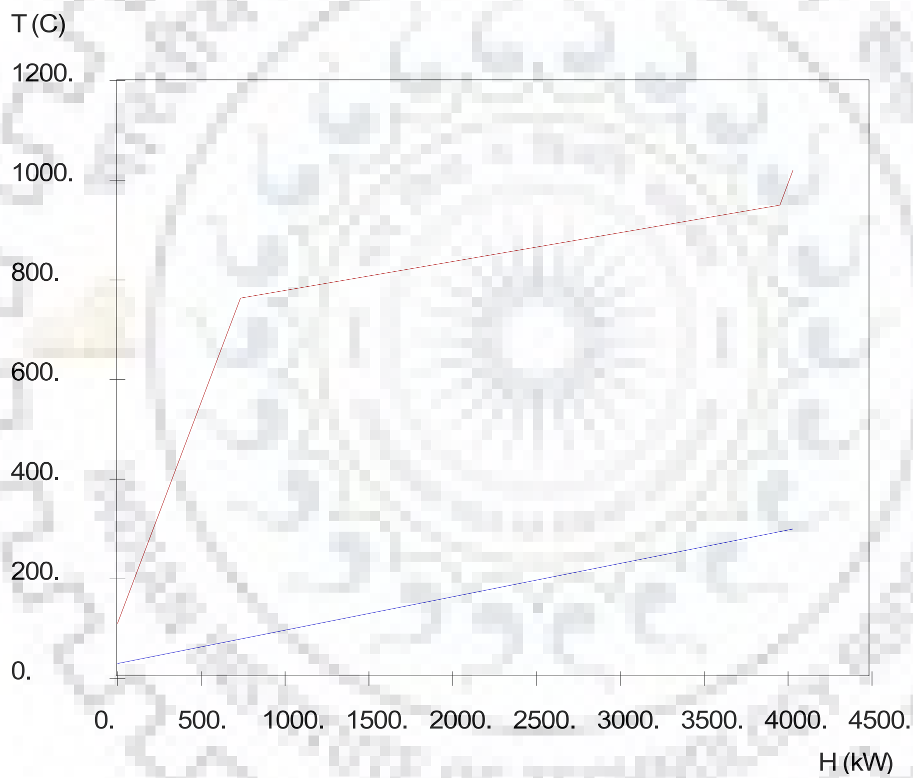


Fig 6.21 Composite curves of Scheme-2 for Process-3

A trial and error computational technique, as used for Process-1 and Process-2, is applied to estimate amounts of coal and air in this process. The coal and air consumption at Iteration-0 (existing system) are 5.81 t/h and 33.231 t/h, respectively, as shown in Fig. 4.3 and Table 6.16. Following similar procedure for computing coal and air requirements, as carried out for other two processes; results of all iterations are computed and tabulated in Table 6.16. In all the iterations air to coal ratio is maintained as that of the existing system, i.e., 5.79. The final coal and air requirements are 4.14 t/h and 23.94 t/h respectively, which shows 36.33% reduction in coal consumption in comparison to the existing process.

Table 6.16 Iteration results of Scheme-2 for Process-3

Iteration No	Stage No.	Coal consumption (kg/h)	air requirement (kg/h)
0 (existing)	0	6,500.0	37,603.5
1	1	5,002.4	28,939.4
	2	4,454.5	25,770.1
	3	4,254.1	24,610.8
	4	4,180.8	24,186.8
	5	4,154.0	24,031.7
	6	4,144.2	23,974.9
	7	4,140.6	23,954.2
	8	4,139.3	23,946.6
	9	4,138.8	23,943.8
	10	4,138.7	23,942.8
	11	4,138.6	23,942.4
	12	4,138.6	23,942.3
	13	4,138.6	23,942.2
	14	4,138.6	23,942.2
2		4,138.6	23,942.2

The final HEN of heat integrated Process-3 found through Scheme-2 is shown in Fig. 6.22 in which temperatures and heat loads are based on final iteration values as reported in Table 6.16.

Modified design of Process-3 requires four new HXs as an additional equipment for energy conservation through Scheme-2 as shown in Fig. 6.22. HX-1, HX-2 and HX-4 are gas-solid heat exchangers having heat loads of 1022.4 kW, 773.8 kW and 131.8 kW respectively as shown in Fig. 6.22. HX-3 is a gas-gas exchanger with 810.8 kW load. No heat exchanger in HEN violates

ΔT_{min}. For Process-3, difference of internal heat change, between hot and cold streams, can be observed before and after modification through Scheme-2. These are 4029.21 kW and 2738.79 kW as shown in composite curve (Fig 6.21) and HEN respectively. Reasons for this difference is similar to that explained in Sections 6.1.2 and 6.1.4.

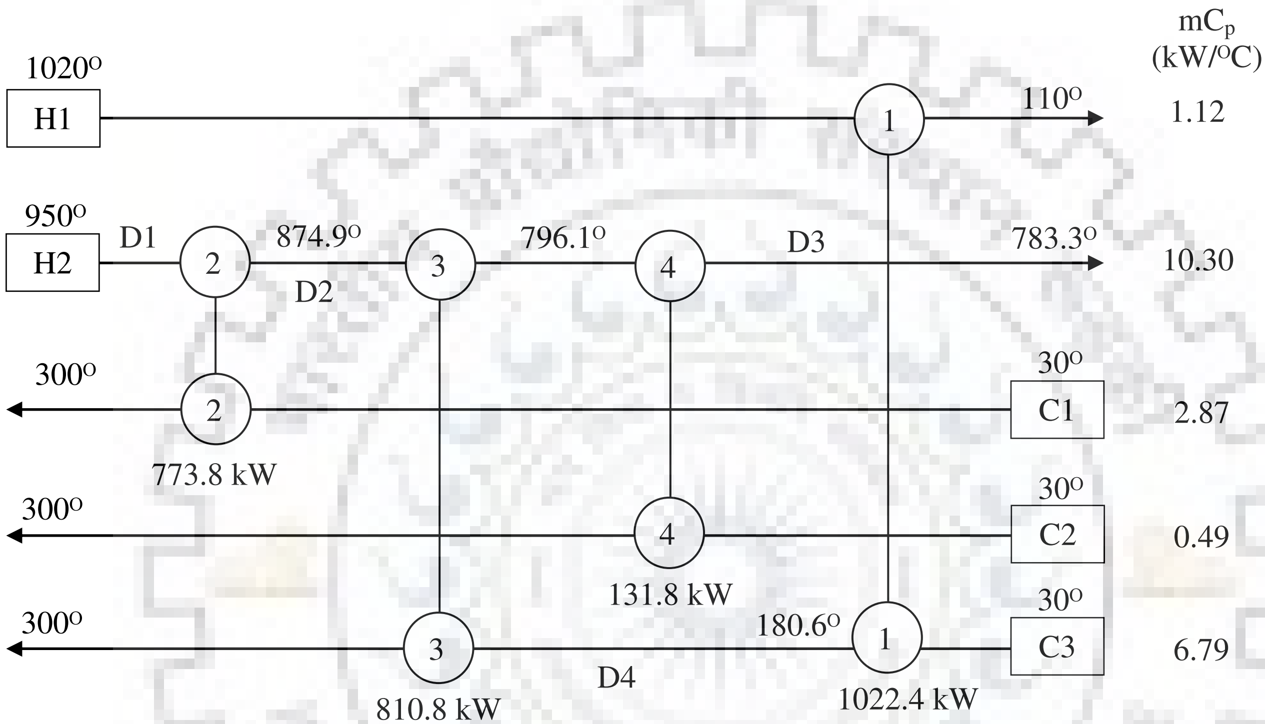


Fig. 6.22 Heat Exchanger Network of Scheme-2 for Process-3

Four ducts, three for carrying ABC exit gas and one for carrying kiln air, are required in the modified PFD of Process-3 for Scheme-2 which is shown in Fig. 6.23. Initially, waste gas is required to bring from ABC to kiln inlet, which needs a duct of length 12m. Further, following ducts are used: Second duct of 40m length from kiln inlet to kiln outlet to preheat kiln air & slinger coal and third duct of length 52m to carry back waste gas from kiln outlet to ABC exit. Finally, a duct of length 7m is employed to carry kiln air from cooler inlet to kiln outlet. The reduced amounts of coal and air cause less waste gas to discharge from the process, which is 37.14% less in comparison to existing PFD. Similarly, 97.5% of less water consumption is found in modified system of Process-3.

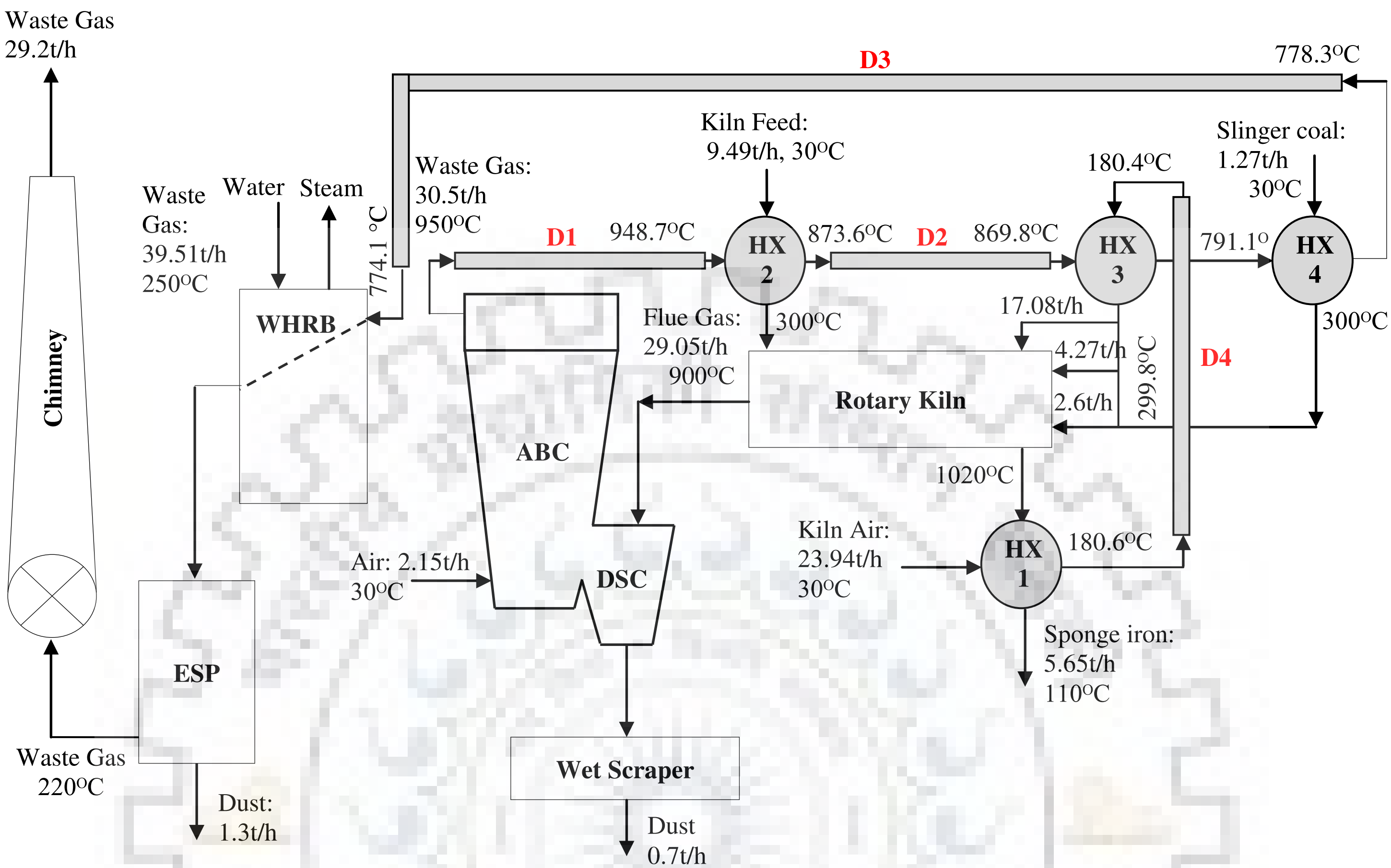


Fig. 6.23 Modified process flow diagram of Process-3 for Scheme-2

6.1.6.1 Operational aspects of Process-3 for Scheme-2

Temperature of ABC exit gas, having CP of 10.3 kW/°C, drops from 950°C to 948.7°C while moving from ABC to kiln inlet through an insulated duct (D1) of length 12 m and diameter 0.3m. After reaching kiln inlet, as shown in Fig. 6.23, ABC exit gas exchanges heat with kiln feed in gas-solid heat exchanger, HX-2, dropping its temperature to 873.6°C instead of from 950°C to 974.9°C shown in HEN (Fig. 6.22). Then it moves through second duct (D2) of 40 m length and 0.3m diameter by dropping its temperature to 869.8°C before entering into gas-gas exchanger, HX-3, where it exchanges 810.8 kW heat with kiln air for preheating it from 180.4°C to 299.8°C instead of from 180.6°C to 300°C as shown in Fig. 6.22. Further waste gas enters into gas-solid exchanger, HX-4, at 791.1°C instead of 796.1°C as shown in Fig. 6.22 and exchanges 131.8 kW of heat with slinger coal. Consequently its temperature drops to 778.3°C. However, in Fig. 6.22 waste gas exits the exchanger, HX-4, at 783.3°C.

Kiln air that was already preheated in gas-solid heat exchanger, HX-1, moves through an insulated duct (D4) of length 7m and diameter 0.3m from cooler inlet to kiln outlet dropping its temperature drops from 180.6^{o}C to 180.4^{o}C as shown in Fig. 6.22. Finally waste gas is carried back to ABC through a non-insulated duct (D3) of 52m length and 0.3m diameter in which its temperature drops from 778.3^{o}C to 774.1^{o}C. The temperature profiles of all four ducts are shown in Fig. 6.24. Mild steel is used for casing of the duct and for insulation ceramic wool is placed at outer surface of the duct.

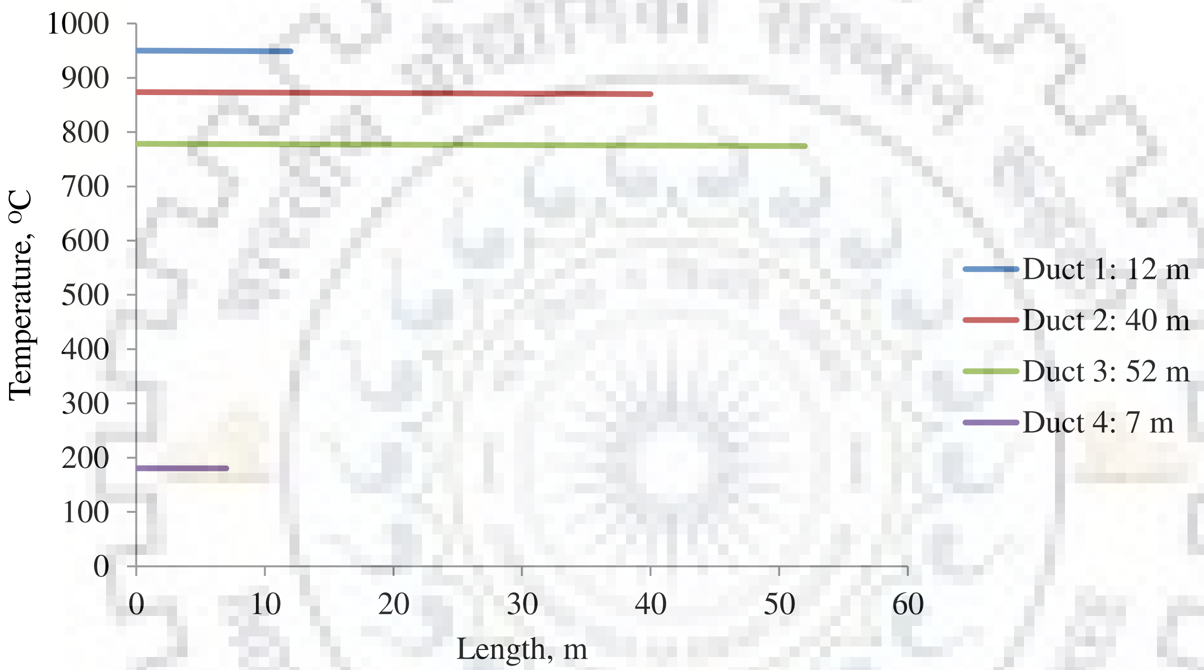


Fig. 6.24 Temperature profile of waste gas in ducts for Process-3 Scheme-2

Pressure drops, created while gases are flowing through these ducts, are evaluated using method discussed by Prasad [133] and are found as 0.056 atm, 0.174atm, 0.211 atm and 0.008 atm for ducts of length 12m, 40m, 52m and 7m respectively as reported in Table 6.17. Four FD fans are required, one for each duct, to compensate these pressure drops in ducts as one FD fan can sustain 1.26 atm pressure drop [177].

Table 6.17 Pressure and temperature drops in ducts for Process-3 Scheme-2

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (°C)	T_o (°C)
D1	12	ABC exit gas	ABC exit	Kiln inlet	0.056	950.0	948.7
D2	40	ABC exit gas	Kiln inlet	Kiln outlet	0.174	873.6	869.8
D3	52	ABC exit gas	Kiln outlet	ABC exit	0.211	778.3	774.1
D4	7	Kiln air	RC inlet	Kiln outlet	0.008	180.6	180.4

6.1.6.2 Economic aspects of Process-3 for Scheme-2

The total economic analysis of Process-3 for Scheme-2 is carried out in terms of operating cost, capital investment, total annual cost (TAC), savings and payback period as discussed in Sections 6.1.2.2 and 6.1.4.2. Results are summarized in Table 6.18. It shows that discounted payback period of 7.1 months is required for the design modification of Process-3 through Scheme-2.

Table 6.18 Economic analysis of Scheme-2 for Process-3

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	815.7	HX-1	117.4	1059.8	1296.04	7266.6	160.0	7.1
Water	31.3	HX-2	103.5					
Electricity	99.3	HX-3	18.7					
		HX-4	67.8					
		Ducts	28.7					
		FD fans	360.9					
		Total	697.0					

6.2 SELECTION OF BEST HEAT INTEGRATION SCHEME

As discussed in Section 6.1 two different schemes, Scheme-1 and Scheme-2 are proposed for heat integration in three coal based sponge iron industries such as Process-1, Process-2 and Process-3 being operated with different capacities in an industrial cluster of India. This section summarizes results predicted in Section 6.1 from heat integration studies carried out for these three processes individually, through Scheme-1 and Scheme-2. Here these two heat integration schemes are examined to select best scheme for each process. For this purpose results of Scheme-1 and Scheme-2 are reported in Table 6.19 and compared with the existing system based on all economical parameters such as coal consumption, water requirement, operating cost, capital investment, total annual cost, total savings and payback period. Efficacies of two schemes are also computed while comparing results of these based on waste gas generation, energy ratio (ER) and %heat recovery.

Table 6.19 shows that Scheme-2, for Process-1, gives 28.88% savings in coal consumption as compared to maximum of 17.72% savings in Scheme-1. This is because in Scheme-2 kiln inlet streams are preheated from 30^{O}C to 300^{O}C utilizing 2901.2 kW of waste heat. However, Scheme-1 utilizes only 1831.1 kW of waste heat for preheating kiln feed and slinger coal from 30^{O}C to 120^{O}C and kiln air from 30^{O}C to 243.7^{O}C as shown in Table 4.1. Consequently, in Scheme-2, kiln inlet streams are heated to reaction temperature inside the kiln, using coal, at higher temperature than that in Scheme-1. For this purpose, Scheme-2 requires 2205.7 kg/h of coal whereas; Scheme-1 considers 2504.6 kg/h, which is 13.6% more. In the existing process requirement of coal consumption for the same purpose is 3185.3.8 kg/h.

Similarly, for Process-2, Scheme-2 shows 33.88% savings in coal consumption though it is 21.83% in Scheme-1. This is because in Scheme-2, similar to Process-1, kiln inlet streams are preheated to higher temperature before these enter to the kiln than in Scheme-1. It is due to the fact that preheating utilizes 3543.8 kW of waste heat in Scheme-2 whereas Scheme-1 requires only 2530.4 kW. Thus, 2848.3 kg/h and 2572.4 kg/h of coal is required in Scheme-1 and Scheme-2, respectively, for preheating kiln inlet streams up to reaction temperature whereas in existing process it is 3537.1 kg/h.

Similar to Process-1 and Process-2, in Process-3 also, heat integration through Scheme-2 gives more coal savings (36.33%) than through Scheme-1 (23.85%) as it utilizes more waste heat

(4029.22 kW) than Scheme-1 (1925.5 kW) for the purpose of preheating kiln inlet streams. Because of this kiln inlet streams are heated to reaction temperature inside the kiln, using coal, at higher temperature than that in Scheme-1. Consequently, Scheme-2 requires 3007.9 kg/h of coal whereas Scheme-1 requires 3596.8 kg/h which is 19.6% more. In the existing process 4135.8 kg/h of coal is required for the same purpose. Thus, it can be concluded from Table 6.19 that though Scheme-1 saves huge amount of coal, Scheme-2 saves even more in sponge iron processes. The variation in coal consumption for all three processes in both schemes is shown in Fig. 6.25.

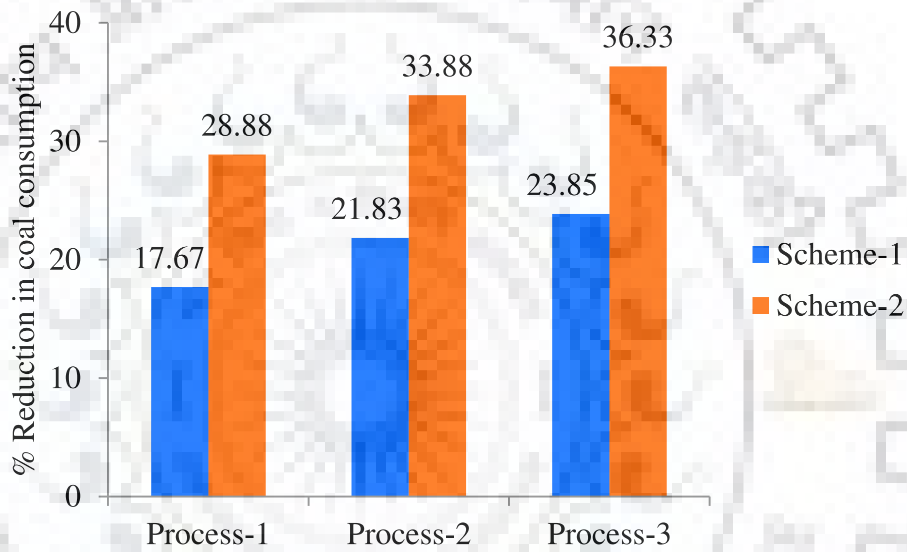


Fig. 6.25 Variation of coal consumption in Scheme-1 and Scheme-2

To estimate the efficiency of proposed design modifications in the present work a parameter ER (energy ratio) is coined, which is defined as the ratio between actual energy consumed in the kiln to theoretical energy required for the reduction reactions to continue. From the material and energy balances, carried out in all three processes in Section 4.1, it is computed that Process-1 requires 19643.2 kW of theoretical energy for reduction reactions to continue in kiln whereas, in actual practice 33886.1 kW of energy is consumed, thus ER is 1.73 under existing conditions. After heat integration through Scheme-1, in Process-1, actual energy consumed is reduced to 27897.8 kW, which is further decreased to 24098.3 kW through Scheme-2. Thus, ER value of Process-1 reduces to 1.42 and 1.23 through Scheme-1 and Scheme-2, respectively. That means, the excess energy consumption of 73%, prior to heat integration, is reduced to 42% and 23% through Scheme-1 and Scheme-2, respectively. The ER values for all schemes and processes are shown in Table 6.19.

Similarly, it is computed that theoretical energy of 20770 kW is required to carry out reduction reactions in Process-2 though in actual practice 36731 kW of energy is consumed. This is reduced to 28713.2 kW and 24284.8 kW through Scheme-1 and Scheme-2 respectively. Thus ER value which is 1.77 under existing conditions, as shown in Table 6.19, is reduced to 1.38 and 1.17 through heat integration schemes. In the similar manner to Process-1 and Process-2, the actual energy consumption for Process-3 is computed and found that it is reduced from 41093.2 kW to 31294.1 kW and 26164.1 kW through Scheme-1 and Scheme-2, respectively, whereas theoretical energy of only 23228.8 kW is required. Thus ER value which is 1.77 under existing conditions is reduced to 1.35 and 1.13 through heat integration schemes. Thus the excess energy consumption, computed in all three processes, is very less in Scheme-2 than in Scheme-1. Hence, actual energy consumption, in Scheme-2, is more close to theoretical energy required than in Scheme-1. The variation in ER values, for all three processes in both schemes, is shown graphically in Fig. 6.26.

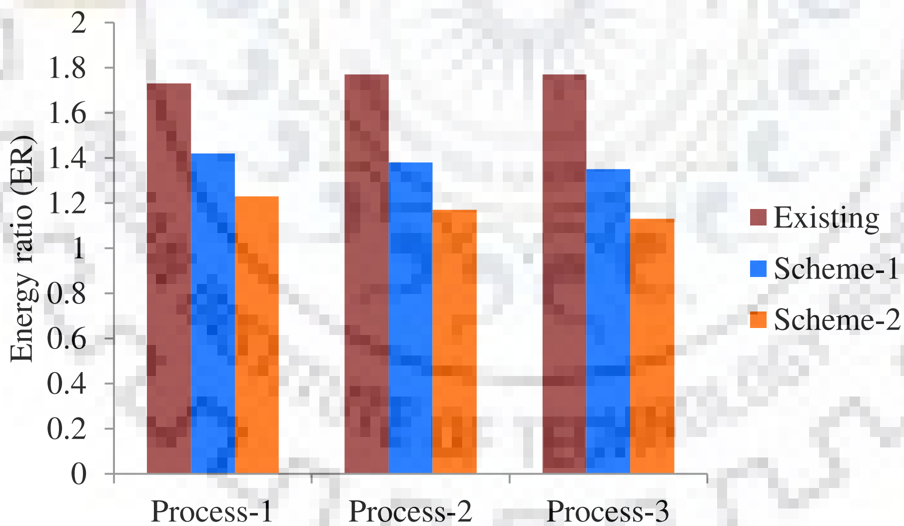


Fig. 6.26 Variation of energy ratio (ER) in Scheme-1 and Scheme-2

The variation in waste gas generation for both schemes in all three processes is also reported in Table 6.19. These results further indicate that both schemes produce significantly less waste gas in comparison to existing system by proper integration of heat in sponge iron processes. Scheme-1 produces 35.23 t/h of waste gas whereas; Scheme-2 generates 28.37 t/h in comparison to existing sponge iron process which produces 44.17 t/h as shown in Table 6.19. Thus, for Process-1 Scheme-1 and Scheme-2 release 35.78% and 20.24%, respectively, less waste gas in comparison to the existing system. Similarly, Scheme-2 produces less waste gas than Scheme-1 in Process-2 and Process-3 as shown in Table 6.19 and Fig. 6.27. The primary fossil fuel energy consumption of India by coal gives strong concern on greenhouse gas emissions and climate change. Reduction in coal (energy) consumption through heat integration schemes reduces waste gas released into the atmosphere and thereby making the sponge iron production processes more environment friendly.

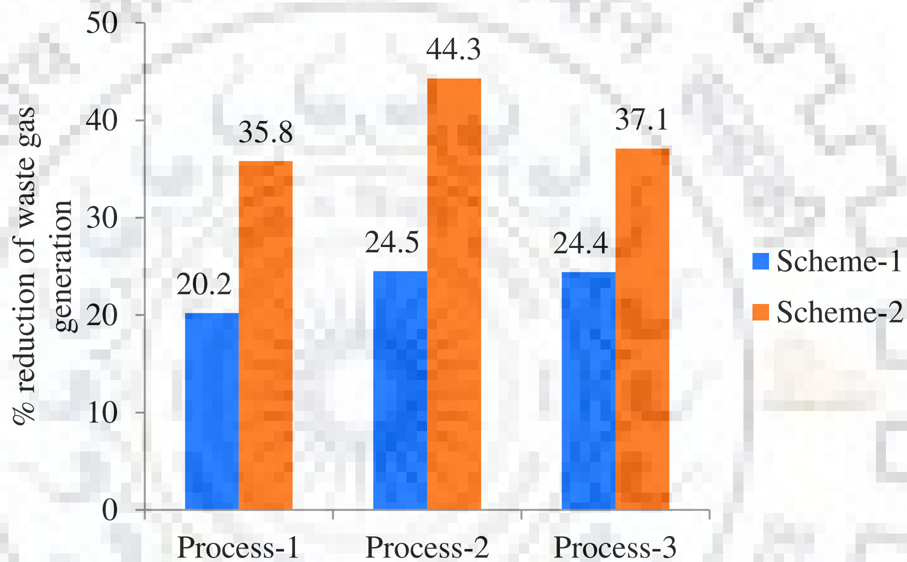


Fig. 6.27 Variation of reduction in waste gas generation in Scheme-1 and Scheme-2

The %heat recoveries in all three processes, through design modifications Scheme-1 and Scheme-2 are provided in Table 6.19 and also shown graphically in Fig. 6.28. It is computed that total waste heat that can be possibly recovered in Process-1 is 11,255.8 kW and Scheme-1 and Scheme-2 recovers 1,987.7 kW (17.66%) and 2,901.2 kW (25.78%), respectively. In the case of Process-2, total waste heat available to recover is 11454.6 kW and Scheme-1 and Scheme-2 recovers only 2801 kW (24.45%) and 3543.8 kW (30.94%), respectively. Similarly in Process-3, Scheme-1 recovers 2083.1 kW (86.63%) out of 2404.6 kW waste heat available in the process whereas, Scheme-2 utilises only 1022.4 kW (42.52%). The reason for lower heat recovery in Process-3 through Scheme-2 is explained as: in Process-3 waste heat, available in Area-3, is already utilized in existing conditions and only Area-2, effectively, is considered to compute %heat recovery in Scheme-2 whereas in Scheme-1 both Area-1 and Area-2 are used for the same purpose. Areas in different processes are defined in Section 4.2. Thus, it can be concluded from Table 6.19 that for Process-1 and Process-2 which do not have power generation unit Scheme-2 recovers maximum amount of waste heat than Scheme-1. On the other hand, for Process-3 which already has power generation unit, as shown in Fig. 4.3, Scheme-1 recovers more amount of waste heat as compared to Scheme-2.

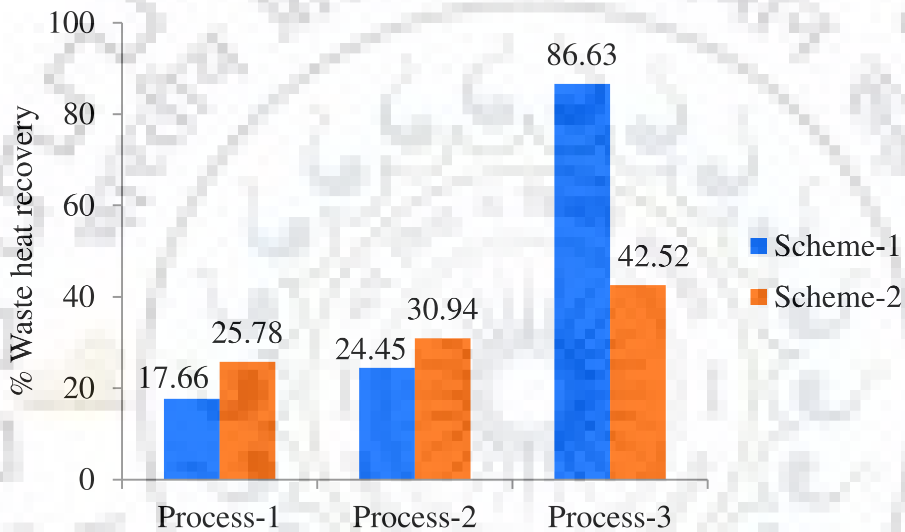


Fig. 6.28 Variation of % waste heat recovery in Scheme-1 and Scheme-2

Further, Table 6.19 shows that Scheme-2 saves 96.97% water, for Process-1, in comparison to existing system which means that Scheme-2 will save 17.29 MT of water, in Process-1, during the plant life of 10 years. This amount is highly significant in the present day scenario because of increasing fresh water crisis. Similarly, Scheme-2 saves 97.28% and 97.51% water in comparison to existing system for Process-2 and Process-3, respectively, as shown in Table 6.19. These savings are higher than that offered by Scheme-1 although the difference is very small.

It is interesting to note that Scheme-2 gives maximum savings and also involves minimum investment. As shown in Table 6.19, the capital investment required for design modification through Scheme-2, in Process-1, is 657.4 lakh and savings it offers is 1243.3 lakh/year. On the other hand, the design modification through Scheme-1 requires capital investment of 705.2 lakh while offering 1084.7 lakh/year as savings, which is 14.62% lower than Scheme-2. TAC presented in Table 6.19, shows that it is mainly controlled by annual operating cost and for Scheme-2 it is 14.68% less as compared to Scheme-1. As coal and water consumptions are less in Scheme-2 than in Scheme-1 the operating cost for Scheme-2 is 15.2% less than that of Scheme-1. Due to 14.62% higher savings in comparison to Scheme-1, Scheme-2 offers less payback period (7.91 months) than Scheme-1 (6.43 months).

Similar trends in results that are observed for Process-1 can also be found for Process-2 and Process-3, as shown in Table 6.19. Here, Scheme-2 offers minimum investment with more savings and less payback periods than Scheme-1. In Process-3, difference of savings between Scheme-1 and Scheme-2 is less in comparison to other two processes as Process-3 is already utilizing heat of ABC waste gas for power generation. The variations in capital investment and savings for both schemes in all three processes are shown in Fig. 6.29 and Fig. 6.30, respectively.

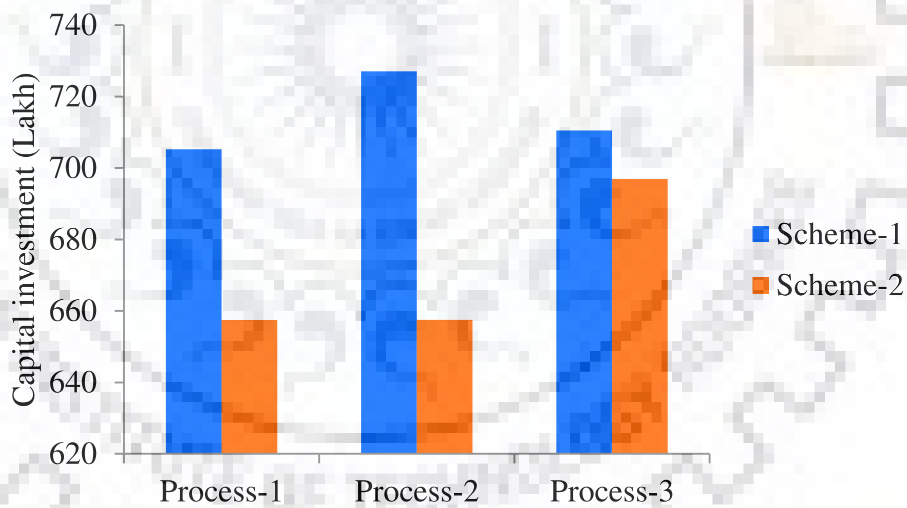


Fig. 6.29 Variation of capital investment in Scheme-1 and Scheme-2

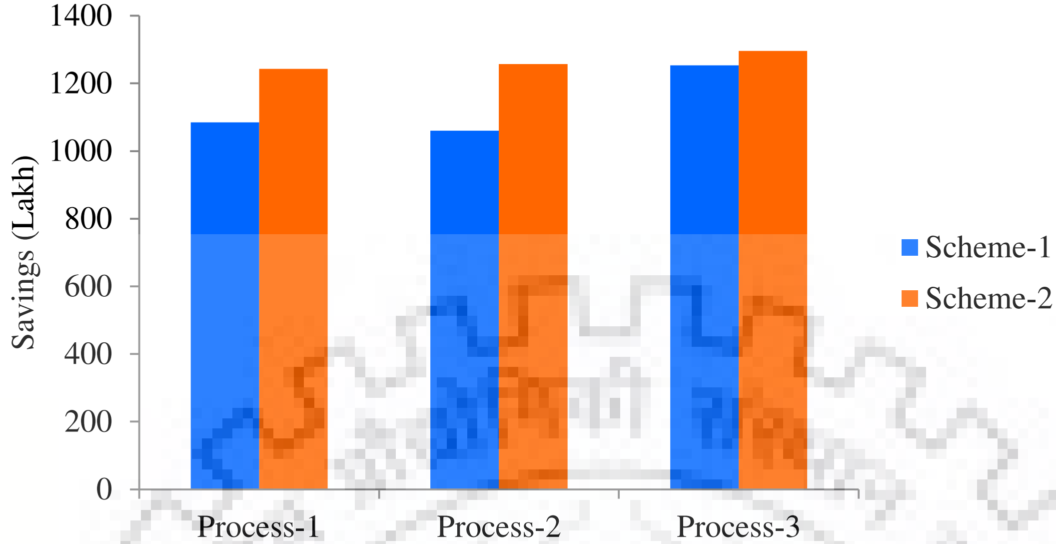


Fig. 6.30 Variation of savings in Scheme-1 and Scheme-2

Thus, based on energy consumption, water requirement, waste gas generation, energy ratio (ER) and %heat recovery along with economic analysis discussed in this section Scheme-2 is selected as better heat recovery option over Scheme-1 for all sponge iron processes of industrial cluster.

Table 6.19 Comparative analysis of Scheme-1 and Scheme-2

Case	Resource Consumption (t/h)		Operating cost	Capital investment (Lakh)	TAC (Lakh/ year)	Waste gas generated (t/h)	Energy ratio (ER)	(%) Heat recovery	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
	Coal	Water	(Lakh/ year)									
						Process-1						
Existing	5.36	226.17	-	-	-	44.17	1.73	-	-	-	-	-
Scheme-1	4.41	10.09	1041.67	705.23	1156.4	35.23	1.42	17.66	1084.68	5959.7	130.7	8.58
Scheme-2	3.81	6.86	883.08	657.44	990.1	28.37	1.23	25.78	1243.26	6981.9	162.8	6.98
						Process-2						
Existing	5.81	210.34	-	-	-	61.05	1.77	-		-	-	-
Scheme-1	4.54	12.75	1079.69	726.96	1198.0	46.09	1.38	24.45	1060.43	5788.9	123.5	9.05
Scheme-2	3.84	5.72	883.50	657.50	990.5	33.99	1.17	30.94	1256.62	7063.9	164.7	6.91
						Process-3						
Existing	6.5	265.33	-	-	-	46.45	1.77	-	-	-	-	-
Scheme-1	4.95	10.64	1150.17	710.51	1265.8	35.13	1.35	86.63	1253.52	6991.8	151.3	7.48
Scheme-2	4.14	6.62	946.36	696.96	1059.8	29.20	1.13	42.52	1296.04	7266.6	160.0	7.10

6.3 ANALYSIS OF DIFFERENT STRATEGIES OF TOTAL SITE INTEGRATION

In the present work three different strategies are developed for integration of total site. For this purpose three different sponge iron processes, Process-1, Process-2 and Process-3, which are located nearby in an industrial cluster of India, are considered. These processes are discussed in detail in Section 3.1. Detailed discussion of three strategies is carried out in Section 4.4. Strategy-0 is developed based on the best scheme for Process-1, Process-2 and Process-3 found in Section 6.2. Strategy-1 and Strategy-2 are proposed based on the modified approach of total site integration developed by Ramanaiah and Khanam [142] to conserve energy in total site of plants of similar type where conventional methods are not applicable. Detailed results of these strategies are discussed in this section.

6.3.1 Strategy-0

Scheme-1 and Scheme-2 are proposed for heat integration in individual processes, separately. Among these, Scheme-2 is found as best for each process as discussed in Section 6.2. In Strategy-0 the best scheme, selected for individual process, is used for total site integration. Thus, the stream data of Scheme-2 for Process-1, Process-2 and Process-3, shown in Table 4.2 is used for total site integration also. For all three processes, Process-1, Process-2 and Process-3, hot utilities are found to be zero, as mentioned in Section 6.1, whereas cold utilities are found to be 0.176 kW, 0.0013 kW and 0.007 kW, respectively. Site source sink profiles (SSSP) is constructed for the same stream data, shown in Table 4.2, using the principles of conventional total site integration [80] and shown in Figure 6.31.

The plot, shown in Fig. 6.31, indicates that hot and cold utilities of total site are 0 kW and 0.184 kW, respectively. It is the sum of hot and cold utilities of individual processes. As hot and cold utilities of total site are not changing the HENs of individual processes, shown in Fig. 6.6, Fig. 6.14 and Fig. 6.22, remain unaltered while considering these for total site. As all three processes are similar, there is no relative advantage of conventional total site integration, which is more suitable to integrate heat within the site of different processes. In the present case, composite curves of individual processes, drawn in Fig. 6.5, Fig. 6.13 and Fig. 6.21 for the stream data shown in Table 4.2, are similar and the opportunity to match streams among different processes is same as matching within the individual process. Thus, total site integration through usual method is not beneficial as no improved modification is possible in the case of coal based sponge iron industrial cluster, which is considered in the present work.

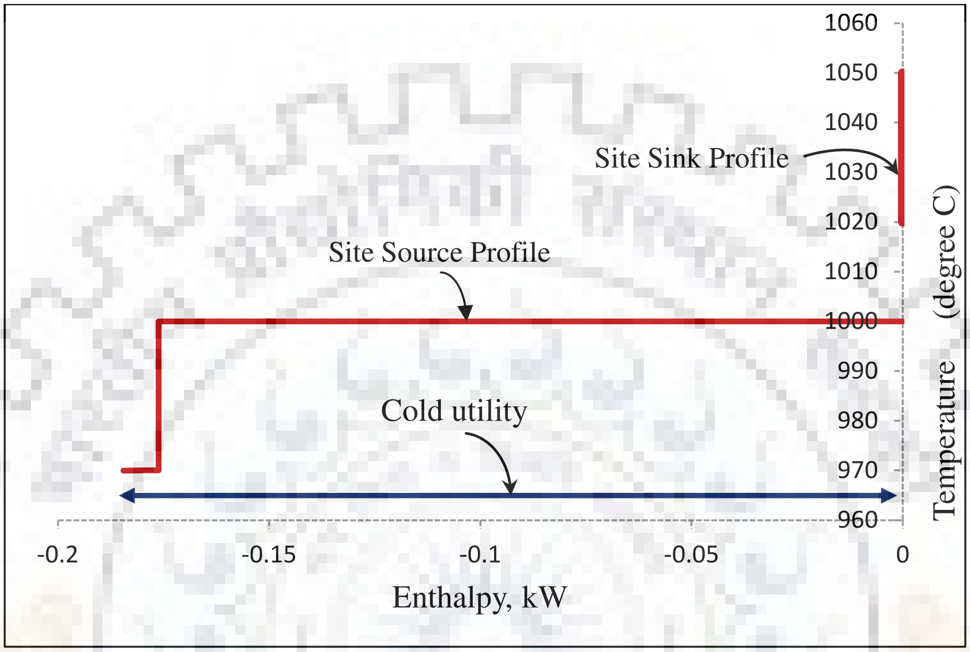


Fig. 6.31 Site source sink profiles for Strategy-0

Further, coal is used as a process stream in sponge iron process as it is heated inside the kiln up to the reaction temperature and at the same time coal is used as utility stream as it provides heat through the combustion. In fact, in sponge iron process coal is the only source of heat. Thus, for such process the separation of process/utility data is difficult, which makes integrating the utility network infeasible through conventional total site integration concept.

Therefore, two different strategies for total site integration of sponge iron cluster are proposed as Strategy-1 and Strategy-2 based on the modified approach of total site integration developed by Ramanaiah and Khanam [142] to conserve energy in total site of plants of similar type. These are formulated under Section 4.4. Detailed discussion on modified approach of total site integration is given in Section 5.2.

6.3.2 Strategy-1

Based on the guidelines given in Section 4.4, two different energy conservation strategies, Strategy-1 and Strategy-2, are proposed. These strategies are proposed considering the working areas discussed in Section 4.2. These two strategies vary in terms of modification, investments and savings.

In Strategy-1, waste gas exiting ABC of Process-1 splits into three branches and is used for preheating kiln feed and slinger coal of all three processes, individually, from 30^{O}C to 300^{O}C. Kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air of its respective process. Further, kiln air is preheated using waste gas to maximum possible temperature (421^{O}C in this case) such that ΔT_{min} equal to 50^{O}C is maintained. After preheating kiln feed and slinger coal of all three processes three branches of waste gas are again mixed to enter into ESP of Process-1. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation. Once power is generated waste gas stream further splits into two parts. One part enters ESP of Process-2 whereas the other ESP of Process-3. The partial schematic diagram of Strategy-1 is shown in Fig. 6.32. The stream data of Strategy-1, which consists of 5 hot and 9 cold streams, are shown in Table 4.3. In fact, this table includes stream data of all three plants together and thus, all three plants are considered as single plant in Strategy-1. The number shown as a subscript in Fig. 6.32 denotes the specific process in the total site as reported in Table 4.3 and discussed in Section 4.4.1.

The energy requirement for this strategy is targeted using pinch analysis [161] considering ΔT_{min} as 50^{O}C. For stream data of Strategy-1, shown in Table 4.3, the hot and cold utilities are found to be 0 kW and 23617.2 kW, respectively, which indicate that it is the threshold problem. Fig. 6.33 and Fig. 6.34 depict composite and grand composite curves (GCC) with an internal heat exchange of 13850.4 kW. For computation of coal and air requirements a trial and error technique, as shown in Fig. 5.1, is used as these pursue dual nature i.e. both utility stream as well as process stream. The computation is similar to as discussed in the Section 6.1.1. The results of different iterations are reported in Table 6.20.

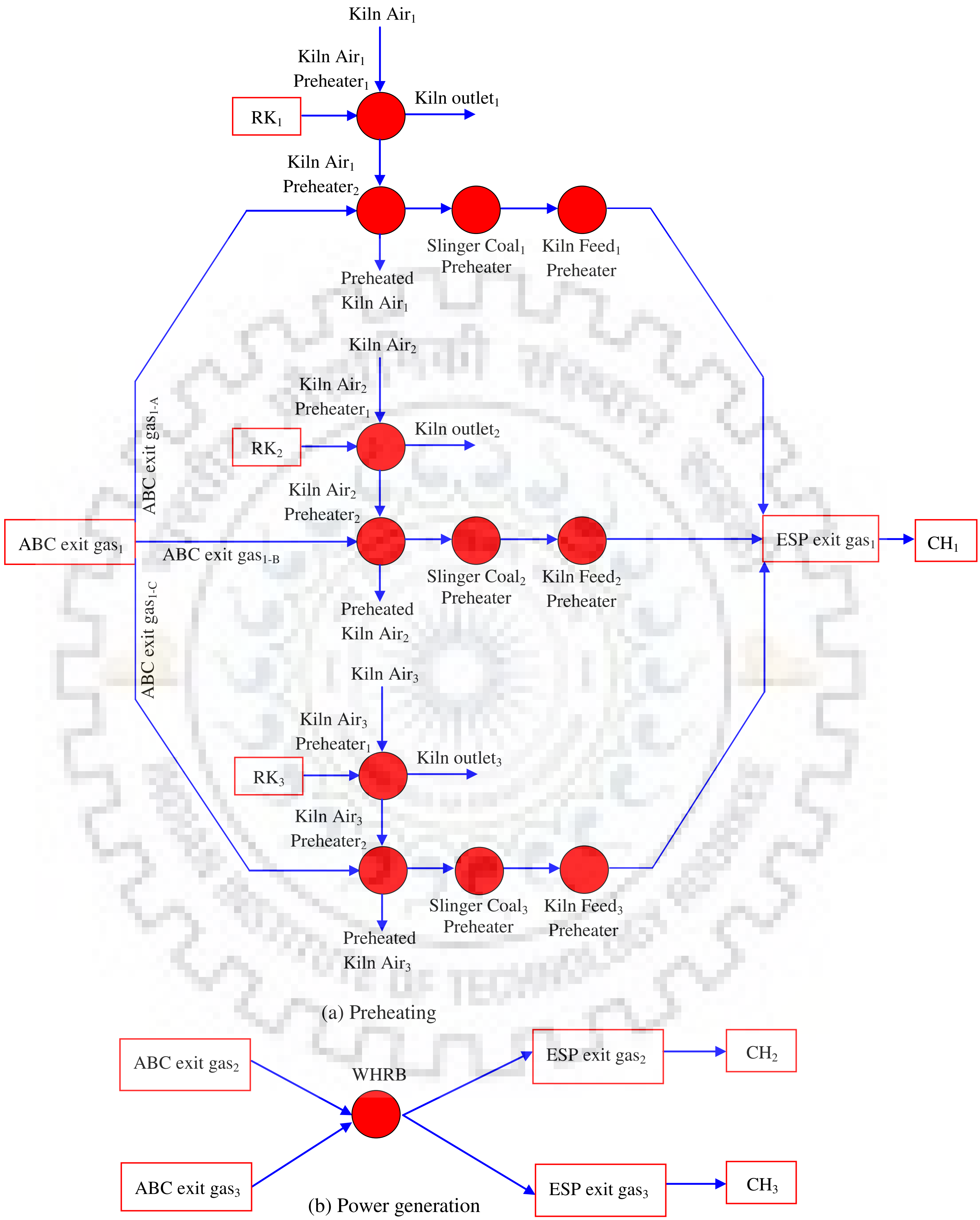


Fig. 6.32 Partial schematic diagram of Strategy-1

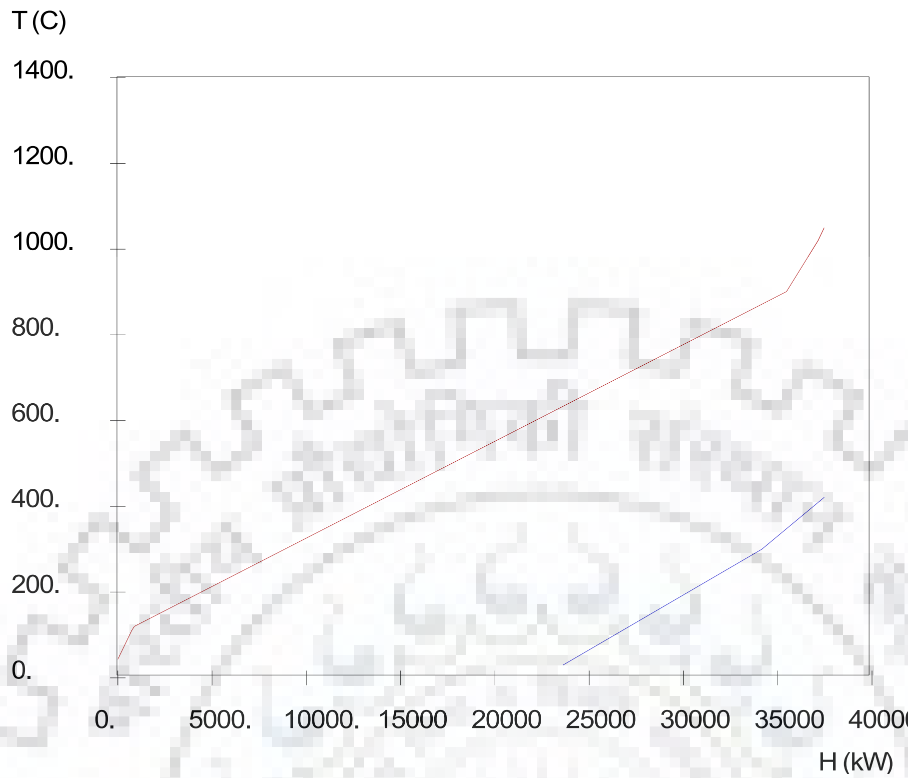


Fig 6.33 Composite curves for Strategy-1

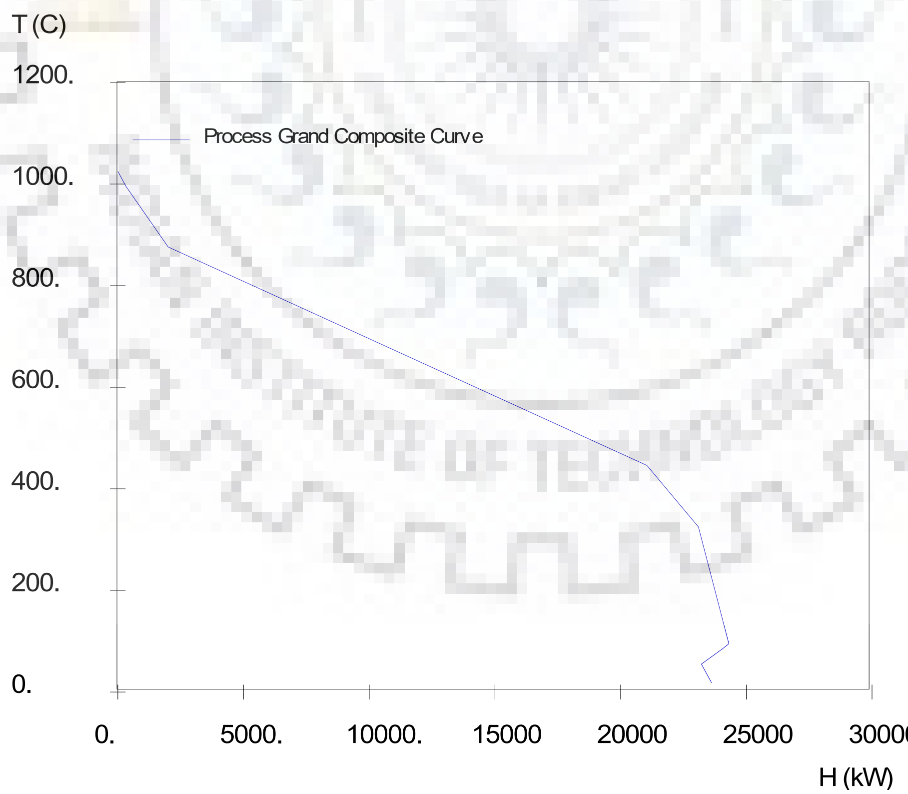


Fig 6.34 Grand composite curve for Strategy-1

Table 6.20 Iteration results for Strategy-1

Iteration No	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	17670.0	95636.5
1	0	1	12838.8	69488.4
		2	11456.6	62007.5
		3	11061.2	59867.2
		4	10948.0	59254.9
		5	10915.7	59079.7
		6	10906.4	59029.6
		7	10903.8	59015.3
		8	10903.0	59011.2
		9	10902.8	59010.0
		10	10902.7	59009.7
		11	10902.7	59009.6
		12	10902.7	59009.5
		13	10902.7	59009.5
2	0	-	10902.7	59009.5

For Strategy-1, total coal consumption, for all plants, at Iteration-0 (existing system) is 17.67 t/h (Process-1 = 5.36 t/h, Process-2 = 5.81 t/h, Process-3 = 6.5 t/h) whereas, kiln air is 95.64 t/h (Process-1 = 24.80 t/h, Process-2 = 33.23 t/h, Process-3 = 37.60 t/h), respectively, as shown in Table 6.20. For iteration-1 coal is calculated using Equation 4.9 and found as 12.84 t/h. The corresponding amount of air is computed using ratio of air to coal as maintained in iteration-0, which is 5.41, and found as 69.49 t/h. However, while computing coal consumption in first stage of iteration-1, through Equation 4.9, air available in iteration-0 i.e. 95.64 t/h is used. This air to coal is not in same ratio as 5.41. Thus, revised value of air i.e. 69.49 t/h is used in second stage of iteration-1 to compute amount of coal, which is found to be 11.46 t/h as shown in Table 6.20. Corresponding amount of air is 62.01 t/h. Similarly, different stages of iteration-1 are performed till the values of coal and air become equal in two consecutive stages and results are reported in Table 6.20. Final amounts of coal and air are 10.90 t/h (Process-1 = 3.31 t/h, Process-2 = 3.59 t/h, Process-3 = 4.01 t/h) and 59.01 t/h (Process-1 = 15.30 t/h, Process-2 = 20.50 t/h, Process-3 = 23.20 t/h), respectively, and these values maintain the ratio of 5.41. In Strategy-1, as hot utility is zero, amounts of coal and kiln air remain same in iteration-2. Therefore, 38.3 % of coal requirement is reduced in comparison to the existing process. Reduction in coal consumption is due to total site integration through Strategy-1, where coal and kiln air are preheated from 30 ^{o}C to 300 ^{o}C and from 30 ^{o}C to 421 ^{o}C, respectively.

Based on guidelines proposed in Section 4.4, HEN for the modified design of total site integration through Strategy-1 is shown in Fig. 6.35. It includes twelve heat exchangers, which do not violate ΔT_{min}. These are additional equipment to be employed in the modified design of total site integration through Strategy-1. Temperatures and heat loads shown in HEN (Fig. 6.35) are based on values of final iteration reported in Table 6.20.

The exchanger, HX-1, is a gas-solid heat exchanger with 953.1 kW heat load is used between kiln outlet stream (H1) and kiln air (C1) of Procees-1. Hence, kiln outlet stream of Process-1 cools from 1020^{O}C to 110^{O}C whereas kiln air of Process-1 heats up from 30^{O}C to 247.7^{O}C. Waste gas that exits from ABC of Process-1 at 1050^{O}C (Stream no. H5 in HEN) splits into three branches as discussed in Section 4.4.1 and shown in Fig. 6.35. These are denoted as ABC exit gas$_{1\text{-}A}$, ABC exit gas$_{1\text{-}B}$ and ABC exit gas$_{1\text{-}C}$ and also shown in Fig. 6.32. Using first branch of waste gas (ABC exit gas$_{1\text{-}A}$) kiln air of Process-1, which is already preheated up to 247.67^{O}C in heat exchanger HX-1, is further preheated in exchanger HX-2 to 421^{O}C. It is a gas-gas heat exchanger having heat load of 758.9 kW where temperature of waste gas drops from 1050^{O}C to 596.7^{O}C, as shown in Fig 6.35. Same branch of waste gas then enters into gas-solid heat exchanger HX-3 at 596.7^{O}C, to preheat slinger coal of Process-1 (C4) from 30^{O}C to 300^{O}C. In this heat exchanger waste gas temperature drops from 512.8^{O}C while exchanging 140.5 kW of heat load as shown in Fig. 6.35. Further, same waste gas stream is cooled to 125^{O}C in HX-4 while transferring 649.2 kW of heat to preheat kiln feed of Process-1 (C7) from 30^{O}C to 300^{O}C.

A gas-solid heat exchanger, HX-5, transfers 795.2 kW heat from kiln outlet stream (H2) to kiln air (C2) of Procees-2. Hence, kiln outlet stream of Process-2 cools from 1020^{O}C to 110^{O}C whereas kiln air of Process-2 heats up from 30^{O}C to 165.6^{O}C. Using second branch of waste gas (ABC exit gas$_{1\text{-}B}$), as shown in Fig. 6.35, kiln air of Process-2, which is already preheated up to 165.6^{O}C in heat exchanger HX-5, is further preheated in exchanger HX-6 to 421^{O}C. HX-6 is a gas-gas heat exchanger of 1498.6 kW load where temperature of waste gas drops from 1050^{O}C to 438.7^{O}C, which is shown in Fig 6.35. Same branch of waste gas then enters into gas-solid heat exchanger HX-7 at 438.7°C, to preheat slinger coal of Process-2 (C5) from 30°C to 300°C. In this heat exchanger waste gas temperature drops to 371°C while exchanging 166 kW of heat load as shown in Fig. 6.35. Further the same waste gas stream is cooled to 125°C in HX-8, which has load of 603.1 kW. It is a gas-solid exchanger where kiln feed of Process-2 (C8) is preheated from 30°C to 300°C.

HX-9, a gas-solid heat exchanger, is used to cool kiln outlet of Process-3, from 1020°C to 110°C, using kiln air of the same process. Consequently, as shown in Fig. 6.35, kiln air is preheated to 184°C from 30°C, which is further preheated to 421°C, in HX-10, using third branch of waste gas. Similar to first and second branches of waste gas, which preheat slinger coal and kiln feed of Process-1 and Process-2, third branch of waste gas preheats slinger coal and kiln feed of Process-3 from 30°C to 300°C in gas solid exchangers, HX-11 and HX-12, respectively. As a result of this, temperature of waste gas drops to 411.8°C in HX-11 and finally to 125°C in HX-12. Heat loads of these exchangers are shown in Fig. 6.35.

Thus, first branch of waste gas from ABC exit of Process-1 preheats kiln air, slinger coal and kiln feed of Process-1 while exchanging total 1548.6 kW of heat load as shown in Fig. 6.35. Similarly, second branch of waste gas stream transfers 1548.6 kW of heat to preheat kiln air, slinger coal and kiln feed of Process-2 whereas, 2465.5 kW of waste heat, available with third branch of waste gas stream, is utilized for preheating kiln inlet streams of Process-3 as shown in Fig. 6.35. Thus, HEN shown in Figure 6.35 includes twelve heat exchangers, in which no heat exchanger violates ΔT_{min}.

However, internal heat exchange, shown in composite curve (Fig 6.33), before total site integration through Strategy-1 is different from that shown in HEN, after modification and these are 13850.4 kW and 9052.5 kW, respectively. This difference in internal heat exchange is due to the change in flow rates of kiln inlet streams (kiln feed, slinger coal and kiln air) and waste gas streams of total site and also change in temperature of waste gas stream (H5). After design modification through Strategy-1 amounts of kiln inlet streams and waste gas generation of total site reduce and hence, net heat contents of these also decrease. For example, the flow rate of total site kiln air is 95.64 t/h and 59.01 t/h, before and after total site integration, respectively. Also, before total site integration, waste gas from ABC exit of Process-1 (H5) is cooled to 43.2°C whereas in actual practice it is cooled to only 125°C as shown in Fig. 6.35.

Based on guidelines discussed in Section 4.4, the modified PFD of total site through Strategy-1 is shown in Fig. 6.36. Reduced amounts of coal and air cause less waste gas to discharge from the site. In the existing system, shown in Fig. 4.5, waste gas generation is 151.67 t/h which is reduced to 73.23 t/h in the modified site shown in Fig. 6.36. Thus, 51.72% less waste gas is generated in modified site which is an additional benefit of total site integration through Strategy-1.

In Strategy-1 power generation is carried out using waste gas exiting ABCs of Process-2 and Process-3. For this purpose these two streams of waste gases are mixed and the resultant waste gas stream (Stream no. H4 in HEN) having CP 18.42 kW/^{O}C and temperature 900.5^{O}C is used to generate power. This can be understood through GCC shown in Fig. 6.37 where total cold utility is found as 14379.8 kW. In fact, this value is different than that is shown in Fig. 6.34, which is 23617.2 kW. This is due to the fact that Fig. 6.34 is drawn for existing total site whereas Fig. 6.37 is drawn for modified site where cold utility is reduced due to reduction in amount of waste gas generated as coal and air requirements are decreased after total site integration. Detailed computation of total power produced is illustrated in Appendix B. For producing steam first water is preheated from 30^{O}C to 345^{O}C. At this temperature water is boiled up to saturation condition (155.45 bar) and then the saturated steam is superheated up to 700^{O}C and 300 bar. In this process 14.93 t/h superheated steam is generated, which is expanded through steam turbine having 85% isentropic efficiency [155]. Consequently, 5.19 MW of power is produced. This power is assumed to be exported. Once superheated steam is generated temperature of waste gas cools to 120^{O}C. Total water requirement in the existing site is 705.94 t/h as shown in Fig. 4.5. The modification in the present design eliminates water requirement in rotary cooler (RC) of all three plants and in evaporating chamber (EC) of Process-1 and Process-2. For power generation 14.93 t/h water is required, therefore, 97.9% of water requirement is reduced in comparison to the existing process.

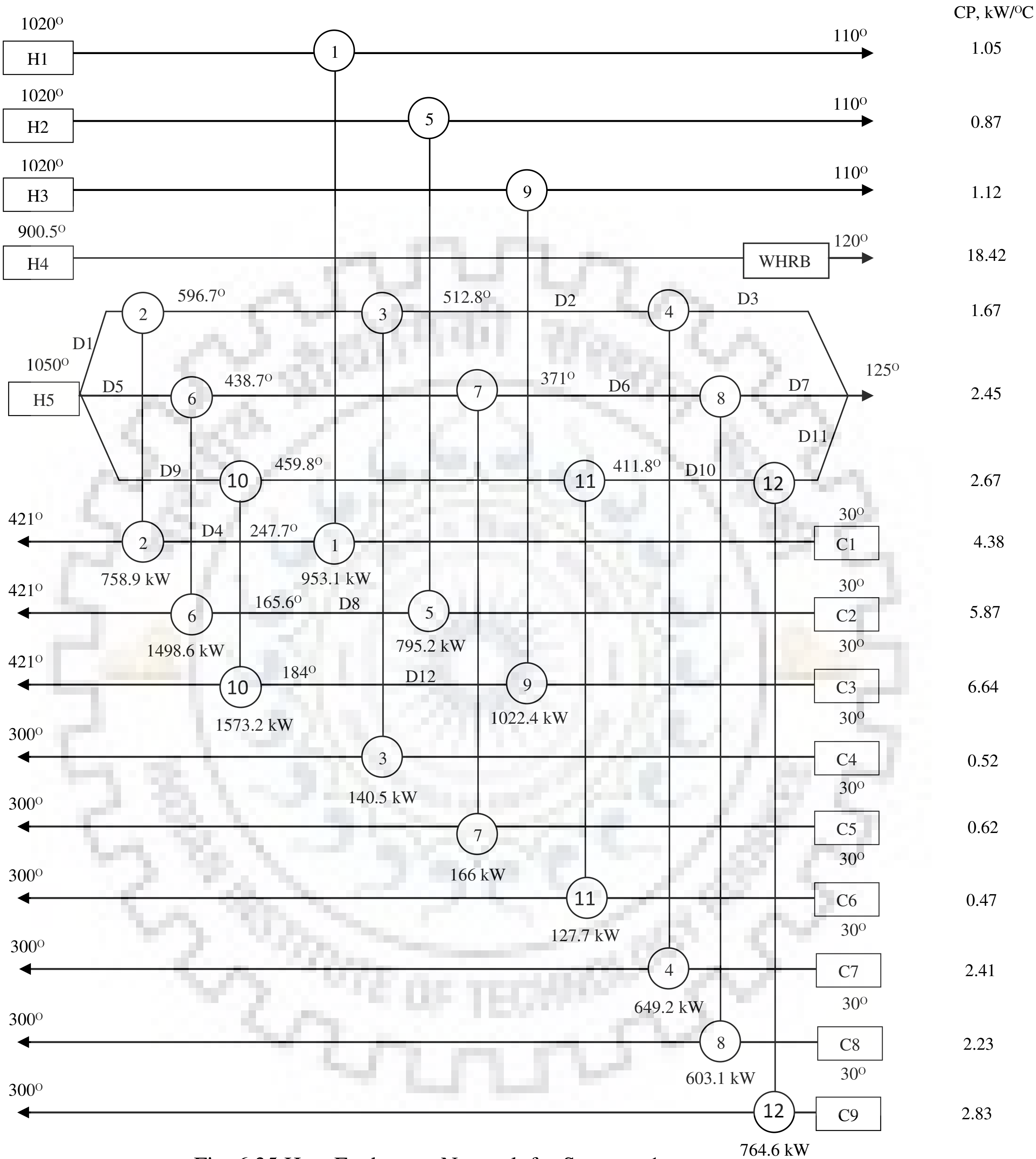


Fig. 6.35 Heat Exchanger Network for Strategy-1

6.3.2.1 Operational aspects of total site integration for Strategy-1

In Strategy-1, as in most of exchangers heat transfer occurs between gas and solid streams, ducts and conveyors are used to transport. However, following guidelines given in Section 4.4, movement of gas is preferred over solid and so, ducts are employed. Thus, Fig. 6.38 includes total sixteen ducts for Strategy-1, thirteen ducts for carrying waste gas from ABC of three processes to different places where they exchange heat before they are directed back to ESP of respective processes and three ducts for carrying kiln air. Initially, waste gas (Stream no. H5) coming out from ABC of Process-1 is splitted into three streams, which are denoted as ABC exit gas$_{1\text{-}A}$, ABC exit gas$_{1\text{-}B}$ and ABC exit gas$_{1\text{-}C}$ as discussed in Section 4.4.1 and shown in Fig. 6.32. Amongst these, first stream (ABC exit gas$_{1\text{-}A}$) is used for preheating kiln air, slinger coal and kiln feed of Process-1. Similarly, second and third streams (ABC exit gas$_{1\text{-}B}$ and ABC exit gas$_{1\text{-}C}$) are used to preheat kiln air, slinger coal and kiln feed of Process-2 and Process-3, respectively.

As distance between ABC of Process-1 and kiln outlet of the same process is 52 m, as shown in Fig. 4.5, an insulated duct (D1) of same length is employed to transfer first branch of waste gas (ABC exit gas$_{1\text{-}A}$), having CP of 1.67 kW/^{O}C, from ABC to kiln outlet of Process-1. While passing through this duct temperature of waste gas reduces from 1050^{O}C to 1038.7^{O}C as shown in Table 6.21. The temperature profile for this duct is calculated using the method given by Prasad [130]. At this temperature waste gas enters to a gas-gas heat exchanger, HX-2, as shown in Fig. 6.36, where kiln air of Process-1 is preheated while exchanging heat of 758.9 kW. Further, it enters into the gas-solid exchanger, HX-3, at 585.4^{O}C, which was 596.7^{O}C in Fig. 6.35. Exchanging 140.5 kW of heat with slinger coal in HX-3 waste gas leaves the exchanger at 501.5^{O}C. However, in Fig. 6.35 waste gas exits the exchanger, HX-3, at 512.8^{O}C. At 501.5^{O}C gas moves through an insulated duct (D2) of 40 m length and 0.3 m diameter and drops its temperature to 497.2^{O}C before entering to the gas-solid exchanger, HX-4. Here, kiln feed is preheated from 30^{O}C to 300^{O}C while exchanging load of 649.2 kW. Temperature of waste gas reduces to 109.4^{O}C in contrast to 125.0^{O}C shown in Fig. 6.35. The kiln air that was already preheated in gas-solid heat exchanger, HX-1, is passed through an insulated duct (D4) of length 5 m and diameter 0.3 m from cooler inlet to kiln outlet. Consequently, kiln air temperature drops from 247.7^{O}C to 247.6^{O}C. Finally, after preheating C1, C4 and C7 streams the first branch of waste gas (ABC exit gas$_{1\text{-}A}$) moves, from kiln inlet to ESP of Process-1, through an non-insulated duct (D3) of 25 m length and 0.3 m diameter dropping its temperature from 109.4^{O}C to 109^{O}C.

Further, an insulated duct (D5) of length 126 m and diameter 0.3 m is employed to transfer second branch of waste gas (ABC exit $gas_{1\text{-}B}$), having CP of 2.45 kW/^{O}C, from ABC to kiln outlet of Process-2. While passing through this duct temperature of the waste gas drops down from 1050^{O}C to 1031.4^{O}C as shown in Table 6.21. At this temperature waste gas enters to a gas-gas heat exchanger, HX-6 shown in Fig. 6.36, where kiln air of Process-2 is preheated while exchanging heat of 1498.6 kW. Further, it enters into the gas-solid exchanger, HX-7, at 420.1^{O}C, which is in contrast to value shown in Fig. 6.35, where waste gas enters at 438.7^{O}C. Waste gas leaves the exchanger, HX-7, at 352.3^{O}C while transferring 166 kW of heat to slinger coal. However, in Fig. 6.35 waste gas exits the exchanger, HX-7, at 371^{O}C. Waste gas exiting exchanger, HX-7, at 352.3^{O}C moves through an insulated duct (D6) of 42 m length and 0.3 m diameter and drops its temperature to 350.1^{O}C. Further, it exchanges load of 603.1 kW in HX-8 to preheat kiln feed from 30^{O}C to 300^{O}C. The temperature of the waste gas reduces to 104.1^{O}C whereas, it is 125.0^{O}C in HEN, shown in Fig.6.35. Kiln air gets preheated to 165.6^{O}C in gas-solid heat exchanger, HX-5, using 795.2 kW of heat of kiln exit stream of process-2. Further, heated air at 165.6^{O}C is passed through an insulated duct (D8) of small length 5 m and diameter 0.3 m from cooler inlet to kiln outlet. Subsequently, its temperature decreases to 165.5^{O}C. Finally, at 104.1^{O}C waste gas (ABC exit $gas_{1\text{-}B}$) moves from kiln inlet of Procees-2 to ESP of Process-1, through an non-insulated duct (D7) of 168 m length and 0.3 m diameter, while decreasing its temperature to 102.6^{O}C.

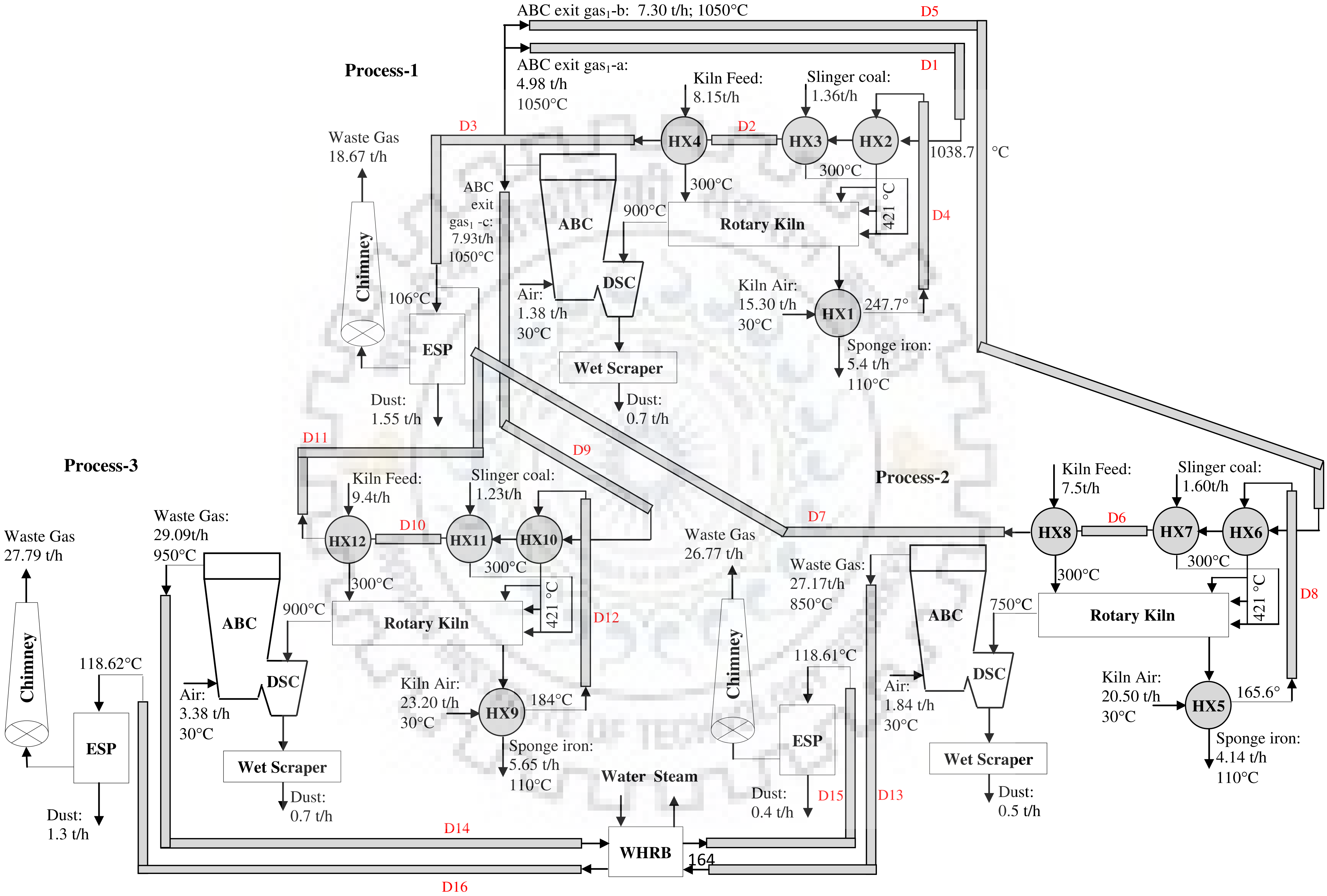


Fig. 6.36 Modified process flow diagram of Total Site through Strategy-1

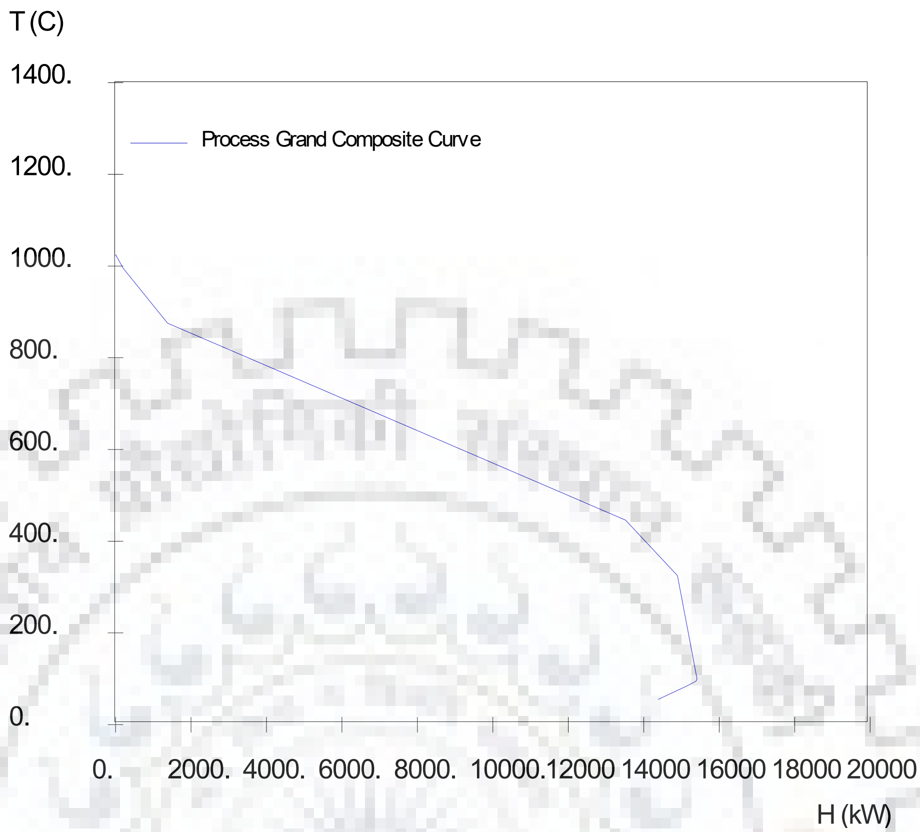


Fig 6.37 Grand composite curve for modified site through Strategy-1

Similarly, waste gas (ABC exit $gas_{1\text{-}C}$), at 1050^{O}C with CP of 2.67 kW/^{O}C, moves from ABC to kiln outlet of Process-3 through an insulated duct (D9) of length 101 m and diameter 0.3 m. After reaching kiln outlet of Process-3 it enters to a gas-gas heat exchanger, HX-10 at 1036.2^{O}C instead of 1050^{O}C shown in Fig. 6.35. Here it cools to 446^{O}C by exchanging 1573.2 kW of heat with kiln air and then enters gas-solid exchanger, HX-11, for exchanging 127.7 kW of heat with slinger coal. Waste gas exiting, HX-11, at 398.1^{O}C instead of 411.8^{O}C as shown in Fig. 6.35, moves through an insulated duct (D10) of 40 m length and 0.3 m diameter and drops its temperature to 395.9^{O}C before entering into gas-solid exchanger, HX-12 where it preheats kiln feed from 30^{O}C to 300^{O}C through exchanging load of 764.6 kW. Temperature of waste gas, thus, reduces to 109^{O}C in contrast to 125^{O}C shown in Fig.6.35. An insulated duct (D12) of small length 5 m and diameter 0.3 m is used to carry kiln air which was already preheated in gas-solid heat exchanger, HX-9. Finally, it moves from kiln inlet of Process-3 to ESP of Process-1 through a non-insulated duct (D11) of 140 m length and 0.3 m diameter dropping its temperature from 109.4^{O}C to 109.0^{O}C.

Thus, after preheating kiln air, slinger coal and kiln feed of all three processes in the site, three branches of waste gas stream are again mixed before they are combined to enter into ESP of Process-1 and acquire a new value of CP 6.79 kW/^{O}C and temperature, 106.2^{O}C instead of 125^{O}C as shown in Fig. 6.35.

An insulated duct (D13) of length 35 m and diameter 0.3 m carries waste gas having CP of 9.59 kW/^{O}C from ABC exit of Process-2 to the power generation plant at 850^{O}C. Similarly, another duct (D14) of same dimensions is used to bring the waste gas having CP of 10.38 kW/^{O}C that exits the ABC of Process-3 to the power generation plant at 950^{O}C. While passing through these ducts temperatures of waste gas streams drop to 848.86^{O}C and 948.81^{O}C. These streams of Process-2 and Process-3 are mixed in the power generation plant to acquire temperature 899.3^{O}C. Once power generation is completed waste gas splits back into two streams at 118.8^{O}C and move through ducts (D15 and D16) of 38 m length and 0.3 m diameter to ESP of their respective processes while dropping temperatures to 118.6^{O}C. The temperature profiles of all sixteen ducts are shown in Fig. 6.38. Mild steel is used for casing of the duct and for insulation ceramic fibre pipe and glass wool, at higher and lower temperatures, respectively, are placed at outer surface of the duct. Details of all sixteen ducts, used for total site integration through Strategy-1, is tabulated in Table 6.21. As gases pass through these sixteen ducts these create a certain amount of pressure drop. For all sixteen ducts pressure drops are evaluated using method discussed by Prasad [133] and tabulated in Table 6.21. These pressure drops have to be compensated either by the process or by outside means using some drives. FD fans are used for this purpose where one FD can sustain 1.26 atm pressure drop [177]. Accordingly, sixteen FD fans are required, one for each duct, to maintain the necessary pressure drops in these ducts for Strategy-1.

Table 6.21 Pressure and temperature drops in different ducts used in Strategy-1

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	52	ABC exit gas_{1-A}	ABC $exit_1$	Kiln $outlet_1$	0.0093	1050.0	1038.7
D2	40	ABC exit gas_{1-A}	Kiln $outlet_1$	Kiln $feed_1$	0.0045	501.5	497.2
D3	25	ABC exit gas_{1-A}	Kiln $feed_1$	ESP_1	0.0013	109.4	109.0
D4	5	Kiln air_1	RC $inlet_1$	Kiln $outlet_1$	0.0028	247.7	247.6
D5	126	ABC exit gas_{1-B}	ABC $exit_1$	Kiln $outlet_2$	0.0447	1050.0	1031.4
D6	42	ABC exit gas_{1-B}	Kiln $outlet_2$	Kiln $feed_2$	0.0075	352.3	350.1
D7	168	ABC exit gas_{1-B}	Kiln $feed_2$	ESP_1	0.0171	104.1	102.6
D8	5	Kiln air_2	RC $inlet_2$	Kiln $outlet_2$	0.0039	165.6	165.5
D9	101	ABC exit gas_{1-C}	ABC $exit_1$	Kiln $outlet_3$	0.0416	1050.0	1036.2
D10	40	ABC exit gas_{1-C}	Kiln $outlet_3$	Kiln $feed_3$	0.0090	398.1	395.9
D11	140	ABC exit gas_{1-C}	Kiln $feed_3$	ESP_1	0.0166	109.0	107.8
D12	5	Kiln air_3	RC $inlet_3$	Kiln $outlet_3$	0.0051	184.0	184.0
D13	35	ABC exit gas_2	ABC $exit_2$	WHRB	0.1206	850.0	848.9
D14	35	ABC exit gas_3	ABC $exit_3$	WHRB	0.1472	950.0	948.8
D15	38	ABC exit gas_2	WHRB	ESP_2	0.0421	118.8	118.6
D16	38	ABC exit gas_3	WHRB	ESP_3	0.0478	118.8	118.6

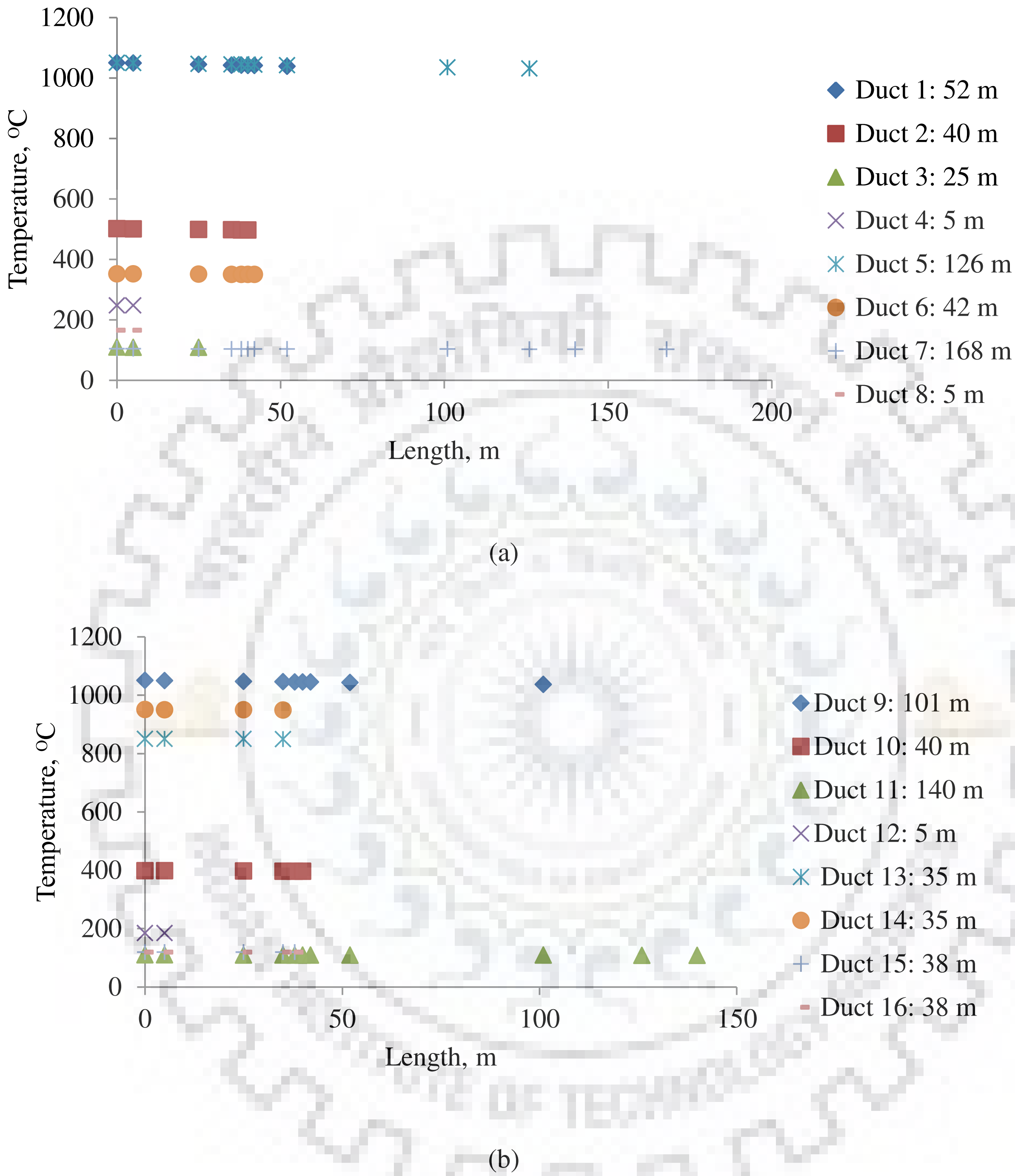


Fig. 6.38 Temperature profiles in ducts for Strategy-1

6.3.2.2 Economic aspects of total site integration for Strategy-1

Development of integrated total site through Strategy-1, requires a few new equipment. These are twelve heat exchangers, a steam turbine to generate power, sixteen ducts to carry the waste gas and kiln air and sixteen FD fans, one for each duct, to sustain the pressure drop. Amongst twelve heat exchangers nine are gas-solid heat exchangers (HX-1, HX-3, HX-4, HX-5, HX-7, HX-8, HX-9, HX-11 and HX-12) and three are gas-gas shell and tube heat exchangers (HX-2, HX-6 and HX-10). Capital costs of shell and tube heat exchangers are computed based on the approach discussed in the work of Shenoy [150] and shown in Table 6.22. The capital costs of gas-solid heat exchangers, HX-1, HX-3, HX-4, HX-5, HX-7, HX-8, HX-9, HX-11 and HX-12, are estimated as Rs. 113.5 lakh, Rs. 68.31 lakh, Rs. 96.60 lakh, Rs. 104.72 lakh, Rs. 69.73 lakh, Rs. 94.03 lakh, Rs. 117.35 lakh, Rs. 67.60 lakh and Rs. 103.02 lakh, respectively. The capital costs of sixteen ducts of total length 895 m are calculated to be Rs. 119.44 lakh. For sixteen FD fans total capital cost is computed as Rs. 1068.09 lakh. Further, electricity cost for sixteen fans is predicted as Rs 1.09 lakh/day. The costing method of gas-solid heat exchangers, ducts and FD fans is taken from the work of Prasad [133]. The capital cost of steam turbine is found to be Rs. 2430.87 lakh.

Due to 38.30% reduction in coal consumption Rs. 15226.37/h is saved. Similarly, the modification in the present design reduces water consumption by 97.89%, and accordingly Rs 37314.56/h is saved. In the Strategy-1, 5.19MW of power is generated, which include net export cost of Rs. 3.45 lakh/day. This cost is added to the savings. Thus, total net savings in terms of coal, water, power export and electricity consumption is found to be Rs 14.03 lakh/day. Based on total capital cost of new equipment and savings discounted payback period of 11.65 months is predicted. If capital recovery factor is considered with a plant life of 10 years and 10% rate of return, TAC of the modified site comes out to be Rs 3434.6 lakh. The total economic analysis for the modified design of total site integration found through Strategy-1 is shown in Table 6.22.

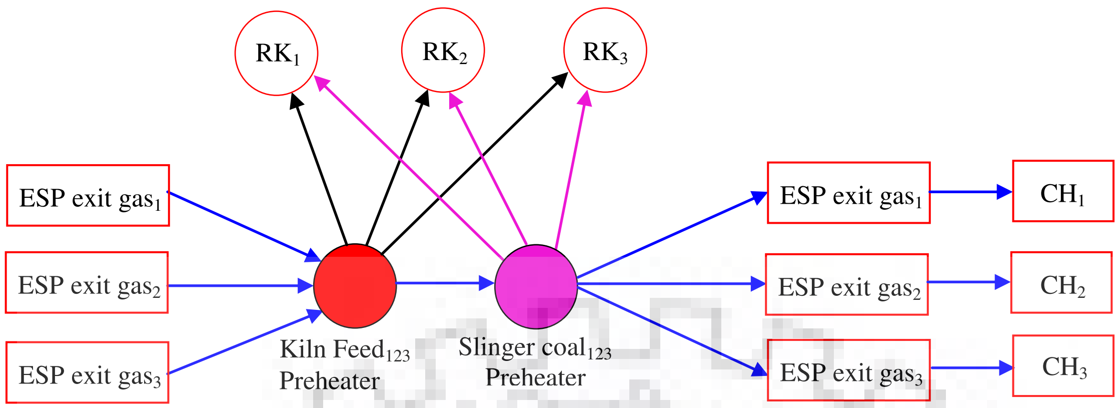
Table 6.22 Economic analysis of total site integration for Strategy-1

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	2148.93	HX-1	113.50	3434.6	5122.4	26953.6	94.0	11.65
Water	70.62	HX-2	19.31					
Electricity	397.35	HX-3	68.31					
Turbine	81.90	HX-4	96.60					
		HX-5	104.72					
		HX-6	24.27					
		HX-7	69.73					
		HX-8	94.03					
		HX-9	117.35					
		HX-10	24.57					
		HX-11	67.60					
		HX-12	103.02					
		Duct	119.44					
		FD fans	1068.09					
		Turbine	2430.87					
		Total	4521.42					

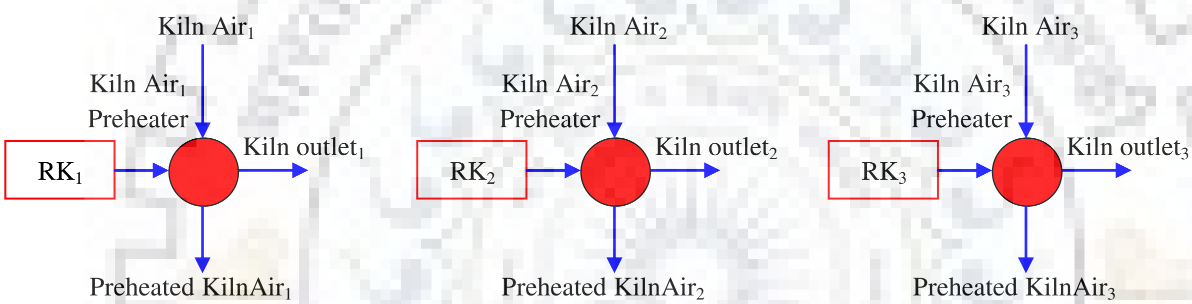
6.3.3 Strategy-2

Under Strategy-2 waste gas streams exiting the ESPs of Process-1, Process-2 and Process-3 are mixed. This mixed stream is used as hot stream to preheat kiln feed and slinger coal of all the three processes. For this purpose total kiln feed, used in all three processes, is preheated at Process-1 using mixed stream of waste gases. Further, it is also used to preheat total slinger coal of three processes at Process-1 only. Then preheated kiln feed and slinger coal is carried from Process-1 to respective processes. Further, kiln outlet stream is cooled from 1020 ^{o}C to 110 ^{o}C using kiln air of its respective process. After exchanging waste gas heat with kiln feed and slinger coal of total site, mixed waste gas stream is again splits into three different streams to move to chimney of three processes. It is noted from Fig. 4.1, Fig. 4.2 and Fig. 4.3 that, amongst these three processes only Process-3 has a unit of waste heat recovery boiler (WHRB) that recovers heat from waste gas exiting from ABC and rest of the processes cool the waste gas of ABC using water in EC. Thus, Strategy-2 proposes WHRBs, for power generation, in Process-1 and Process-2 using waste gas exiting from ABC of respective processes. The partial schematic diagram of Strategy-2 is shown in Fig. 6.39. The stream data of Strategy-2, consisting of 6 hot and 5 cold streams, are reported in Table 4.4. This Strategy, similar to Strategy-1, also considers stream data of all three processes together as single process. The number shown as a subscript in Fig. 6.39, as in Table 4.4, denotes the specific process in the total site as discussed in Section 4.4.

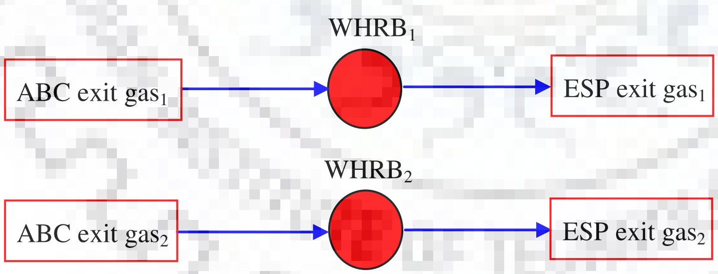
For stream data of Strategy-2, shown in Table 4.4, hot and cold utilities are found to be 0 kW and 18040.7 kW, respectively, considering ΔT_{min} as 50^{O}C. It is a threshold problem as hot utility is zero. Internal heat exchange of 4171.3 kW is observed in Strategy-2 through composite curve and GCC shown in Fig. 6.40 and Fig. 6.41, respectively. Similar to Strategy-1 a trial and error technique, as shown in Fig. 5.1, is used for computation of coal and air requirements. The results of different iterations are reported in Table 6.23.



(a) Preheating



(b) Preheating of Kiln air



(c) Power generation

Fig. 6.39 Partial schematic diagram of Strategy-2

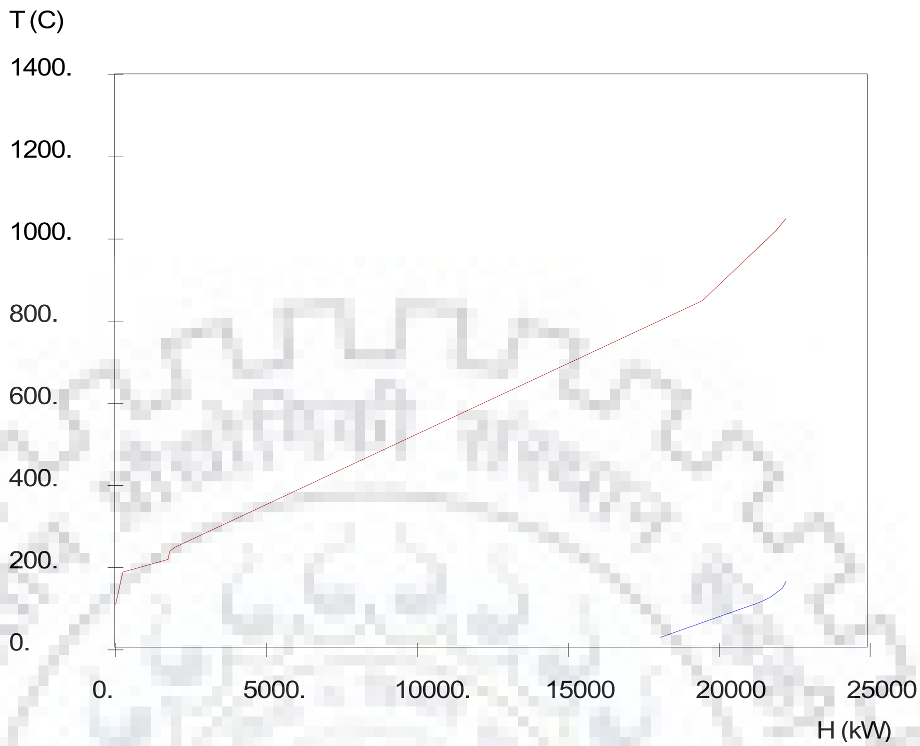


Fig 6.40 Composite curves for Strategy-2

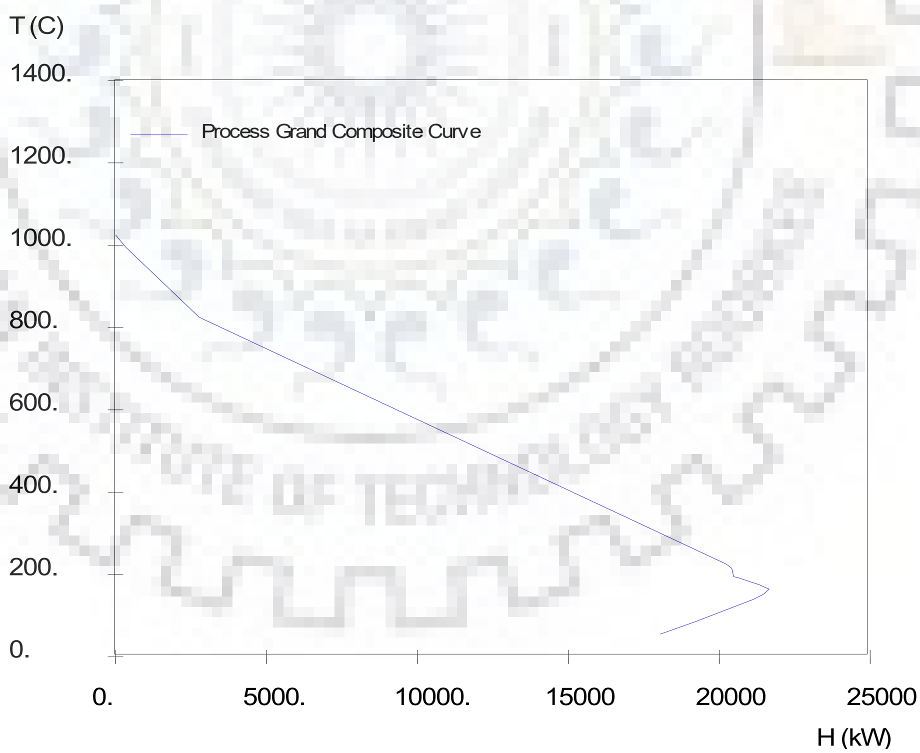


Fig. 6.41 Grand composite curve for Strategy-2

Table 6.23 Iteration results for Strategy-2

Iteration No.	Hot utility (kW)	Stage No.	Coal consumption (kg/h)	Air requirement (kg/h)
0 (existing)	-	0	17670.0	95636.5
1	0	1	15638.3	84640.4
		2	14777.9	79983.3
		3	14413.5	78010.9
		4	14259.1	77175.6
		5	14193.8	76821.8
		6	14166.1	76672.0
		7	14154.4	76608.6
		8	14149.4	76581.7
		9	14147.3	76570.3
		10	14146.4	76565.5
		11	14146.0	76563.4
		12	14145.9	76562.6
		13	14145.8	76562.2
		14	14145.8	76562.1
		15	14145.8	76562.0
		16	14145.7	76562.0
		17	14145.7	76562.0
		18	14145.7	76561.9
		19	14145.7	76561.9
2	0		13861.6	75024.3
3	0		13835.1	74880.5
4	0		13832.2	74865.1
5	0		13831.9	74863.5
6	0		13831.9	74863.3
7	0		13831.9	74863.3

For Strategy-2, total coal consumption, for all processes, at Iteration-0 (existing system) is 17.67 t/h (Process-1 = 5.36 t/h, Process-2 = 5.81 t/h, Process-3 = 6.5 t/h) whereas, kiln air is 95.64 t/h (Process-1 = 24.80 t/h, Process-2 = 33.23 t/h, Process-3 = 37.60 t/h), respectively, as shown in Table 6.23. For iteration-1 coal is computed as 15.64 t/h using Equation 4.9 and corresponding amount of air is found as 84.64 t/h using ratio of air to coal as 5.41 maintained in iteration-0. As discussed in Section 6.3.2, this revised value of air is used in second stage of iteration-1 to compute amount of coal, which is found to be 14.78 t/h as shown in Table 6.23. Corresponding amount of air is 79.98 t/h. Similarly, different stages of iteration-1 are performed till the values of coal and air become equal in two consecutive stages. Final amounts of coal and air, in iteration-1, are 14.15 t/h and 76.56 t/h. The reduction in coal and air consumption is due to total site heat integration of sponge iron cluster through Strategy-2 where which kiln feed and slinger coal of total site are preheated from 30°C to 150°C. Along with this, kiln air in each process is also preheated to maximum possible temperature using waste heat available from kiln outlet of its respective process. Therefore, the amount of coal required to preheat feed material inside the kiln is reduced. Now, revised values of coal and air i.e. 14.15 t/h and 76.56 t/h are considered to modify the stream data, shown in Table 4.4, which are used to compute coal consumption in iteration-2. Following similar procedure for computing coal and air consumptions as carried out in iteration-1, results for other iterations are found in different stages as shown in Table 6.23. Final amounts of coal and air are 13.83 t/h (Process-1 = 4.20 t/h, Process-2 = 4.55 t/h, Process-3 = 5.08 t/h) and 74.86 t/h (Process-1 = 19.41 t/h, Process-2 = 26.01 t/h, Process-3 = 29.44 t/h), respectively, and these values maintain the ratio of 5.41. Therefore, 21.72% of coal is reduced in comparison to the existing process.

HEN for the modified design of total site integration through Strategy-2 is shown in Fig. 6.42. It includes five heat exchangers. The temperatures and heat loads shown in HEN (Fig. 6.42) are based on values of final iteration reported in Table 6.23.

Stream (H4) shown in Fig. 6.42 is a mixed stream of waste gases exiting from ESP of all three processes. Temperature of mixed stream of waste gas is 220°C. It is splitted into two streams to maintain ΔT_{min} at 50°C in the HEN shown in Fig. 6.42. One part of waste gas, having CP 22.41 kW/°C, exchanges heat with total site kiln feed used in all three processes (C1) to get cooled from 220°C to 176.3°C through HX-1, which has load of 979.3 kW. It is a gas-solid exchanger where kiln feed of total site is preheated from 30°C to 150°C. Another part of waste gas with CP

5.6 kW/^{O}C enters to a gas-solid exchanger, HX-2, which has a load of 244.9 kW. It is employed for preheating slinger coal stream (C2), which is total slinger coal used in all three processes, from 30^{O}C to 150^{O}C. As a result of heat exchange in HX-1 and HX-2 the mixed stream of waste gas drops from 220^{O}C to 176.3^{O}C. Further, gas-solid heat exchanger, HX-3, exchanges 953.1 kW heat between kiln outlet stream (H1) and kiln air (C3) of Procees-1. Hence, kiln outlet stream of Process-1 cools from 1020^{O}C to 110^{O}C whereas; kiln air of Process-1 heats up from 30^{O}C to 204.1^{O}C. Similarly, exchangers HX-4 and HX-5 are gas-solid heat exchangers with heat loads of 795.2 kW and 1022.4 kW, respectively. In these exchangers kiln outlet streams (H2 and H3) of Process-2 and Process-3 are cooled from 1020^{O}C to 110^{O}C using kiln air streams (C4 and C5) of respective processes available at 30^{O}C. As a result of it, kiln air is heated up to 138.8^{O}C and 153.5^{O}C in Process-2 and Process-3, respectively, as shown in Fig 6.42. Thus, HEN shown in Fig. 6.42 includes five gas-solid heat exchangers, in which no heat exchanger violates ΔT_{min}.

Difference of internal heat change, between hot and cold streams, can be observed before and after total site integration through Strategy-2. These are 4171.3 kW and 3994.9 kW as shown in composite curve (Fig 6.40) and HEN (Fig 6.42), respectively. It is due to the fact that after design modification through Strategy-2 amounts of kiln inlet streams and waste gas generation of total site reduce and hence, net heat contents of these streams also reduce. For example, the heat capacity flow rate of total site kiln feed (stream No. C1) is 9.07 kW/^{O}C and 8.16 kW/^{O}C, before and after total site integration, respectively. Along with this, as shown in Stream data of Strategy-2 in Table 4.4, waste gas from ESP exit of total site (stream no. H4) is aimed to cool from 220^{O}C to 151.7^{O}C whereas, after total site integration, it is cooled to only 176.3^{O}C as shown in Fig. 6.42.

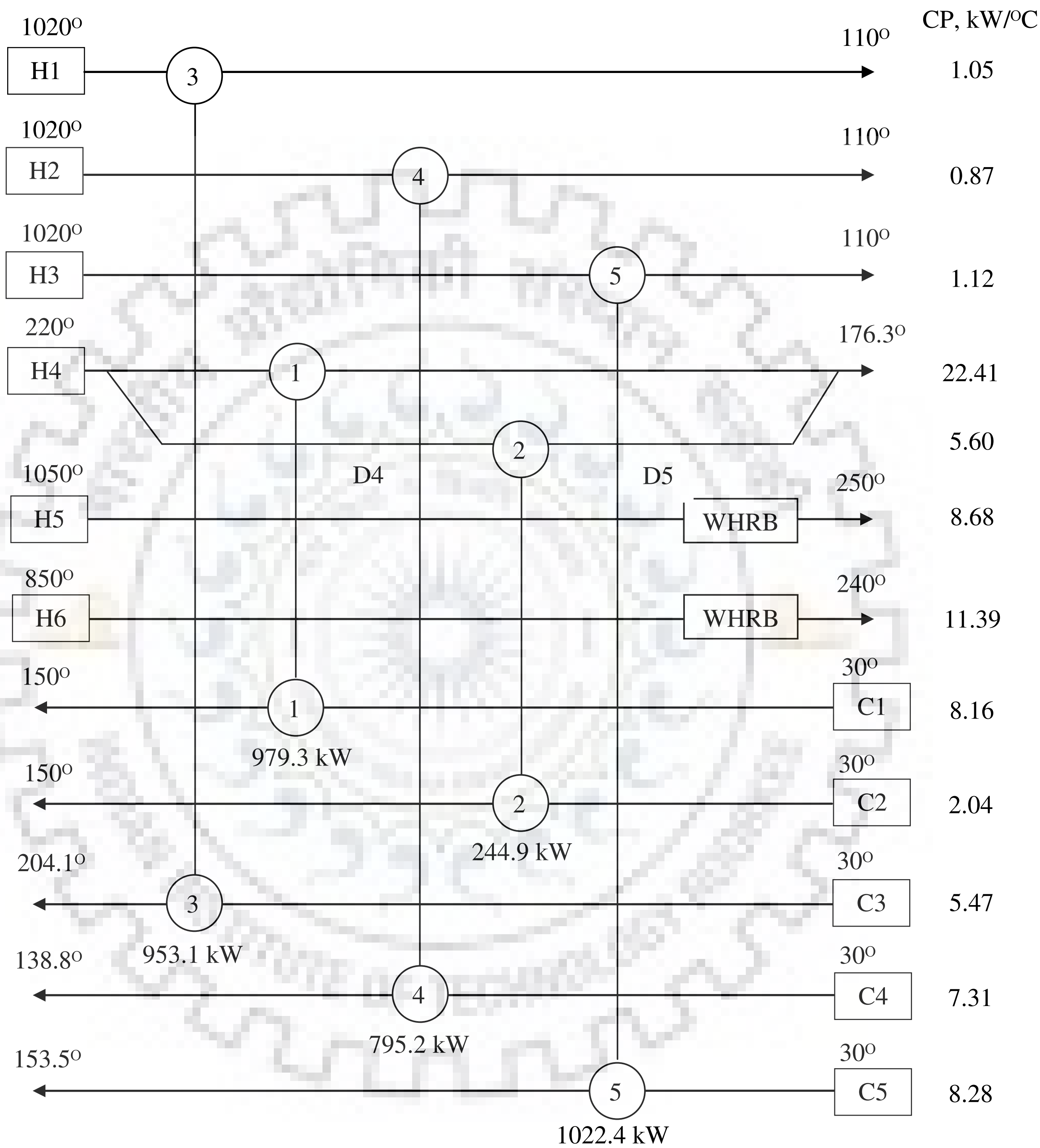


Fig. 6.42 Heat Exchanger Network for Strategy-2

Based on the guidelines described in Section 4.4, the modified PFD of total site through Strategy-2 is drawn and shown in Fig. 6.43. Here, waste gas generation is reduced to 93.44 t/h which is 38.4% less in comparison to the existing system, shown in Fig. 4.5, where it is 151.67 t/h.

Strategy-2 proposes WHRBs, for power generation in Process-1 and Process-2 using waste gas exiting ABC of respective processes. Total power produced through heat of waste gas is predicted as 2.59 MW and 2.66 MW for Process-1 and Process-2, respectively. In fact, for this computation GCC is not used as discussed for Strategy-1. Total heat available in the waste gases of individual processes is used for preheating the water, boiling and superheating. Consequently, the produced superheated steam at 700^{O}C and 300 bar is used for power generation. The power produced is assumed to be exported. Total water requirement in the existing site is 705.94 t/h as shown in Fig. 4.5. The modification in the present design eliminates water requirement in rotary cooler of all three processes and in EC of Process-1 and Process-2. For power generation 25.84 t/h water is required and hence, 96.34% of water is reduced in comparison to the existing process.

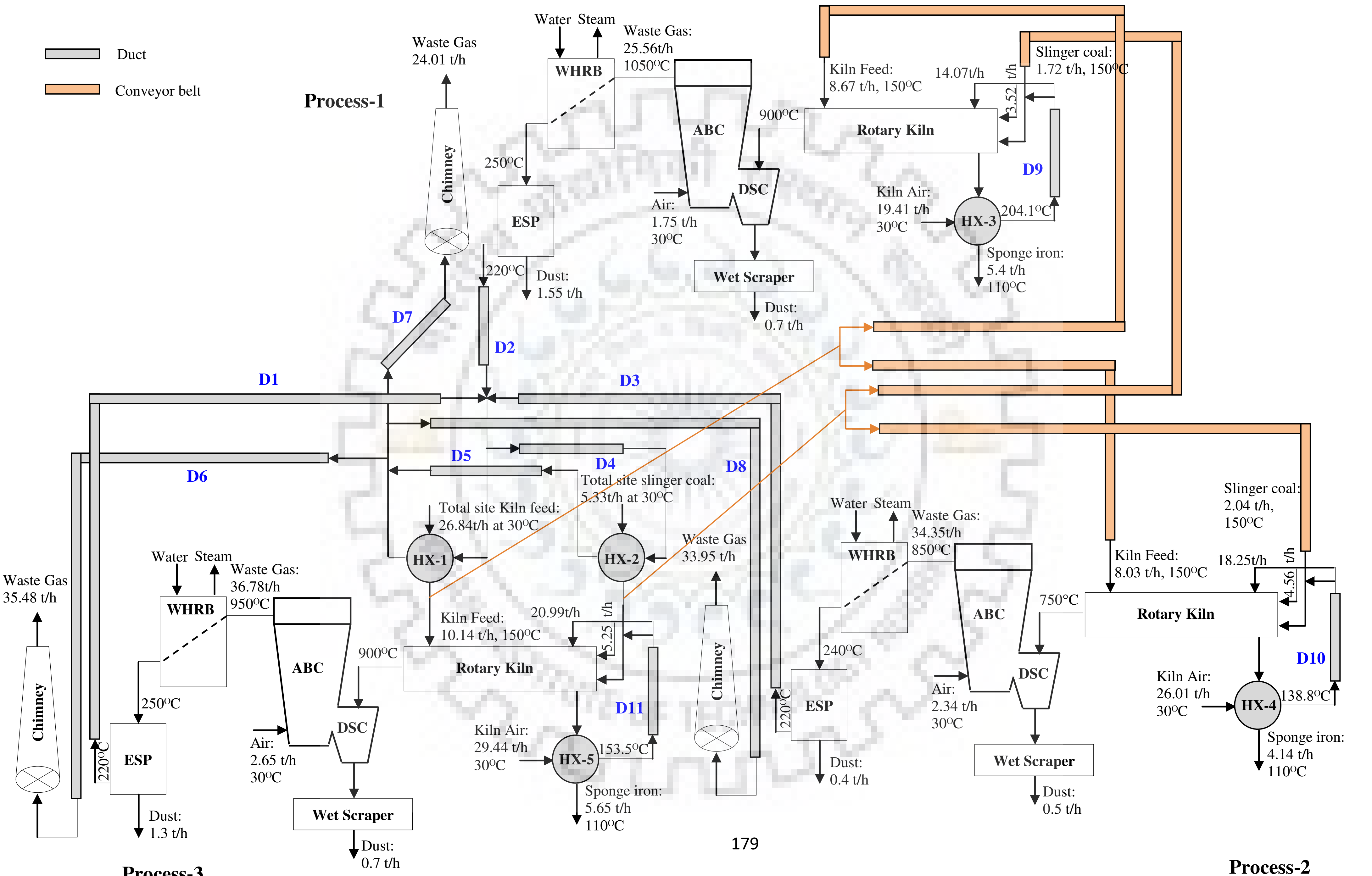


Fig. 6.43 Modified process flow diagram of total site through Strategy-2

6.3.3.1 Operational aspects of total site integration for Strategy-2

Strategy-2 requires eleven numbers of ducts, eight for carrying waste gas streams from exits of ESP to other sections of processes and three for carrying kiln air to kiln outlet in individual processes. Initially, an insulated duct (D1) of length 25 m and diameter 0.3 m is employed to transfer waste gas, having CP of 7.52 kW/^{O}C, from ESP of process-3 to kiln inlet of the same process. While passing through this duct temperature of waste gas drops down from 220^{O}C to 219.8^{O}C as shown in Table 6.24. Similarly, two insulated ducts (D2 and D3) of different lengths 140 m and 79 m are employed to carry waste gas streams from ESP of Process-1 and Process-2 to kiln inlet of Process-3 as shown in Fig. 6.43. These three streams from ESP of Process-1, Process-2 and Process-3 are combined together. Further, combined stream of waste gas with CP 28.02 kW/^{O}C and temperature 219.3^{O}C splits into two streams as shown in Fig. 6.42 (HEN) and Fig. 6.43. First stream of waste gas with CP of 22.41 kW/^{O}C enters into a gas-solid heat exchanger, HX-1, at 219.3^{O}C where kiln feed of all three processes together is preheated from 30^{O}C to 150^{O}C while exchanging heat of 979.3 kW. However, second stream of waste gas with remaining value of CP, i.e., 5.60 kW/^{O}C is transferred to kiln outlet of Process-3 through an insulated duct (D4) of length 40 m and diameter 0.3 m where it enters into a gas-solid heat exchanger, HX-2, at 218.9^{O}C instead of 220^{O}C as shown in Fig. 6.42. In this exchanger slinger coal of all three processes together is preheated from 30^{O}C to 150^{O}C while exchanging heat of 244.9 kW. After exchanging heat in HX-2 second stream of waste gas is carried to kiln inlet of Process-3 through an insulated duct (D5) of length 40 m and diameter 0.3 m where both streams of waste gas are mixed to acquire temperature of 175.4^{O}C. Waste gas now splits into three different streams as shown in Fig. 6.43. The first stream having CP of 11.44 kW/^{O}C enters an insulated duct (D6) of length 25 m and diameter 0.3 m at 175.4^{O}C, which directs waste gas to chimney of Process-3 at 175.3^{O}C. The second stream of waste gas with value of CP as 7.47 kW/^{O}C is carried through an insulated duct (D7) of dimension, 140 m length and 0.3 m diameter. While passing through this duct temperature of waste gas drops to 174.2^{O}C and then it enters the chimney of Process-1. Further, an insulated duct (D8) of length 79 m and diameter 0.3 m carries the third stream of waste gas having CP of 10.69 kW/^{O}C to chimney of Process-2 at 174.9^{O}C. Kiln air of Process-1 that was already preheated in gas-solid heat exchanger, HX-3, is passed through an insulated duct (D9) of small length 5 m and diameter 0.3 m from cooler inlet to kiln outlet. Consequently kiln air temperature drops from 204.1^{O}C to 204^{O}C. Similarly two insulated ducts (D10 and D11) of lengths 5 m are employed to carry kiln air streams that were already preheated in gas-solid heat exchangers, HX-4 and HX-5, from cooler inlet to kiln outlet of Process-2 and Process-3, respectively. Consequently, kiln air temperature, in Process-2, drops from 138.8^{O}C to 138.7^{O}C whereas in Process-3 the temperature drop is negligible. The temperature profile for all these ducts, shown in Fig. 6.44, are predicted, similar to Strategy-1, using the method discussed by Prasad [130]. Mild steel is used for casing of the duct and for insulation glass wool is placed at outer surface of the duct.

Further, it is observed that certain pressure drops are created in eleven ducts when gases traverse through these. These are evaluated using method discussed by Prasad [133] and also tabulated in Table 6.24. These pressure drops are compensated using eleven FD fans, one for each duct, where one FD can sustain 1.26 atm pressure drop [177].

Table 6.24 Pressure and temperature drops in different ducts used in Strategy-2

Duct No	Length (m)	Stream	From	To	ΔP (atm)	T_i (^oC)	T_o (^oC)
D1	25	ESP exit gas$_3$	ESP$_3$	Kiln feed$_3$	0.0569	220.0	219.8
D2	140	ESP exit gas$_1$	ESP$_1$	Kiln feed$_3$	0.152	220.0	218.4
D3	79	ESP exit gas$_2$	ESP$_2$	Kiln feed$_3$	0.163	220.0	219.4
D4	40	ESP exit gas$_{123}$-b	Kiln feed$_3$	Kiln outlet$_3$	0.028	219.3	218.9
D5	40	ESP exit gas$_{123}$-b	Kiln outlet$_3$	Kiln feed$_3$	0.025	175.2	174.8
D6	25	ESP exit gas$_3$	Kiln feed$_3$	CH$_3$	0.050	175.4	175.3
D7	140	ESP exit gas$_1$	Kiln feed$_3$	CH$_1$	0.137	175.4	174.2
D8	79	ESP exit gas$_2$	Kiln feed$_3$	CH$_2$	0.147	175.4	174.9
D9	5	Kiln air$_1$	RC inlet$_1$	Kiln outlet$_1$	0.004	204.1	204.0
D10	5	Kiln air$_2$	RC inlet$_2$	Kiln outlet$_2$	0.006	138.8	138.7
D11	5	Kiln air$_3$	RC inlet$_3$	Kiln outlet$_3$	0.007	153.5	153.5

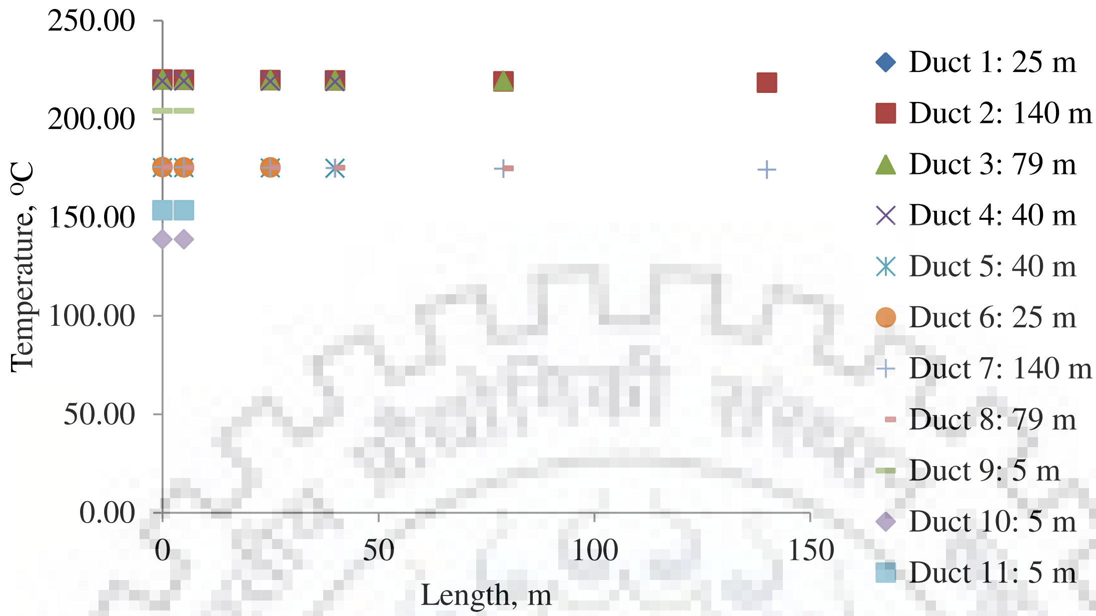


Fig. 6.44 Temperature profiles in ducts for Strategy-2

Along with eleven ducts four covered conveyor belts are required to carry hot streams of solid from preheater to kilns of different processes as the usual belt, used in the processes, can sustain temperature up to 120°C only. First, hot kiln feed (26.84 t/h) coming out from kiln feed preheater, HX-1, is divided into three streams (8.67 t/h, 8.03 t/h and 10.14 t/h) according to requirements of the individual process. A covered conveyer belt of length 143 m and width 0.5 m is used to carry 8.67 t/h of kiln feed from the kiln preheater in Process-3 to the kiln inlet of Process-1. Another conveyor belt of 75 m length and width 0.5 m is employed for transferring 8.03 t/h kiln feed to the kiln of Process-2. Remaining 10.14 t/h of preheated kiln feed directly enters into the kiln inlet of Process-3. Similarly, hot slinger coal (5.33 t/h) coming out from preheater, HX-2, is divided into three streams. Amongst these two streams reach to kiln outlet of Process-1 and Process-2 through covered conveyer belts of 120 m and 75 m, respectively. These belts carry 1.72 t/h and 2.04 t/h of slinger coal whereas remaining 1.57 t/h of slinger coal directly enters into the kiln of Process-3.

6.3.3.2 Economic aspects for Strategy-2

Total economic analysis for the modified design of total site integration found through Strategy-2 is shown in Table 6.25. The modified design for total site of sponge iron processes through Strategy-2 requires following new equipment: five gas-solid heat exchangers, two steam turbines, eleven insulated ducts to carry the waste gas and kiln air, eleven FD fans and four covered conveyor belts. Capital costs of gas-solid heat exchangers, stream turbines, ducts and FD fans are computed in the similar line as discussed for Strategy-1 under Section 6.3.2.2. These costs are shown in Table 6.25. Further, electricity cost for eleven fans is predicted as Rs 0.76 lakh/day. The capital cost of four conveyor belts of total length of 413m is computed as Rs. 43.55 lakh. Total coal consumption reduced by 21.72% in comparison to the existing system, which saves Rs. 8635.72/h. Similarly, the modification in the present design reduces the water consumption by 96.39%, which saves Rs 37262.06/h. In Strategy-2, 2.59 MW and 2.66 MW of power is generated in Process-1 and Process-2, respectively, which is having net export cost of Rs. 3.71 lakh/day. This cost is added to the savings of the process. Thus, the total net savings in terms of coal, water, power export and electricity consumption is found to be Rs 13.51 lakh/day. If capital recovery factor is considered with a plant life of 10 years and 10% rate of return, TAC of the modified site comes out to be Rs 3839.2 lakh. Considering total capital cost of new equipment and savings the discounted payback period of 10.46 months is predicted as shown in Table 6.25.

Table 6.25 Economic analysis of total site integration for Strategy-2

Operating cost (Lakh/year)		Capital investment (Lakh)		TAC (Lakh/ year)	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
Commodity	Cost	Item	Cost					
Coal	2726.27	HX-1	114.96	3839.2	4930.16	26135.2	106.0	10.46
Water	122.22	HX-2	74.12					
Electricity	278.17	HX-3	113.50					
Turbine	82.79	HX-4	104.71					
		HX-5	117.35					
		Duct	77.87					
		FD fans	766.38					
		Conveyor belts	43.55					
		Turbine-1	1213.65					
		Turbine-2	1243.56					
		Total	3869.64					

6.4 COMPARATIVE ANALYSIS OF STRATEGY-1 AND STRATEGY-2

As discussed in Section 6.3 three strategies are proposed for energy conservation in coal based sponge iron cluster. Strategy-0 is developed based on the best scheme for Process-1, Process-2 and Process-3 found in Section 6.2. As discussed in Section 6.3.1 there is no relative advantage of conventional total site integration through Strategy-0, which is more suitable to integrate heat within the site of different processes. Thus, two different strategies Strategy-1 and Strategy-2 are proposed based on the modified approach of total site integration developed by Ramanaiah and Khanam [142] to conserve energy in total site of plants of similar type where conventional methods are not applicable. Detailed results of these two strategies are discussed in Section 6.3. Here these two heat integration strategies, Strategy-1 and Strategy-2, are examined to select best strategy for total site of sponge iron cluster. For this purpose, results of Strategy-1 and Strategy-2 are summarized in Table 6.26 and compared with the existing system based on all economical parameters such as coal consumption, water requirement, operating cost, capital investment, total annual cost, total savings and payback period. Efficacies of these two strategies are also computed while comparing results of these based on waste gas generation, energy ratio (ER) and %heat recovery.

Table 6.26 shows that Strategy-1 gives 38.3% savings in coal consumption whereas, 21.72% coal savings is observed in Strategy-2. Reduction in coal consumption is more for Strategy-1 in comparison to Strategy-2. This is explained as: In Strategy-1 kiln feed is preheated, outside the kiln, from 30^{O}C to 300^{O}C utilizing 2,448 kW of waste heat whereas in Strategy-2 it is only from 30^{O}C to 150^{O}C involving 1,088 kW of waste heat. Consequently, for Strategy-1 and Strategy-2 kiln feed is heated from 300^{O}C and 150^{O}C, respectively, to reaction temperature inside the kiln using coal. For this purpose, Strategy-1 utilizes 1,820 kg/h of coal; whereas, 2,199.8 kg/h is consumed in Strategy-2, which is 20.9% more. On the other hand, this value of coal consumption is 2,503.2 kg/h in the existing cluster. Similarly, other inlet streams i.e. slinger coal and kiln air require less coal inside the kiln in Strategy-1 as compared to Strategy-2. In fact, in Strategy-1 total 13,850.4 kW of waste heat is utilized to preheat the kiln inlet streams; whereas, in Strategy-2 it is only 4,171.3 kW. Thus, coal consumption to heat these streams inside the kiln reduces more in Strategy-1 than that in Strategy-2. The % coal savings, through both strategies in comparison to the existing system, is shown in Fig. 6.45.

Strategy-1 saves 97.9% water, as shown in Table 6.26, in total site as compared to the existing system, which means that Strategy-1 will save 54.48 MT of water during the plant life of 10 years. Similarly, as shown in Table 6.26, Strategy-2 saves 96.34% water in total site. Thus savings offered by Strategy-1 are higher than that through Strategy-2 although the difference is very small. The % water savings, through both strategies in comparison to existing system, is shown in Fig. 6.45.

These results further indicate that both strategies produce significantly less waste gas in comparison to existing system by proper total site integration of sponge iron cluster. Strategy-1 produces 73.2 t/h waste gas whereas Strategy-2 releases 93.4 t/h in comparison to existing sponge iron cluster where it is 151.7 t/h as shown in Table 6.26. Thus, Strategy-1 and Strategy-2 produce 51.72% and 38.4%, respectively, less waste gas in comparison to the existing system due to less coal consumption. The primary fossil fuel energy consumption of India by coal gives strong concern on greenhouse gas emissions and climate change. Reduction in coal (energy) consumption through total site integration makes sponge iron processes more environment friendly while decreasing waste gas released to the atmosphere. The % reduction in waste gas generation through both strategies in total site is also shown graphically in Fig. 6.45.

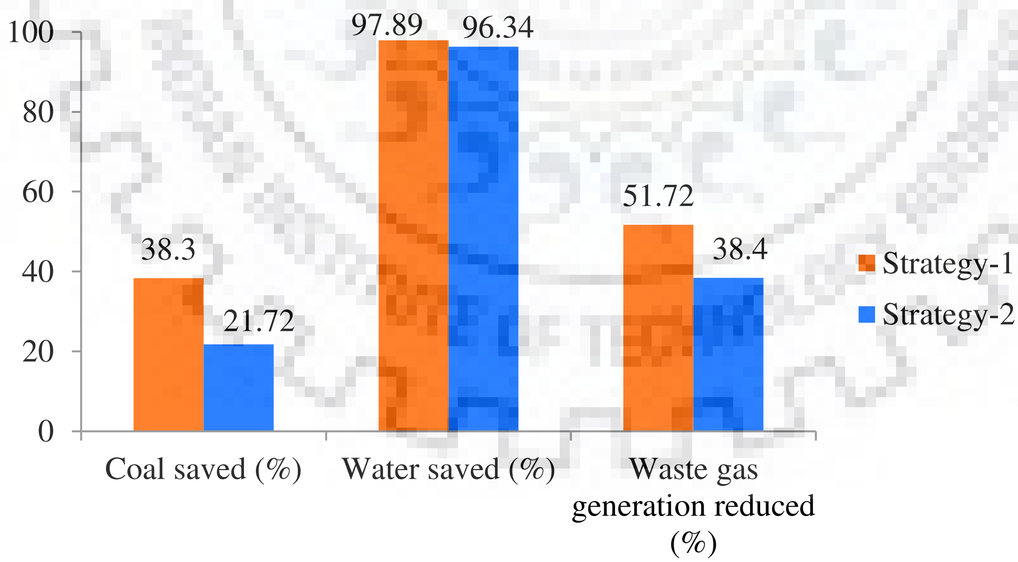


Fig. 6.45 Coal-water savings and reduction of waste gas generation in Strategy-1 and Strategy-2

As discussed in Section 6.2, to estimate the efficiency of proposed design modifications ER is computed and shown in Table 6.26. From the material and energy balances it was computed that all three processes combined together as a total site requires 63,642 kW of theoretical energy for reduction reactions to continue in kiln whereas, in actual practice 1,11,710.2 kW of energy is consumed, thus ER value is 1.76 under existing conditions, as shown in Table 6.26. After total site integration through Strategy-1 actual energy consumed is reduced to 68,927.3 kW giving ER value 1.08, which is almost equal to 1.0. Similarly, after total site integration through Strategy-2 actual energy consumed is reduced to 87445.7 kW giving ER value 1.37. That means, the excess energy consumption which is 76%, prior to total site integration, is reduced to 8% and 37% through Strategy-1 and Strategy-2 respectively. Hence actual energy consumption, in Strategy-1, is more close to theoretical energy required than in Strategy-2. The variation in excess energy consumption, in both strategies, is shown graphically in Fig. 6.46.

After final design it is found that Strategy-1 recovers 99.8% of waste heat available in the modified site whereas, Strategy-2 recovers 86%. It is computed as: total heat that can be possibly recovered in the existing site is 25,602.7 kW and Strategy-1 is able to recover it completely whereas Strategy-2 recovers only 22,211.5 kW. After energy conservation through Strategy-1, total heat that can be recovered from modified site is reduced to 15,481.9 kW, out of which, 15,448.2 kW (99.8%) of heat is recovered. Similarly, through Strategy-2 total recoverable heat is reduced to 20,768.1 kW where 17,890.2 kW (86%) is recovered. Also, through Strategy-1 overall amount of energy wasted in the cluster, in terms of coal consumption, is reduced from 43% to 7.6%. As more % of heat is recovered the requirement of coal is reduced more in Strategy-1 than in Strategy-2. The %heat recovery, in total site, through design modifications Strategy-1 and Strategy-2 is given in Table 6.26 and also shown graphically in Fig. 6.46.

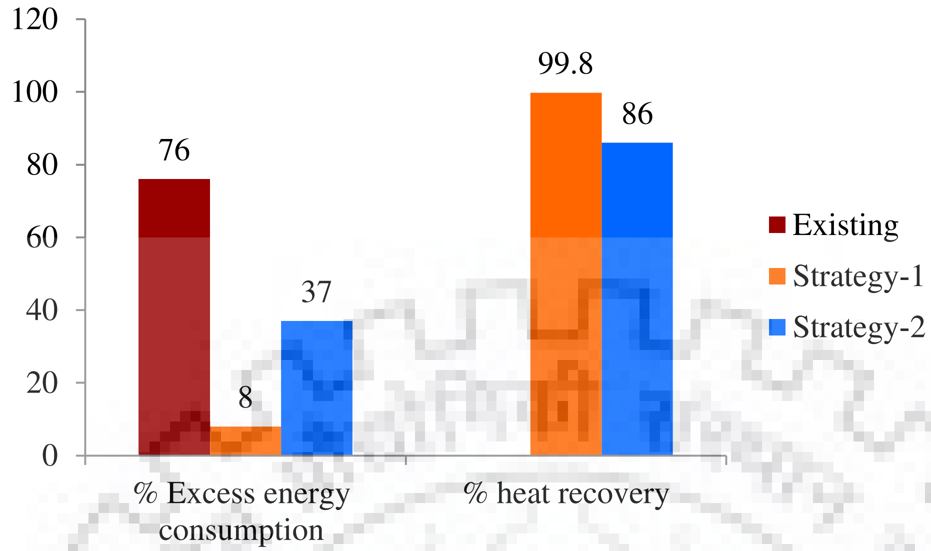


Fig. 6.46 % Excess energy consumption and % heat recovery in Strategy-1 and Strategy-2

It is interesting to note that Strategy-1 gives maximum savings and also involve minimum operating cost. TAC that is presented in Table 6.26 shows that it is mainly controlled by annual operating cost and for Strategy-1 and it is 18.92% less than that for Strategy-2 as coal and water consumptions are less in Strategy-1 than in Strategy-2. As shown in Table 6.26, the capital investment required for design modification through Strategy-1 is 4516.6 lakh whereas the design modification through Strategy-2 requires capital investment of 3869.64 lakh. Both Strategies offer huge savings where Strategy-1 and Strategy-2 give savings of 5122.41 lakh/year and 4883.15 lakh/year, respectively. The payback period for Strategy-1 is 11.65 months and that of Strategy-2 is 10.46 months. Although payback period is little more for Strategy-1, it offers more savings than that of Strategy-2. Thus, based on energy consumption, water requirement, waste gas generation, TAC and savings Strategy-1 is selected as best heat recovery option. The variations in capital investment, savings and payback period for both strategies in sponge iron cluster are shown in Fig. 6.47.

Thus based on energy consumption, water requirement, waste gas generation, energy ratio (ER) and %heat recovery and based on economic analysis discussed in this section Strategy-1 is selected as better heat recovery option over strategy-2 for sponge iron industrial cluster.

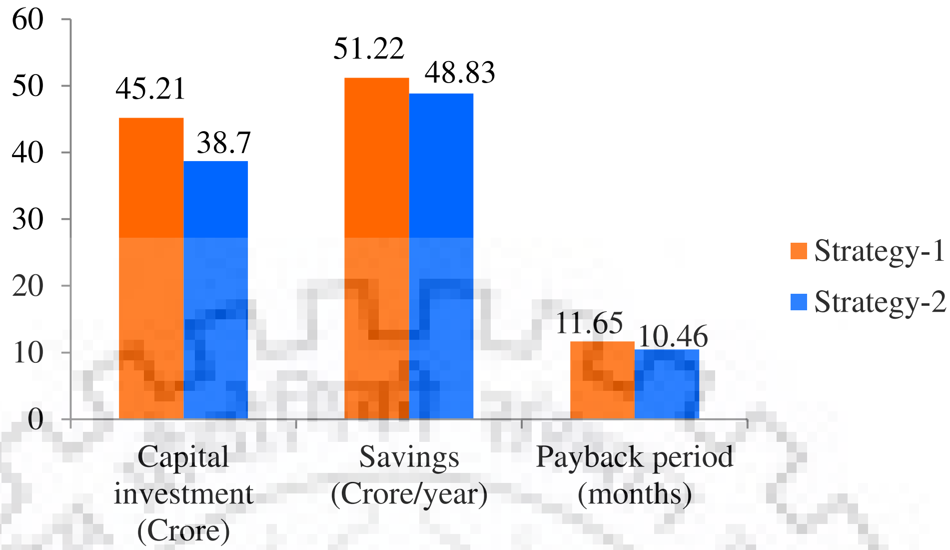


Fig. 6.47 Economic analysis of Strategy-1 and Strategy-2

6.5 COMPARATIVE ANALYSIS OF INDIVIDUALLY INTEGRATED SYSTEM WITH TOTAL SITE INTEGRATED SYSTEM

As discussed under Section 4.3, two design modifications, Scheme-1 and Scheme-2, are proposed for heat integration in individual processes, Process-1, Process-2 and Process-3, of sponge iron cluster. Results of these two schemes are compared in Section 6.2 where Scheme-2 is selected as best design for individually integrated systems. Here, Table 6.26 also summarizes Scheme-2 results of all three processes together for the purpose of comparing its efficacy with Strategy-1 and Strategy-2, which are proposed for total site integration.

Table 6.26 shows that Scheme-2 requires total 11.79 t/h of coal for all three processes combined together whereas Strategy-1 and Strategy-2 requires 10.9 t/h and 13.83 t/h of coal, respectively.

Coal savings offered by Scheme-2 (33.27%) are more than the savings offered by Strategy-2 (21.72%) in comparison to existing system whereas Strategy-1 (38.3%) offers more than Scheme-2. Similarly, Scheme-2 consumes 19.2 t/h of water, which is again more than the requirement of Strategey-1, though lower than that of Strategy-2, as shown in Table 6.26. Accordingly, Scheme-2 generates more waste gas than Strategy-1 but less than Strategy-2. From values of ER, shown in Table 6.26, it can be said that the actual energy consumption required in Scheme-2 is more close to theoretical energy consumption required in Strategy-2 whereas, Strategy-1 offers better ER as compared to Scheme-2. Although Scheme-2 requires less capital cost, it also offers less savings than that of strategies of total site integration. Scheme-2 offers only 3796.92 lakh/year savings whereas Strategy-1 and Strategy-2 offer 5122.41 lakh/year and 4883.15 lakh/year, respectively.

Thus it can be concluded that total site integrated systems offer better results than individually integrated systems although it depends upon the strategies one select. Here, individually integrated systems modified through Scheme-2 offers better results, in terms of coal and water savings, than total site integrated system modified through Strategy-2 but offers less savings than Strategy-2. However, as shown in Table 6.26, results offered by Strategy-1 that is applied on total site integrated system are significantly better than Scheme-2 which is applied on individually integrated systems. Hence, for sponge iron cluster, considered in the present work, integrating as a total site is more profitable than individually integrated systems.

The four design modifications proposed in the current work through both energy integration schemes (Scheme-1 and Scheme-2) and total site integration strategies (Strategy-1 and Strategy-2) offer better energy savings. All the proposed modifications require same level of disturbance during start-up and shut down but energy integration schemes offer minimum process modifications in comparison to total site integration strategies as they require set up of power generation plants. However, once these are set up ease of operation is similar in all the four cases. Savings offered by total site integration strategies are better than that offered by single site integration schemes. Hence Strategy-1 is considered as the best design modification for coal based sponge iron cluster.

Table 6.26 Comparative analysis of Strategy-1 and Strategy-2

Case	Resource Consumption (t/h)		Operating cost (Lakh/ year)	Capital investment (Lakh)	TAC (Lakh/ year)	Waste gas generated (t/h)	Energy ratio (ER)	(%) Heat recovery	Savings (Lakh/ year)	NPV (Lakh)	IRR (%)	DPP (months)
	Coal	Water										
Existing (Total site)	17.67	705.94	-	-	-	151.67	1.76	-	-	-	-	-
Scheme-2 (All three processes)	11.79	19.20	2712.94	2011.90	3040.4	91.56	1.17	29.7	3795.92	21312.4	162.0	7.0
Strategy-1	10.90	14.93	2698.80	4521.42	3434.6	73.23	1.08	99.8	5122.41	26953.6	94.0	11.65
Strategy-2	13.83	25.84	3209.46	3869.64	3839.2	93.44	1.37	86.1	4883.15	26135.2	106.0	10.46

6.6 COMPARATIVE ANALYSIS OF RESULTS OF STRATEGY-1 WITH THAT OF PUBLISHED LITERATURE

The conventional method, developed for total site integration, is used to handle clusters of different type plants only and these methods cannot be used directly for energy integration in site of similar type plants. Further, it appears from literature that total site integration of sponge iron cluster is not carried out. Thus, advantages of Strategy-1, which has been selected as the best proposed design in the present work for energy conservation in sponge iron cluster, cannot be compared with the available published literature. However, some investigators, although few, developed energy conservation scenarios for sponge iron plants of different capacities using the principles of process integration. Results of these studies can be compared with that of Strategy-1 on the basis of % reduction in coal consumption, water consumption and waste gas generation and also on the basis of % heat recovery, ER and savings/t of sponge iron to observe the benefits of total site integration. Advantages of Strategy-1 over the published literature are summarized in the Table 6.27. It also shows the production capacity of the plant that is considered for energy integration and the savings/t of sponge iron for the purpose of comparison of Strategy-1 with the published literature as different plants operate at different capacities.

Kumar and Khanam [83] proposed preheating of kiln air to 80^{O}C by kiln outlet with the use of water bath. The proposed design of heat integration reduces coal consumption only by 7.2%, which is very low as compared to the reduction found by total site integration through Strategy-1 in the present work. Though reduction in water consumption is significant Strategy-1 saves more water in comparison to the proposed design. The % reduction of waste gas generation in comparison to existing system is just 15.3% through the modification authors proposed whereas Strategy-1 reduces 51.7%. It is interesting to note that though capital investment, as shown in Table 6.27, is much less in design proposed by authors as compared to Strategy-1, TAC and payback period are almost equal. Strategy-1 offers huge savings than published design, which offers only 1.33 Lakh/year/t of SI with payback period of 11.3 months, whereas Strategy-1 offers savings of 19.52 Lakh/year/t of SI with payback period of 10.73 months. Based on these factors total site integration through Strategy-1 can be considered as the best design modification over Kumar and Khanam [83].

Prasad et al. [130] proposed preheating of feed materials to kiln from 30^{O}C to 110^{O}C and air from 30^{O}C to 385.7^{O}C using waste gas coming from ABC. Authors proposed design modifications resulted in 30.5% and 72.6% reductions in coal and water consumptions in comparison to the existing process, which are significantly less than that of Strategy-1. The % reduction of waste gas generation in comparison to existing system is only 27.4% through authors proposed modification whereas Strategy-1 reduces 51.7%. Though capital investment and payback period, as shown in Table 6.27, are less in design proposed by authors Strategy-1 offers huge savings. Strategy-1 offers savings of 19.52 Lakh/year/t of SI with payback period of 10.73 months whereas authors design modification proposed offers only 2.64 Lakh/year/t of SI with payback period of 6.95 months. Based on these factors total site integration through Strategy-1 can be considered as the best design modification over Prasad et al. [130].

Prasad et al. [131] proposed some more energy conservation scenarios for sponge iron plants. They proposed preheating of feed materials to kiln from 30^{o}C to 120^{o}C and air from 30^{o}C to 300^{o}C using waste gas coming from ABC. Along with this they also recovered heat energy from kiln outlet by cooling it from 1020^{o}C to 110^{o}C using waste gas coming from ESP. Authors proposed design modifications resulted in 12.3% and 93.7% less coal and water consumption in comparison to conventional process. Though reduction in water consumption by authors proposed design is comparable with that of Strategy-1, reduction in coal consumption is much less than that of Strategy-1. The % reduction of waste gas generation in comparison to existing system is 12.5% through author's modification whereas Strategy-1 reduces 51.7%. The payback period is less for authors proposed design than that of Strategy-1 though the savings in Strategy-1 is very high, which is 19.52 Lakh/year/t of SI whereas authors proposed design offers only 8.24 Lakh/year/t of SI. Based on coal consumption, water requirement, waste gas generation and possible savings it offer, total site integration through Strategy-1 can be considered as the best design modification over Prasad et al. [131].

Advantages of Strategy-1 over the published literature are also shown graphically in Fig. 6.48. Fig. 6.48(a) shows that Strategy-1 offers large % savings in coal and water consumption and reduction in waste gas generation. Reasons for such variation can be better understood through following points:

1. To recover maximum possible waste heat the present work considers all heat recovering and utilizing areas mentioned in Section 4.2. Though Prasad et al. [131] considered all areas, they were able to recover only 31.9% of waste heat that was available in the plant as shown in Fig. 6.48(a)

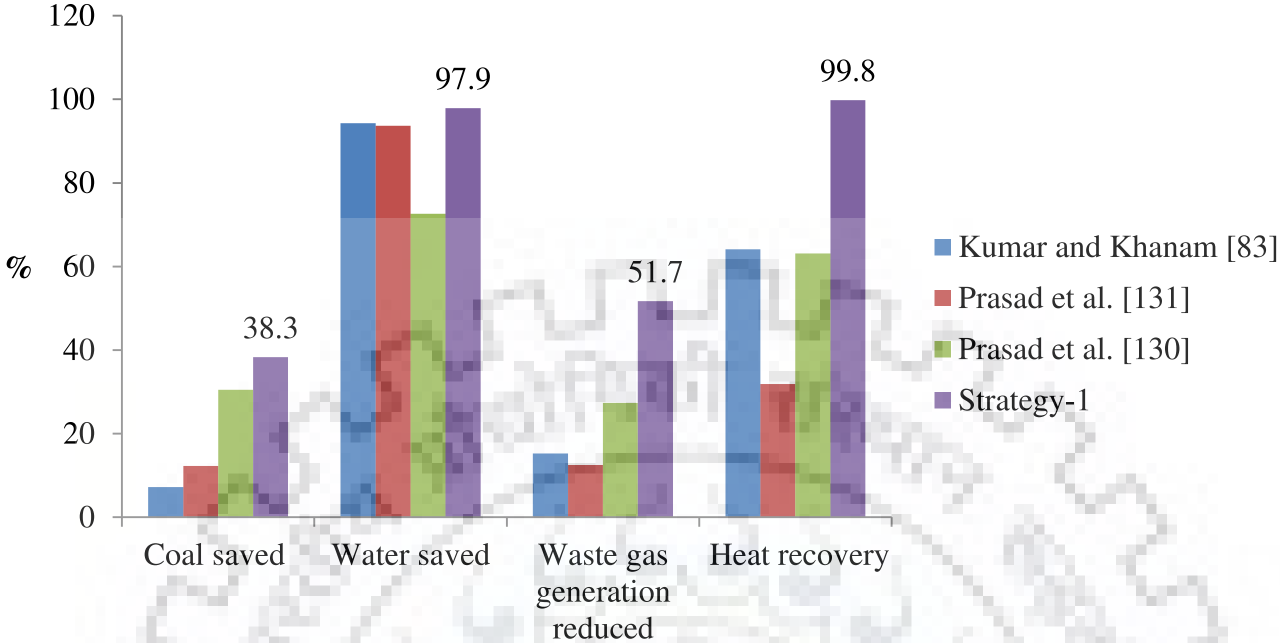
and Table 6.27. However, Kumar and Khanam [83] and Prasad et al. [130] are able to recover maximum 64.1% and 63.1%, respectively, of waste heat only.

2. For Strategy-1, waste heat available from ABC exit of one plant is completely used for preheating kiln inlet streams of all plants; whereas, that of other two plants is used for power generation simultaneously. This ensures maximum heat recovery in the site, which is not possible in the case of single site integration considered in published literature.

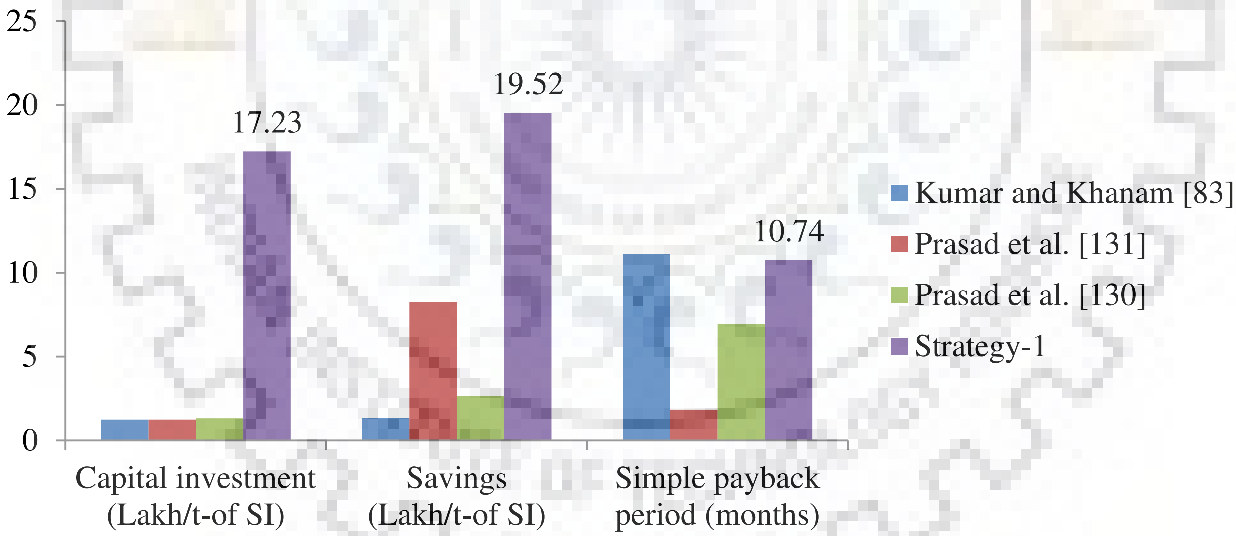
3. The model developed in published literature, for coal utility factor and coal consumption, did not consider all factors such as preheating of dolomite, heat released in its decomposition and also heat required to vaporize coal volatiles as explained in Section 4.5. The modified model, used in the present work, further increased coal savings by 4.3%. Thus, the modified model ensures better coal utility factor, which in turn results in better coal saving.

4. Previous authors considered constant properties for waste gas and kiln air throughout their design computations. But, in all the design modifications through heat integration schemes such as Scheme-1 and Scheme-2 and also through total site integration strategies like Strategy-1 and Strategy-2, in current work, constant properties at average temperature in each iteration are considered. To know the effect of this consideration, on % reduction of coal consumption, separate computations are carried out for both Scheme-1 and Scheme-2 and the results are shown graphically in Fig. 6.49. It shows that reduction of coal consumption is 4.53% and 3.23% more for Process-1 in Scheme-1 and Scheme-2 respectively. For Process-2 also it is 4.32% and 3.83% more, as shown in Fig. 6.49, in respective schemes. Similarly, for Process-3 also it is 5.23% and 3.91% more, as shown in Fig. 6.49, in respective schemes. Consideration of this aspects also, in the current work, results in better coal savings over published literature.

Thus savings in coal and water and hence reduction in waste gas generation, obtained in Strategy-1, are found greater than that of other heat integration schemes and strategies proposed in the current work as well as various design modifications reported in the literature. Also as it offers huge savings with least payback period, Strategy-1 is selected as the best possible design modification for sponge iron cluster.



(a)



(b)

Fig. 6.48 Comparative analysis of Strategy-1 with published literature

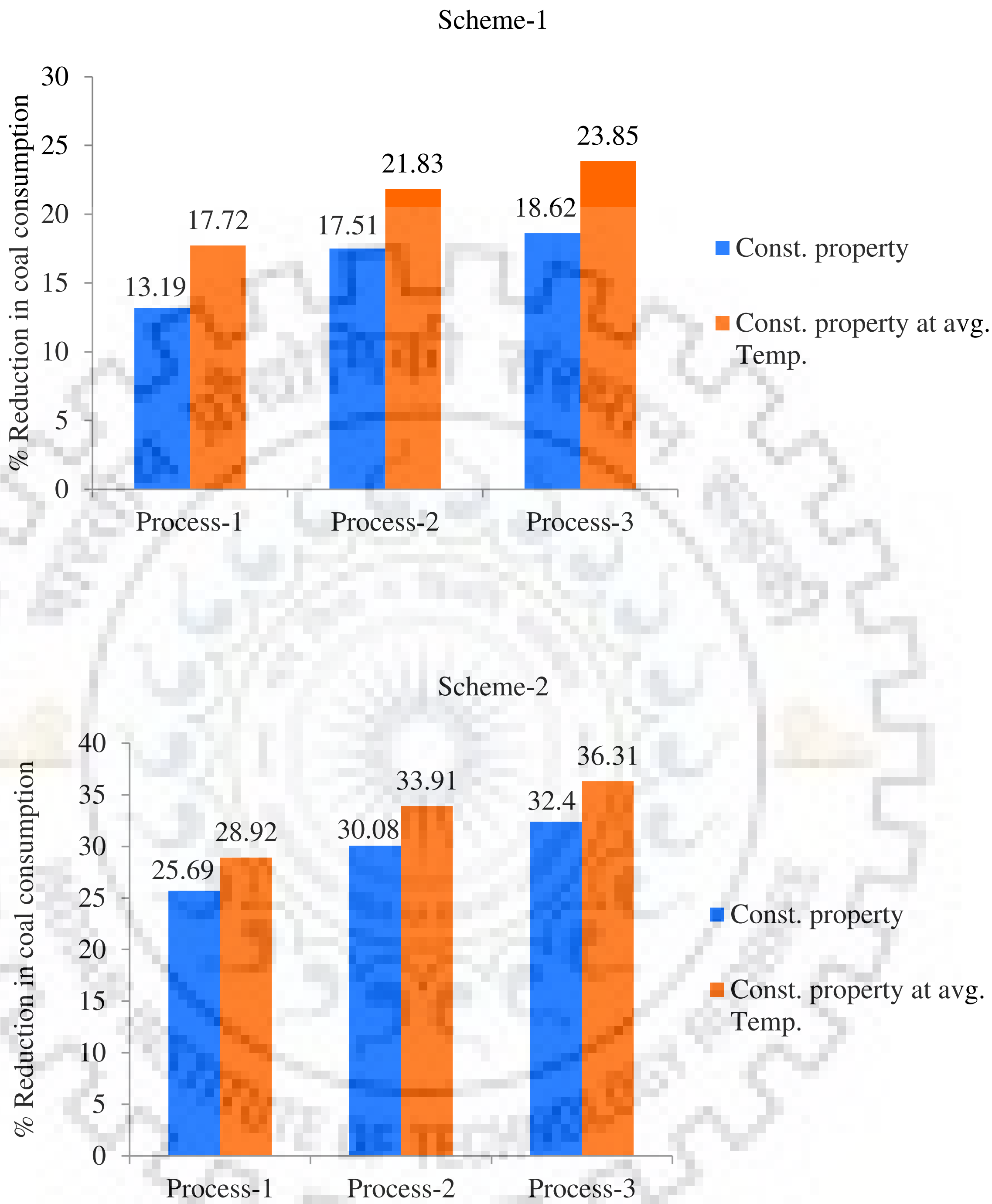


Fig. 6.49 Effect of property data, considered, on % reduction of coal consumption

Table 6.27 Comparative analysis of Strategy-1 with published literature

S. No	Author / Scheme	Capacity of the plant (TPD)	Coal saved (%)	Water saved (%)	Waste gas reduced (%)	(%) Heat recovery	Capital investment (Lakh/ t of SI)	Operating cost	TAC	Savings	Simple payback period (months)
								(Lakh/year/t of SI)			
1	Kumar and Khanam [83]	100	7.2	94.3	15.3	64.1	1.23	10.23	10.35	1.33	11.10
2	Prasad et al. [131]	500	12.3	93.7	12.5	31.9	1.23	8.45	-	8.24	1.83
3	Prasad et al. [130]	446.4	30.5	72.6	27.4	63.1	1.32	6.01	-	2.64	6.95
4	Strategy-1	262.4	38.3	97.9	51.7	99.8	17.21	10.29	12.01	19.52	10.73

6.7 ALL POSSIBLE STRATEGIES FOR ENERGY CONSERVATION IN GIVEN SPONGE IRON CLUSTER

In the present work, two different strategies for total site integration of sponge iron cluster are proposed as Strategy-1 and Strategy-2 based on the modified approach of total site integration developed by Ramanaiah and Khanam [142] to conserve energy in total site of plants of similar type. Although only two strategies are proposed for total site integration, Strategy-1 shows maximum heat recovery. However, Strategy-2 is chosen to perform the comparative analysis. The selection of these strategies, as one may think, is not a hit and miss thing. This is explained hereafter.

Though both Strategy-1 and Strategy-2 are first defined and then whole analysis is carried out, it could be possible that better strategies are missed if the selection of strategies is not done systematically. This can be avoided while identifying strategies using guidelines proposed in Section 4.4, Step-1 and computing % recovery of heat, for each strategy. However, to illustrate that optimal strategy should not be missed while identifying these, following the guidelines all possible strategies are identified based on recovering and utilizing areas, which can be analysed through Table 6.28. In Table 6.28 these areas are identified using data of sponge iron plants. It includes % heat recovery of each strategy, which is computed using stream data of that strategy a priory to detailed design of it. After final design it was found that Strategy-1 recovers 99.8% of waste heat available whereas, Strategy-2 recovers only 86%. Though other strategies are proposed as shown in Table 6.28, it is difficult to recover more than 99.8% of heat through any other strategies. Thus, there is no chance to miss the best possible strategy if % recovery of heat is computed.

For example, for maximum heat recovery one should consider all the possible recovering areas as well as utilizing areas while defining a strategy. Strategy 1, 2 and 6, in Table 6.28, consist of all recovering areas as well as utilizing areas. Further, % heat recoveries for these strategies are also shown in Table 6.28, which is maximum for Strategy-1. Thus, it is selected as optimal strategy. Hence, if one selects the best strategy based on maximum % heat recovery, one cannot miss the optimal strategy. Thus it can be concluded that Strategy-1 that is proposed in the current work is the best design modification for sponge iron cluster. The schematic representation of all possible strategies are given in Appendix D.

Table 6.28 All possible strategies for energy conservation in the given site of sponge iron plants

Strategy No.	Strategy	Heat recovering areas	Heat utilizing areas	Recovered heat (kW)	Recovered heat (%)
1.	Waste gas that exits from ABC of Process-1 is used for preheating the kiln feed and slinger coal of all three plants, individually, from 30°C to 300°C. Kiln outlet stream is cooled from 1020°C to 110°C using kiln air of its respective plant. Further, kiln air is preheated using splitted streams of waste gas to the maximum possible temperature (421°C in this case) such that ΔT_{min} equal to 50°C is maintained. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation.	All three	All three	25602.7	≈100
2.	Waste gas streams exiting the ESPs of Process-1, Process-2 and Process-3 are mixed and used as hot stream to preheat kiln feed and slinger coal of all three processes at Process-1 from 30°C to 150°C. Further, kiln outlet stream is cooled from 1020°C to 110°C using kiln air of its respective process. In addition to it, waste gas exiting from ABC of Process-1 and Process-2 are used for power generation.	All three	All three	22211.5	≈86.8
3	Waste gas streams exiting the ESPs of Process-1, Process-2 and	1	1 and 2	1400.93	≈5.5

	Process-3 are mixed and used as hot stream to preheat kiln feed and slinger coal of all three processes at Process-1 from 30^{O}C to 150^{O}C.				
4	Waste gas streams exiting the ESPs of Process-1, Process-2 and Process-3 are mixed and used as hot stream to preheat kiln feed and slinger coal of all three processes at Process-1 from 30^{O}C to 150^{O}C. Further, kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air of its respective process.	1 and 2	All three	4171.88	≈16.3
5.	Waste gas that exits from ABC of Process-1 is used for preheating the kiln feed and slinger coal of all three processes, individually, from 30^{O}C to 300^{O}C. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation.	1 and 2	1 and 2	14903.57	≈58.2
6.	Waste gas that exits from ABC of Process-1 is used for preheating the kiln feed and slinger coal of all three processes, individually, from 30^{O}C to 300^{O}C. Further, kiln outlet stream is cooled from 1020^{O}C to 110^{O}C using kiln air of its respective process. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation.	All three	All three	17674.52	≈69.0
7.	Waste gas streams exiting the ESPs of Process-1, Process-2 and Process-3 are mixed and used as hot stream to preheat kiln feed	1 and 2	1 and 2	19441.12	≈75.9

	and slinger coal of all three processes at Process-1 from 30^oC to 150^oC. In addition to it, waste gas exiting from ABC of Process-1 and Process-2 are used for power generation.				
8.	Waste gas that exits from ABC of Process-1 is used for preheating the kiln feed and slinger coal of all three plants, individually, from 30^oC to 300^oC and kiln air to the maximum possible temperature such that ΔT_{min} equal to 50^oC is maintained. In addition to it, waste gas streams from ABC exits of Process-2 and Process-3 are combined and used for power generation.	1 and 2	All three	21987	≈85.8

6.8 GENERALIZATION OF THE RESULTS

In the present work the modified approach of total site integration is illustrated through a case study of sponge iron cluster and thus, results found is applicable to these plants. However, a few data may change with capacity such as coal and water consumptions, heat duties of exchangers, waste gas generation, operating and capital costs, savings and payback period, etc. The limitation of the proposed approach is that it can be employed to the industrial cluster where all available plants are of same process. However, these plants may have same process and utility streams or different. Sponge iron, steel, cement, thermal power plants, etc. used coal as utility and so, in these plants process and utility streams are same. Thus, the approach illustrated through sponge iron cluster, can be applicable effectively in these plants also irrespective of variation in operating conditions. On the other hand, in plants such as oil refineries, petrochemical, etc. process and utility streams are different. For such cases, another approach, proposed in the current work, is applicable. As it works on pinch analysis principles, it is also employed effectively in different clusters. However, a few factors like design parameters of new equipment, utility consumption, economic data, etc. may vary according to operating conditions.

CHAPTER 7

CONCLUSIONS AND FUTURE RESEARCH DIRECTIONS

In this Chapter salient conclusions drawn from the present investigation along with various recommendations for future work are summarized.

7.1 CONCLUSIONS

An enormous potential for saving energy is available in coal based sponge iron plants as these are associated with high percentage of energy wastage per unit sponge iron production. To deal with the current issue of energy crisis, two design modifications, Scheme-1 and Scheme-2, are proposed, based on principles of process integration, for a site of three sponge iron plants such as Process-1, Process-2 and Process-3 being operated with different capacities in India. Further, two strategies, Strategy-1 and Strategy-2, are suggested for total site integration based on the modified approach developed in the present work to conserve energy in total site of plants of similar type where conventional methods are not applicable. Based on results, obtained from implementation of these design modifications and comparison of these with the existing system and published literature on the basis of coal consumption, water requirement, capital investment, savings, payback period, waste gas generation, energy ratio (ER) and %heat recovery several noteworthy conclusions can be drawn as:

1. Maximum coal and water savings are found in Scheme-2 over Scheme-1. In Scheme-2 coal savings are 28.9%, 33.9%, 36.3% in Process-1, Process-2 and Process-3, respectively, whereas water savings are found as 96.97%, 97.28%, 97.49%. On the other hand, in Scheme-1 coal savings are only 17.7%, 21.8% and 23.85% and water savings are 95.54%, 93.94% and 95.96% in same processes in comparison to the existing systems.

2. Waste gas generation, in comparison to existing system, in Scheme-2 is much lower than in Scheme-1. Scheme-2 generates 35.8%, 44.3%, 37% less waste gas in Process-1, Process-2 and Process-3, respectively, whereas in Scheme-1 it is 20.2%, 24.5% and 24.4% less in same processes in comparison to the existing systems.

3. Actual energy consumption, in Scheme-2, is found close to theoretical energy required as compared to Scheme-1. Excess energy consumption, which is 73% for Process-1 is reduced to

42% and 23% through Scheme-1 and Scheme-2, respectively, whereas for Process-2 this is decreased to 38% and 17% through respective schemes from 77%. Same trend is found in Process-3 also. Similarly, ER is found minimum in Scheme-2 for all three processes than Scheme-1.

4. In Process-1 and Process-2, which do not have power generation unit, Scheme-2 recovers maximum amount of available waste heat than Scheme-1. On the other hand, for Process-3, which already has power generation unit, Scheme-1 recovers more amount of waste heat as compared to Scheme-2.

5. Scheme-2 offers significant savings along with minimum capital investment and payback period over Scheme-1 in all three processes. Thus, Scheme-2 can be considered as best design modification over Scheme-1.

6. Among two strategies proposed for total site integration, Strategy-1 offers better coal and water savings than Strategy-2. In Strategy-1 coal savings are 38.3% and water savings are 97.9% whereas, in Strategy-2 coal and water savings are 21.72% and 96.34%, respectively, in comparison to the existing site system. Consequently, waste gas generation reduces by 51.72% and 38.4%, through Strategy-1 and Strategy-2, respectively, as compared to the existing system.

7. For all three processes combined together as a total site, theoretical energy required for reduction reactions to continue in kiln is 63,642 kW. However, in existing conditions actual energy consumed for the same purpose is 1,11,710.2 kW which is reduced to 87445.7 kW and 68,927.3 kW through Strategy-1 and Strategy-2, respectively. Thus, ER which is 1.76, prior to total site integration, is reduced to 1.08 and 1.37 through Strategy-1 and Strategy-2 respectively. Hence, actual energy consumption, in Strategy-1, is more close to theoretical energy required as compared to Strategy-2.

8. Strategy-1 recovers almost all available waste heat, i.e. 99.8%, in total site than Strategy-2, which recovers only 86%.

9. Although capital cost required for Strategy-1 is more than that of Strategy-2, it requires less operating cost and TAC than Strategy-1. Both Strategies offer huge savings where Strategy-1 and Strategy-2 give savings of 5122.41 lakh/year and 4883.15 lakh/year, respectively.

10. Savings in coal and hence reduction in waste gas generation offered by Strategy-1 are found significantly greater than savings offered by various design modifications reported, based on single site integration, in literature. Consequently, total site integration through Strategy-1 offers

minimum 2.37 times to maximum 14.68 times more savings than that shown in published literature.

11. Since total site integration through Strategy-1 offers better results over Strategy-2 and also over Scheme-1 and Scheme-2, along with other design modifications reported, based on single site integration, in literature it is considered to be the best possible design for sponge iron cluster of any capacity. Also integrating as a total site is more profitable than single system. Along with this, Strategy-1 reduces waste gas emissions significantly while making sponge iron cluster more environment friendly. Hence, total site integration can be concluded to be an important tool to save energy in any coal based sponge iron cluster.

12. Based on results obtained, in present study, it can be concluded that through economic analysis best energy conservation strategy can be selected. However, that can also be chosen at initial stage while comparing %heat recovery of each strategy.

The four design modifications proposed in the current work through both energy integration schemes (Scheme-1 and Scheme-2) and total site integration strategies (Strategy-1 and Strategy-2) offer better energy savings. Among three processes (Process-1, Process-2 and Process-3) individually and as a total site various possibilities are identified to integrate heat recovering areas (heat sources) with heat utilizing areas (heat sinks). But only total site integration strategies involving all heat sources and sinks are able to recover maximum unutilized heat and also able minimize GHG emissions. Along with this, all the proposed modifications require same level of disturbance during start-up and shut down but energy integration schemes offer minimum process modifications in comparison to total site integration strategies as they require set up of power generation plants. However, once these are set up ease of operation is similar in all the four cases. Savings offered by total site integration strategies are better than that offered by single site integration schemes. Hence Strategy-1 is considered as the best design modification for coal based sponge iron cluster.

7.2 FUTURE RESEARCH DIRECTIONS

For the advancement of knowledge in the area of energy integration of sponge iron industries, following future research directions are identified:

1. Variation in physical properties of different streams can be considered for heat integration of sponge iron cluster.

3. Based on this approach, other commodities such as water, separating agents, etc. can be conserved as it requires areas from where these commodities can be recovered (as being wasted) and where these can be utilized.

4. A mathematical model to optimize TAC in sites of similar plants can be proposed and if options are available, how to select best one, can be understood through the present approach.

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APPENDIX A

MATERIAL AND ENERGY BALANCE

This Appendix deals with the material and energy balances carried out on Process-3, shown in Fig. 3.1. The calculations shown in the Appendix are made in the standard spread sheet.

A.1 COMPONENT MASS BALANCE AROUND ROTARY KILN

The main components of the feed material (iron ore, Feed coal and dolomite) are iron, coal, gangue and ash. The mass balance of the components are shown here. The calculations are done on hourly basis.

A.1.1 Iron

Input

Fe present in iron ore (Fe_T) = {Iron Ore feed rate × % of Fe (Total) in Fe_2O_3}

= {6500×0.63) = 4095 kg (A.1)

Output

Fe present in Sponge iron = {Sponge iron produced × % of Fe_M present in sponge iron}

= {4166×0.926) = 3857.7 kg (A.2)

Fe lost in the fly ash = Production of fly ash × % of Fe_2O_3 in fly ash × Fe/Fe_2O_3 ratio =

= 684.45×0.424×112/160 = 203.14 kg (A.3)

Other losses by difference (Oversize from Cooler outlet, accretion) = 34.16 kg (A.4)

Total output = (A.2) + (A.3) + (A.4) = 4095 kg

A.1.2 Gangue

Input

Gangue = {Iron Ore feed rate × % of gangue in iron ore}

= {6500×0.08) = 520 kg (A.5)

Output

Gangue = {Sponge iron produced × % of gangue present in sponge iron}

$= \{4166 \times 0.0462) = 192.47$ kg (A.6)

Fe lost in the Fly Ash = Production of Fly Ash × % of Fe_2O_3 in Fly Ash × Fe/Fe_2O_3 ratio

= Difference of (A.5) and (A.6) = 327.53 kg (A.7)

Total output = (A.6)+ (A.7) = 520 kg

A.1.3 Carbon

Input

C (total) present in feed coal = {feed coal feed rate × % of C in feed coal}

$= \{4500 \times 0.606) = 2727$ kg (A.8)

C (total) present in slinger coal = {slinger coal feed rate × % of C present in slinger coal}

$= \{2000 \times 0.6221) = 1244.2$ kg (A.9)

C (total) present in dolomite (present as $CaCO_3$) = {dolomite feed rate × % of $CaCO_3$ present in dolomite × C associated with $CaCO_3$}

$= \{120 \times 0.5 \times 12/100) = 7.2$ kg (A.10)

Total input = (A.8) + (A.9) + (A.10) = 3978.4 kg

Output

C in sponge iron (as FC) = Sponge iron flow rate × % of C in sponge iron

$= 4166 \times 0.0017 = 7.08$ kg (A.11)

C in fly Ash (as FC) = Fly ash rate × % of C in fly ash

$= 684.45 \times 0.1348 = 92.26$ kg (A.12)

C in Char (as FC) = {Char flow rate existing from cooler ×% of FC present in Char}

$= \{554 \times 0.285) = 157.89$ kg (A.13)

$$\text{C in Char (as VC)} = \left(\frac{\text{VC in slinger coal}}{\text{VM in slinger coal}}\right)\text{VM in char} = \left(\frac{382.2}{634.2}\right)10.526 = 6.34kg$$

= (382.2/634.2)*10.526 = 6.34 kg (A.14)

C in CO_2 generated from $CaCO_3$ = 7.2 kg (A.15)

C consumed in Reaction-1 = (Fe_2O_3 consumed) x 12/160

= 5559.79 x 12/160 = 416.98 kg (A.16)

C consumed in Reaction-2 = (FeO consumed) x 12/72

= 4515.98 x 12/72 = 752.66 kg (A.17)

Excess FC burnt = (FC in feed coal + FC in slinger coal) - (FC in sponge iron + FC in fly ash + FC in char + C consumed in Reaction-1 + C consumed in Reaction-2)

= (1842.75 + 862) - (7.08 + 92.26 + 157.89 + 416.98 + 752.66)

= 1277.87 kg (A.18)

Feed coal VC = 884.25 kg (A.19)

Slinger coal VC burnt = (Slinger coal VC - Char VC) * 0.67

= 251.82 kg (A.20)

Slinger coal VC unburnt = (Slinger coal VC - Char VC) * 0.33

= 124.03 kg (A.21)

Total output = (A.11) + (A.12) + (A.13) + (A.14) + (A.15) + (A.16) + (A.17) + (A.18) + (A.19) + (A.20) + (A.21)

= 3978.4 kg

A.1.4 Ash

Input

Ash in feed coal = feed coal feed rate × % of ash in feed coal

= 6500 × 0.2374 = 1068.3 kg (A.22)

Ash in slinger coal = slinger coal feed rate × % of ash in slinger coal

= 2000 × 0.2129 = 425.8 kg (A.23)

Total input = (A.22) + (A.23) = 1494.1 kg

Output

Ash in char = char flow rate × % of ash in char = 554 × 0.696 = 385.58 kg (A.24)

Ash in fly ash = 684.45 × 0.3594 = 245.99 kg (A.25)

Free ash to cooler (from kiln outlet) = 862 kg (A.26)

Total ash output = (A.24) + (A.25) + (A.26) = 1494.1 kg

A.2 MASS BALANCE AROUND DIFFERENT EQUIPMENT

The flow sheet for sponge iron production process is shown in Fig. 3.1 which includes rotary kiln, ABC and DSC, EC (Evaporating Chamber), rotary cooler (RC), electrostatic precipitator (ESP) and Chimney. The mass balance around the above equipment on per hour basis are carried out to know the unknown flow rates of the process flow sheet, shown in Fig. 3.1. These balances are based on values obtained in Step 8 of material and energy balance discussed under section 4.1.2.

A.2.1 Rotary Kiln

Input

Iron ore = 6500 kg

Feed coal = 4500 kg

Dolomite = 120 kg

Slinger coal = 2000 kg

Air = kiln air + air from CBA fan + air with slinger coal = 37603.5 kg

Total input = 50723.5 kg

Output

Sponge iron = 4166 kg

Char = 554 kg

Spent lime = 70 kg

Ash = 862 kg

Mass of a mixture of fly ash, waste gas and fines existing from rotary kiln

= 45071.5 kg

Total output = 50723.5 kg

A.2.2 ABC+DSC

Input

Fly ash, waste gas and fines = 45071.5 kg

Air = 3381.4 kg

Total input = 48452.9 kg

Output

Dust = 700 kg

Waste gas = 47752.9 kg

Total output = 48452.9 kg

A.2.3 WHRB

Input

Waste gas = 47752.9 kg

Water = 13953.6 kg

Total input = 61706.5 kg

Output

Waste gas = 47752.9 kg

Steam = 13953.6 kg

Total input = 61706.5 kg

A.2.4 Electro-static precipitator (ESP)

Input

Dusty waste gas = 47752.9 kg

Output

Dust = 1300 kg

Clean waste gas = 46452.9 kg

Total output = 47752.9 kg

A.3 OVERALL MASS BALANCE AROUND COMPLETE PLANT

The overall mass balance on per hour basis around the complete plant considering all input and output from Process-3 is given in the section below:

Input

Iron ore = 6500 kg

Feed coal = 4500 kg

Dolomite = 120 kg

Slinger coal = 2000 kg

Kiln air = 26818 kg

CBA fan air = 6704.5 kg

Air with slinger coal = 4081 kg

Air in ABC = 3381.4 kg

Water to WHRB = 13953.6 kg

Water to rotary cooler = 251380 kg

Total input = 319438.5 kg

Output

Sponge iron = 4166 kg

Char = 554 kg

Spent lime = 70 kg

Ash = 862 kg

Steam from WHRB = 13953.6 kg

Water from rotary cooler = 241324.8 kg

Vapor from rotary cooler = 10055.2 kg

Waste gas from chimney = 46452.9 kg

Dust from wet scrapper = 700 kg

Dust from ESP = 1300 kg

Total output = 319438.5 kg

A.4 ENERGY BALANCE AROUND ROTARY KILN

Energy balance around rotary kiln is required to compute the heat content of different streams inside the kiln. Total heat contents involve heat of reactions and radiation loss from kiln. Along with this, sensible heat lost in sponge iron, char, dust, flue gas and volatile matter are also considered.

A.4.1 Heat of reaction

In the present process reduction as well as combustion reactions, shown in Section 3.1.3, are involved. The heat of reaction involved in all reactions are computed and shown in this section. For energy balance the specific heat of components is considered as variable.

A.4.1.1 Reduction reactions

Heat of reaction at $T^{O}C$ = Enthalpy of product at $T^{O}C$ - Enthalpy of reactants at $T^{O}C$ + Heat of formation of reaction $25^{O}C$ (A.27)

Heat of formation of reaction at $25^{O}C$ = heat of formation of products at $25^{O}C$ - heat of formation of reactants at $25^{O}C$ (A.28)

Reaction-1: $Fe_2O_3 + CO \leftrightarrow 2\ FeO + CO_2$

To compute the enthalpy of products and reactants total number of moles of each component is required which is calculated on per hour basis as mention below:

Fe_2O_3 present in iron ore = {Iron ore feed rate × % of Fe (Total) in Fe_2O_3 × Ratio of Fe_2O_3 /Fe_2}

= {6500×0.63×((2×56+3×16)/(2×56)) = 5850 kg

Fe_2O_3 lost in the fly ash = Production of fly ash × % of Fe_2O_3 in fly ash = 684.45×0.424

=290.21 kg

Fe_2O_3 reacted = 5850-290.21 = 5559.79 kg

Moles of Fe_2O_3 = (Weight of Fe_2O_3/Molecular weight of Fe_2O_3) = 5559.79/160 = 34.75 kmol

CO consumed = 34.75 kmol

FeO produced = 2 × 34.75 = 69.50 kmol

CO_2 produced = 34.75 kmol

Reaction-1 is taking place at 1020^{O}C inside the kiln. The enthalpy of Fe_2O_3 from 25^{O}C to 1020^{O}C is calculated and given below where 25^{O}C is the reference temperature.

Enthalpy of FeO = $mC_p\Delta T$

C_p (Fe_2O_3) = 36.38 kCal/kmol-K

Thus, enthalpy of Fe_2O_3 = (34.75)(36.38)(1293-298) = 1257.87 MCal

Similarly, enthalpies of CO, FeO and CO_2 are computed and found as 261.2 MCal, 941.07 MCal and 415.32 MCal respectively.

The heat of formations for Fe_2O_3, CO, FeO and CO_2 are -198.5, -26.416, -64.62 and -94.052 kCal/mol, respectively [64].

Heat of formation for the reaction at 25^{O}C :

= {2×(-64.62) +(-94.052)}- {(-198.5) + (-26.416)} = 1624 kCal/mol = 56.432 MCal

Heat of reaction for Reaction-1 :

[$(\text{Enthalpy})_{FeO}$ + $(\text{Enthalpy})_{CO2}$ - $(\text{Enthalpy})_{CO}$ - $(\text{Enthalpy})_{Fe2O3}$]+ $(\text{Heat of formation})_1$

= [941.07 + 415.32 - 1257.87 - 261.2]+[56.432] = -106.244 GCal (A.29)

Reaction-2: FeO + CO ↔ Fe + CO_2

FeO reacted = FeO produced in Reaction-1 - FeO went out through sponge iron

= 69.5 - (4166 × 0.1171)/72 = 62.72 kmol

CO consumed = 62.72 kmol

Fe produced = 62.72 kmol

CO_2 produced = 62.72 kmol

The enthalpies of FeO, CO, Fe and CO_2 are calculated in the similar manner as computed for Fe_2O_3 in Reaction-1 and found as 849.32, 471.47, 548.75 and 749.67 MCal respectively.

Heat of formations for FeO, CO, Fe and CO_2 are taken from Green and Perry [64] and based on these values heat of formation of Reaction-2 is computed as -189.169 MCal

Heat of reaction for Reaction-2

= [$(\text{Enthalpy})_{Fe}$ + $(\text{Enthalpy})_{CO2}$ - $(\text{Enthalpy})_{CO}$ - $(\text{Enthalpy})_{FeO}$]+ $(\text{Heat of formation})_2$

= -211.546 MCal (A.30)

A.4.1.2 Combustion reactions

Heat of reaction at T^{O}C = Enthalpy of product at T^{O}C - Enthalpy of reactants at T^{O}C + Heat of combustion of reaction (A.31)

Heat of combustion = Heat of combustion of reactants at 25^{O}C - Heat of combustion of products at 25^{O}C (A.32)

Reaction-3: $C + CO_2 \leftrightarrow 2CO$

CO_2 produced in Reaction-1 and Reaction-2 = 97.47 kmol

C consumed = 97.47 kmol

CO produced = 2 × 97.47 kmol = 194.94 kmol

The enthalpies of C, CO_2 and CO are calculated in the similar manner as computed for Fe_2O_3 in Reaction-1 and found as 431.71, 1164.99 and 1465.34 MCal respectively.

Heat of combustion for C, CO_2 and CO at 25^{O}C are taken from Green and Perry [64] and based on these values heat of combustion of Reaction-3 is computed as 4017.74 MCal.

Heat of reaction for Reaction-3 = {1465.34-431.71-1164.99}+{4017.74}

= 3886.37 MCal (A.33)

Reaction-4: $2CO + O_2 \leftrightarrow 2CO_2$

CO available to react = (CO produced in Reaction-3) - (CO consumed in Reaction-1 and 2)

= 97.47 kmol

O_2 consumed = 48.74 kmol

CO_2 produced = 97.47 kmol

The enthalpies of CO, O_2 and CO_2 are calculated in the similar manner as computed for Fe_2O_3 in Reaction-1 and found as 732.67, 387.36 and 1164.99 MCal respectively.

Heat of combustion for CO, O_2 and CO_2 at 25^{O}C are taken from Green and Perry [64] and based on these values heat of combustion of Reaction-4 is computed as -6592.52 MCal.

Heat of reaction for Reaction-4 = {1164.99-732.67-387.36}+{-6592.52}

= -6547.56 MCal (A.34)

Reaction-5: C + O_2 ↔ CO_2

Fixed carbon burnt to CO_2 is computed as given below: (M5+J12-J27-M25-J34-L41)/12

Fixed carbon burnt to CO_2 = Total fixed carbon - Fixed carbon lost

Total fixed carbon = 6500×0.4095 + 2000×0.431 = 2704.75 kg

Fixed carbon lost = Carbon lost in sponge iron, char, fly ash, and carbon consumed in Reaction-1 and 2 (i.e., Reaction-3) = A.11 + A.12 + A.13 + 97.47

= 106.49 kmol

Thus, Fixed carbon burnt to CO_2 = 106.49 kmol

No. of moles of carbon = 106.49 kmol

O_2 consumed = 106.49 kmol

Heat of reaction for Reaction-5 = -10060.76 MCal (A.35)

Reaction-6a: C + O_2 ↔ CO_2

Carbon of V.M. present in slinger coal burns to CO_2 = (382.2-6.34)×0.42/12

= 13.15 kmol

O_2 consumed = 13.15 kmol

CO_2 produced = 13.15 kmol

Heat of reaction for Reaction-6a = -1242.84 MCal (A.36)

Reaction-6b: 2C + O_2 ↔ 2CO

Carbon of V.M. present in slinger coal burns to CO = (382.2-6.34)×0.25/12

= 7.83 kmol

O_2 consumed = 3.92 kmol

CO produced = 7.83 kmol

Heat of reaction for Reaction-6b = -213.79 MCal (A.37)

Reaction-6c: $2H_2 + O_2 = 2H_2O$

H_2 of V.M. present in slinger coal burns to H_2O

Total H_2 in slinger coal = 2000 × 0.0404 = 80.8 kg

H_2 of V.M. present in char is computed from following relation:

$$\frac{\text{C in VM of slinger coal}}{\text{H}_2\text{ of VM in slinger coal}} = \frac{\text{C in VM of char}}{\text{H}_2\text{ of VM in char}} = \frac{1477.203}{312.292} = \frac{33.206}{\text{H}_2\text{ of VM in char}}$$

H_2 of V.M. present in char = 1.34 kg

Jena et. al. [77] shows that 33% VM of slinger coal burns in ABC and only 67% remains for reaction.

Thus, H_2 remains for Reaction = (80.8-1.34) × 0.67 = 53.24 kg

No. of moles of H_2 available for reaction in the kiln = 26.62 kmol

O_2 consumed = 13.31 kmol

H_2O produced = 26.62 kmol

Heat of reaction for Reaction-6c = -1590.43 MCal (A.38)

A.4.2 Heat of decomposition of dolomite

The heat involved in decomposition of dolomite is computed in a manner similar to that is shown for Reaction-1.

Heat of reaction for decomposition of dolomite = 42.70 MCal (A.39)

A.4.3 Coal devolatization

The heat utilized in coal devolatilization is computed in a manner similar to that is shown in Jena et. al. [77].

Heat utilized in coal devolatilization = 252.04 MCal (A.40)

A.4.4 Heat of evaporation of moisture in coal

Total moisture in coal = 224.7 kg

Heat of evaporation of moisture = (224.7)*(539.05) = 121.12 MCal (A.41)

A.4.5 Heat loss through kiln wall

The length and diameter of rotary kiln are 40 m and 4 m, respectively. Heat flux through kiln surface is 6366 kCal/m^2 [77].

Thus, heat Loss = 3.1417×4×40×6366 = 3199.9 MCal (A.42)

A.4.6 Sensible heat

A part of total energy produced in the kiln goes outside with sponge iron, char, fly ash, flue gas, volatile matter etc. The calculations of these losses are given below:

A.4.6.1 Kiln out

Kiln out exits at 1020^{O}C. Sensible heat loss with zero is considered as reference temperature and given as:

Sensible heat out with kiln outlet = $mC_p\Delta T$

= 5652 × 0.7156 × (1020-0) = 4125.52 MJ = 986.97 MCal (A.43)

Cp = 0.7156 kJ/kg-K, depends on composition of kiln out.

A.4.6.2 Waste gas with fly ash from kiln

Value of Cp for fly ash is taken as 1.067 kJ/kg^{O}C [174].

Sensible heat out with fly ash = 684.45 × 1.067 × (900-0)

= 657.28 MJ = 157.24 MCal

Sensible heat out with waste gas = 44387.05 × 1.14 × (900-0)

= 45541.11 MJ = 10895 MCal

Combined sensible heat out with flu ash and waste gas = 11052.24 MCal (A.44)

A.4.6.3 Kiln inlet streams

Sensible heat with kiln feed = 11120 x 1.1307 x (30-0)

= 377.2 MJ = 90.24 MCal

Sensible heat with slinger coal = 2000 x 1.38 x (30-0)

= 82.8 MJ = 19.81 MCal

Sensible heat with total kiln air = 37603.5 x 1.01 x (30-0)

= 1138.15 MJ = 272 MCal

Total heat with kiln inlet streams = 406.62 MCal (A.45)

Heat balance around the kiln is given by the following expression:

Total heat in (by different streams) + Heat generated through reactions

= Total heat out + Heat consumed in reactions + other losses (A.46)

Total heat in (by different streams) = 406.62 MCal from (A.45)

Heat generated through reactions = (A.29)+(A.30)+(A.34)+(A.35)+(A.36)+(A.37)+(A.38)

= 19973.17 MCal (A.47)

Total heat out = (A.42) + (A.43) + (A.44)

= 15239.13 MCal (A.48)

Heat consumed in reactions = (A.33)+(A.39)+(A.40)+(A.41)

= 4302.23 MCal (A.49)

Thus, from the heat balance around the kiln the losses are computed as 838.43 MCal

Heat balance shows that total 19.541 GCal (A.48 + A.49) of energy is consumed in kiln from which 12.039 GCal is lost in the form of sensible heat (A.48-3199.9) and only 7.502 GCal is utilized in the process which is 38.39%. The major fraction of total energy is lost to the atmosphere through waste gas existing from the rotary kiln.

Similarly material and energy balances are carried out on Process-1 and Process-2 respectively.

APPENDIX B

POWER GENERATION

In Strategy-1 power generation is carried out using waste gases from ABC exits of Process-2 and Process-3. For this purpose, these two streams of waste gas are mixed and the resultant waste gas stream having CP 18.42 kW/^{O}C and temperature 900.5^{O}C is used to generate power. For producing steam first water is preheated from 30^{O}C to 345^{O}C where saturated steam is generated at 155.45 bar and then it is superheated up to 700^{O}C and 300 bar. Rankine cycle principles [155] are used for power generation.

Total heat available in GCC (Fig. 6.37) for power generation = 14379.79 kW

Maximum saturation temperature of steam to be produced = 345^{O}C

Boiler feed water temperature = 30^{O}C

Specific heat of water = 4.3 kJ/kg^{O}C

Enthalpy of saturated water at 345^{O}C and 155.45 bar = 1632.52 kJ/kg

Enthalpy of superheated steam = H_1 = 3745.67 kJ/kg

Steam production = Total heat available in GCC/((Enthalpy of superheated steam- Enthalpy of saturated water at 345^{O}C and 155.45 bar)+ Preheating of water)

= 4.147 kg/s

The entropy of superheated steam at 700^{O}C and 300 bar, S_2 = 6.5606 kJ/kg/K

Entropy Saturated liquid S_L (at 80^{O}C) = 1.0753 kJ/kg/K

Entropy Saturated vapor S_V = 7.6132 kJ/kg/K

Saturated liquid enthalpy H_L = 334.92 kJ/kg

Saturated vapor enthalpy H_V = 2643.8 kJ/kg

Wetness fraction (X) = $(S_2-S_V)/(S_L-S_V)$

= (6.5606-7.6132)/(1.0753-7.6132)

= 0.161

Isentropic efficiency = 85%

Turbine outlet enthalpy = $H_2 = XH_L + (1-X)H_V$

$= 2272.07$ kJ/kg

For single stage expansion with isentropic efficiency of 85% [$\eta = 0.85$]

$H_2' = H_1 - \eta(H_1 - H_2)$

$= 2493.11$ kJ/kg

Actual wetness fraction $(X') = (H_2' - H_V)/(H_L - H_V)$

$= 0.065$

Power generated $= 4.147 \times (3745.67 - 2493.11)/1000$

$= 5.19$ MW

APPENDIX C

ECONOMIC ANALYSIS

Development of integrated total site through Strategy-1, requires twelve heat exchangers, a steam turbine to generate power, sixteen ducts to carry the waste gas and kiln air and sixteen FD fans, one for each duct, to sustain the pressure drop. First detailed procedure and formulae used in the study are given then the economic analysis for Strategy-1 is given as a reference. All cost coefficients used in current study is tabulated in Table 6.3A of Chapter 6.

Formulae and procedure used in calculations

1. Exit temperature of gas flowing through duct

Exit temperature of the gas flowing through the duct is found using the following equation [130]:

$$T_O = T_i * \exp\left[\frac{-UA}{mC_p}\right] + T_\infty * \left\{1 - \exp\left[\frac{-UA}{mC_p}\right]\right\}$$

Where,

T_O = Exit temperature of the gas

T_i = Inlet temperature of the gas

T_∞ = Ambient temperature

$$\text{Total thermal resistance across the duct} = \frac{1}{UA} = R_i + R_{metal\ duct} + R_{insulation} + R_o$$

Where,

$$R_i = \frac{1}{h_i A_i} \text{ and } R_O = \frac{1}{h_O A_O}$$

Here, inside heat transfer coefficient, h_i is computed using Dittus-Boltzer equation. Outside heat transfer coefficient, h_o is computed considering combined convection and radiation heat losses to surroundings using trial and error method.

2. Pressure drop through ducts

Pressure drops for ducts are obtained using following equation [175]:

$$\Delta P = \frac{0.109136\, q^{1.9}}{de^{5.02}}$$

Where,

q = gas volume flow (cfm - cubic feet per minute)

d_e= Equivalent duct diameter (inches)

ΔP = Friction (head or pressure loss) (inches water gauge/100 ft of duct)

3. G-G heat exchanger

Fixed cost of G-G shell and tube heat exchanger is computed from following equation [150]:

Cost (Rs) = 1368000 + 34200 $(A)^{0.81}$

The cost coefficients are modified based on the year of study (2017) and the modified equation is,

Cost (Rs) = 1935946 + 48400 $(A)^{0.81}$

Heat exchanger area, A is computed from the equation,

$Q = UA(\Delta T)_{LMTD}$

Here Q is heat exchanger heat load

U is taken as 50W/m^2-°C for gas-gas heat transfer in shell and tube heat exchanger [154]

4. G-S heat exchanger

Standard cost components of G-S heat exchanger (load = 4136 kW) are given below [133]:

(i) Cost of rotary drier with accessories such as girth gears, support rollers, drier mounted instruments, weigh feeders, conveyors and vibrofeeders = Rs. 23 Lakh

(ii) Cost of three phase induction motor for exchanger (Capacity: 55kW) = Rs. 22 Lakh

(iii) Cost of gear box for exchanger = Rs. 31 Lakh

(iv) Approximate installation cost = Rs. 7.5 Lakh

Total cost of G-S heat exchanger is sum of all four components [(i) to (iv)] where component (i) is multiplied with factor size ratio (= Load of G-S HX/load of standard G-S HX).

5. FD fan

Standard cost components of FD fan (Capacity = 122000 m^3/hr) are given below [133]:

(i) Cost of FD fan = Rs. 65 Lakh

(ii) Cost of three phase induction motor for FD fan (Capacity: 80kW) = Rs. 41 Lakh

(iii) Cost of bearing of FD fan = Rs. 11.3 Lakh

(iv) Cost of coupling used to couple fan and motor = Rs. 1.6 Lakh

(v) Approximate installation cost = Rs. 15500/-

(vi) Approximate cost of electrical works including material cost = Rs. 26 Lakh

(vii) Approximate volume of work and cost = Rs. 21 Lakh

Total cost of one FD fan is sum of all seven components [(i) to (vii)] where component (i), (vi) and (vii) are multiplied with factor size ratio (=Volumetric load of FD fan/Volumetric load of standard FD fan).

6. Duct cost

Weight of duct = Duct volume (= Duct surface area x duct thickness) × density

Cost of mild steel = Rs. 33020/t

Standard cost for manufacturing, welding and installation of duct = Rs. 154415/- [133]

Total cost of bare duct

= Duct material cost + manufacturing cost + welding cost + installation cost

Cost of insulation material (glass wool) = Rs. 70000/t

Weight of insulation = Surface area x insulation thickness ×density

Commissioning cost = Rs. 5 Lakh

Total cost of insulation = material cost + commissioning cost

Total duct cost = Total cost of bare duct + Total cost of insulation

C.1 Calculation for Heat exchangers

Amongst twelve heat exchangers nine are gas-solid heat exchangers (HX-1, HX-3, HX-4, HX-5, HX-7, HX-8, HX-9, HX-11 and HX-12) and three are gas-gas shell and tube heat exchangers (HX-2, HX-6 and HX-10) as shown in Fig. 6.36.

C.1.1 Calculation for G-G Heat exchangers

Here calculation for G-G heat exchanger (HX-2) is given.

Parameter	Value
Heat load of G-G heat exchanger (kW)	758.9
U (W/m^2-^{o}C)	50
A (m^2)	31.76
Cost (Lakh)	19.31

Similarly, capital cost required for HX-6 and HX-10 are computed and shown in Table 6.22.

C.1.2 Calculation for G-S Heat exchangers

Here calculation for G-S heat exchanger, HX-9 (Rotary drier) is given.

Parameter	Value
Cost of rotary drier with accessories (Lakh)	56.85
Cost of three phase induction motor to drive the exchanger (Lakh)	22.0
Cost of gear box (Lakh)	31.0
Installation cost (Lakh)	7.5
Total cost of rotary drier (HX-9) with all accessories (Lakh)	117.35

Similarly, capital cost required for other G-S heat exchangers are computed and shown in Table 6.22.

Thus, capital cost of all twelve heat exchangers = 903.02 Lakh (C.1)

C.2 Calculation for Ducts

Here calculation for duct (D1) is given.

Parameter	Value
Steel duct, D1, length (m)	52
Thickness (mm)	6
ID (cm)	30
Density of steel (kg/m^3)	7,876
Weight of the duct (kg)	2,408.6
Cost of steel (Rs/t)	33020
Cost of duct (Rs)	79482
Cost of insulation material (Rs/m)	234
Commissioning charges (Lakh)	5.0
Total cost of insulation including installation (Lakh)	5.12
Thus, total cost of duct, D1,with installation (Lakh)	7.46

Thus, total capital cost of all sixteen ducts with insulation = 119.44 Lakh (C.2)

C.3 Calculation for FD fans

Parameter	Value
Cost of FD fan for duct, D1 (capacity: 8168 m^3/h) (Lakh)	4.35
Cost of three phase motor (capacity: 80 kW) (Lakh)	41.0
Cost of bearing (Lakh)	11.3
Cost of coupling (Lakh)	2,462
Approximate installation cost (Rs)	15500
Approximate cost of electrical works including material cost ($)	2,677
Approximate Volume of Work and Cost (Lakh)	1.41
Total cost for installation FD fan for duct, D1 (Lakh)	61.55

Similar to the calculation of FD fan cost for duct, D1, costs of all sixteen FD fans are calculated.

Total capital cost of all sixteen FD fans = 1068.09 Lakh (C.3)

C.4 Calculation for power generation

Parameter	Value	Remark
Cost of boiler and steam turbine (Rs/kW)	46800	From Gaurav and Khanam [59]
Power generated (MW)	5.19	Computed based on Rankine cycle principles [155].

Thus, capital cost of boiler and steam turbine = 2430.87 Lakh (C.4)

Total capital required = C.1 + C.2 + C.3 + C.4 = 4521.42 Lakh

C.5 Operating cost

Parameter	Value	Remark
The availability of the plant	90%	-
Cost of coal (Rs/t)	2500	-
Flow rate of coal (t/h)	10.9	Computed using Equation 4.9
Total operating cost for coal (Lakh/year)	2148.93	
Cost of water (Rs/t)	60	-
Flow rate of water (t/h)	14.93	-
Total operating cost of water (Lakh/year)	70.62	-
Cost of electricity for all sixteen FD fans (Lakh/year)	397.35	Computed using the method given in the work of Prasad [133].
Maintenance cost of turbine (Lakh/year)	81.9	0.2 Rs/kWh
Total operating cost (Lakh/year)	2698.8	Computed from all individual operating costs

C.6 Savings and Simple payback period

Savings of coal = 6.77 t/h

Savings of water = 691.01 t/h

Savings from coal and water = 4602.59 Lakh/year

Cost of electricity for all sixteen fans = 397.35 Lakh/year

Maintenance cost of turbine = 81.9 Lakh/year

Savings from power = 999.07 Lakh/year (3.5 Rs/kWh)

Savings = 4602.59 + 999.07 - 397.35 - 81.9 = 5122.41 Lakh/year

Simple payback period = total capital investment/yearly savings = 4521.4 /5122.41 = 10.74 months

C.7 Discounted payback period (DPP)

Discounted Payback Period (DPP) in years = $A + \frac{B}{C}$

Where,
A = Last period with a negative discounted cumulative cash flow;
B = Absolute value of discounted cumulative cash flow at the end of the period A;
C = Discounted cash flow during the period after A.

For strategy-1 A, B and C are computed as,

A = 0

B = 4521.42

C = 4656.7 (for savings of 5122.41)

DPP = 0.971 years = 11.65 months

C.8 Total annual cost (TAC) considering straight line depreciation method

If 10 years is the life for each equipment and straight line depreciation is used then,

Annual capital cost = 452.14 Lakh/year

So, TAC = annual capital cost + annual operating cost

= 452.14 + 2698.8= 3150.94 Lakh/year

C.9 Total annual cost (TAC) considering capital recovery factor (CRF)

The total annualized cost is the annualized value of the total net present cost. It is calculated using the following equation:

$TAC_{=} CRF*C_{NPC,tot}$

Accounting 10% rate of return and 10 years of equipment life, CRF = Capital recovery factorCRF = $\frac{i(1+i)^{10}}{(1+i)^{10}-1}$

$C_{NPC,tot}$ = Total net present cost which accounts capital cost and yearly operating cost

For strategy-1, $C_{NPC,tot}$ = = 21104.4 lakh

CRF = 0.163

Hence, TAC = 3434.6 lakh

Similarly net profit value (NPV) and internal rate of return (IRR) are computed and found as,

NPV = 26953.6 lakh

IRR = 94.0 %

C.10 Economic analysis considering cost index factors

The method for computing the cost of some equipment is taken from published articles on sponge iron plants so that the advantage of the proposed schemes can be compared with their findings. It is considered only for comparison purpose. As these costs are older, cost index factors are used to predict recent cost of these components. Cost index factors [173] in 1995, 2009 and 2017 are 381.1, 521.9 and 567.5, respectively. Considering these cost index factors economic analysis, for all design modifications considered in the current study, is carried out and summarized below.

Table: Economic analysis with cost index factors

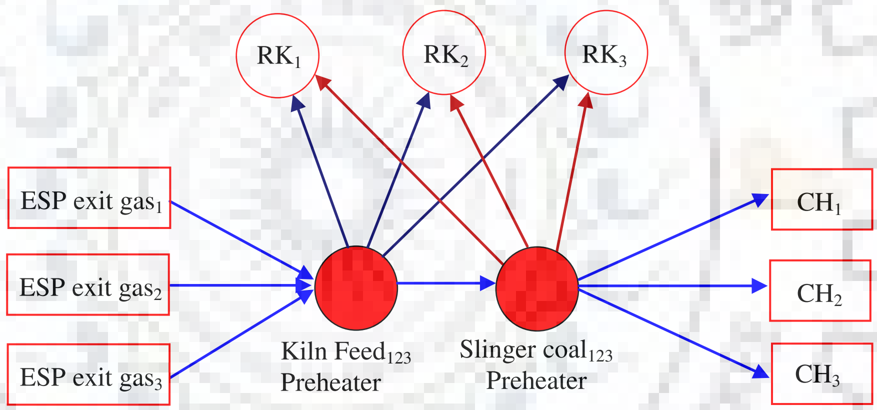
Design modification	Capital cost (Lakh)	Savings (Lakh)	TAC (Lakh)	NPV (Lakh)	IRR (%)	DPP (months)
Process-1: Scheme-1	797.0	1084.68	1171.38	5867.9	114.6	9.7
Process-1: Scheme-2	720.2	1243.26	1000.29	6919.1	147.8	7.65
Process-2: Scheme-1	825.6	1060.43	1214.05	5690.3	107.6	10.28
Process-2: Scheme-2	721.4	1256.62	1000.91	7000.0	149.3	7.58
Process-3: Scheme-1	803.5	1253.52	1280.93	6898.9	132.7	8.46
Process-3: Scheme-2	764.0	1296.04	1070.79	7199.6	145.1	7.78
Strategy-1	4726.4	5122.4	3468.00	26748.6	89.0	12.18
Strategy-2	3989.2	4883.2	3858.69	26015.6	102.0	10.78
All three schemes (Scheme-2)	2205.6	3795.9	3071.89	21118.7	147.0	7.67

APPENDIX D

SCHEMATIC PRESENTATION OF ALL POSSIBLE STRATEGIES FOR ENERGY CONSERVATION IN GIVEN SPONGE IRON CLUSTER

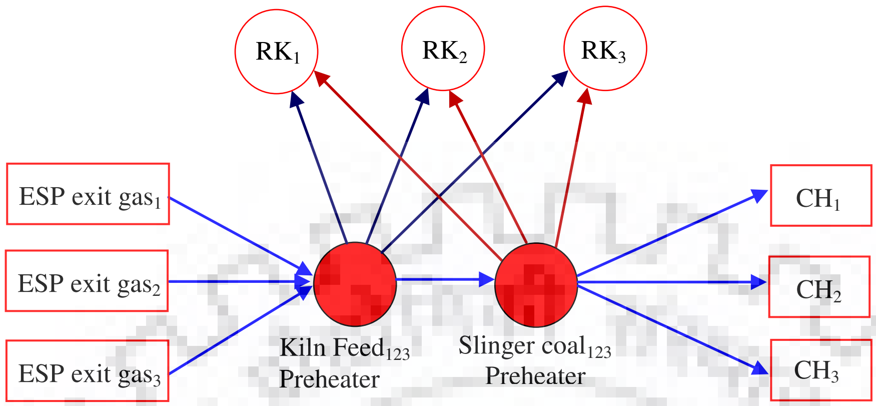
For Strategy-1 and Strategy-2 the schematics are presented in Fig. 6.32 and Fig. 6.39 respectively. The schematic presentation for remaining strategies for energy conservation mentioned in Table 6.28 and also for Scheme-1 and Scheme-2 are given hereunder.

Strategy-3

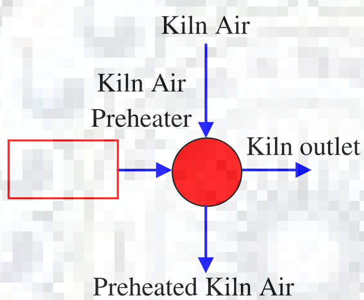


Preheating kiln feed and slinger coal

Strategy-4

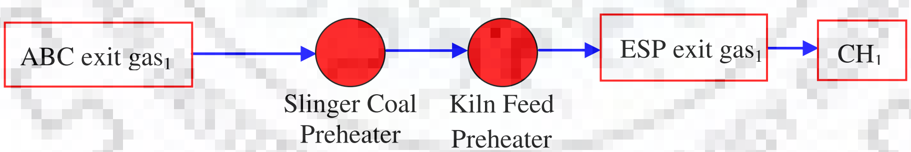


(a) Preheating kiln feed and slinger coal

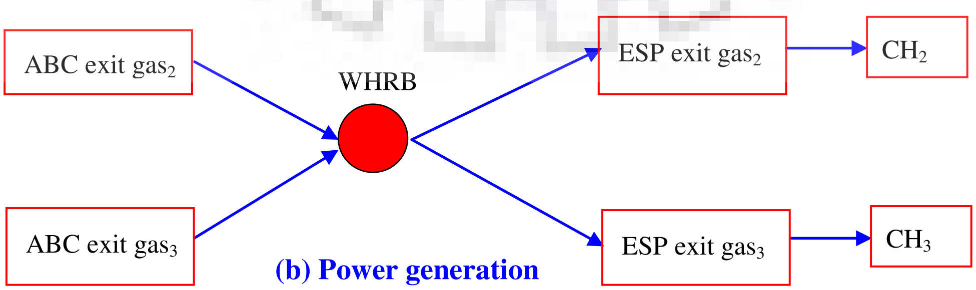


(b) Preheating Kiln air

Strategy-5

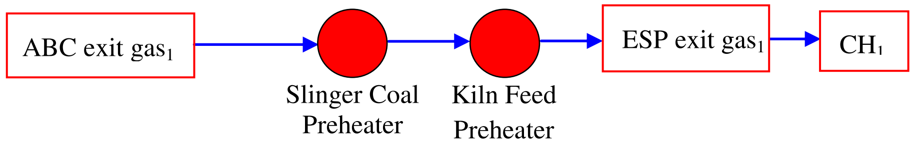


(a) Preheating kiln feed and slinger coal

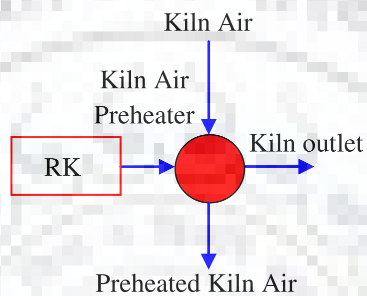


(b) Power generation

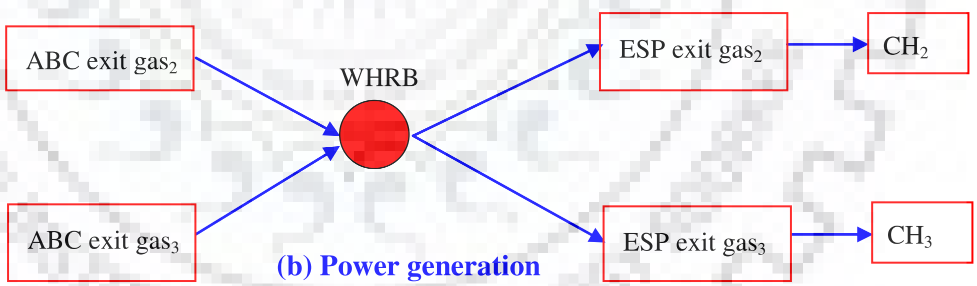
Strategy-6



(a) Preheating kiln feed and slinger coal

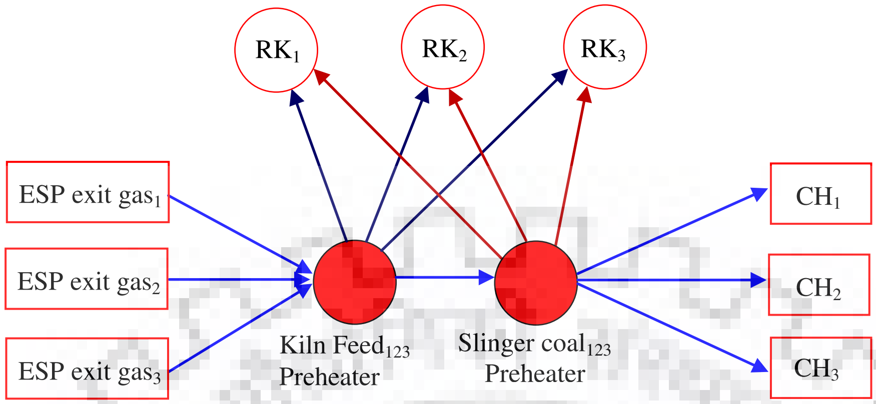


(b) Preheating Kiln air

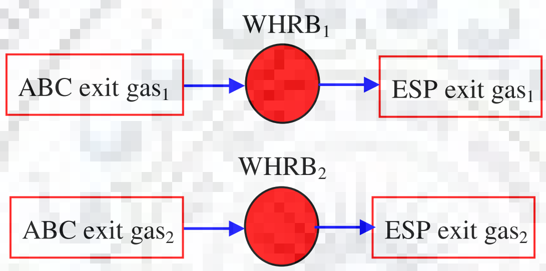


(b) Power generation

Strategy-7

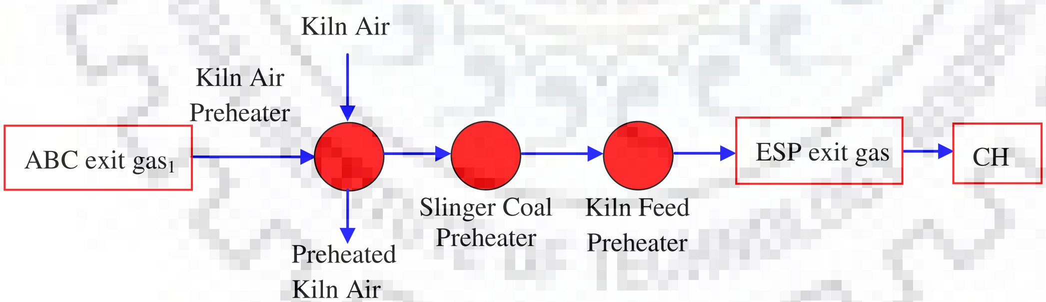


(a) Preheating kiln feed and slinger coal

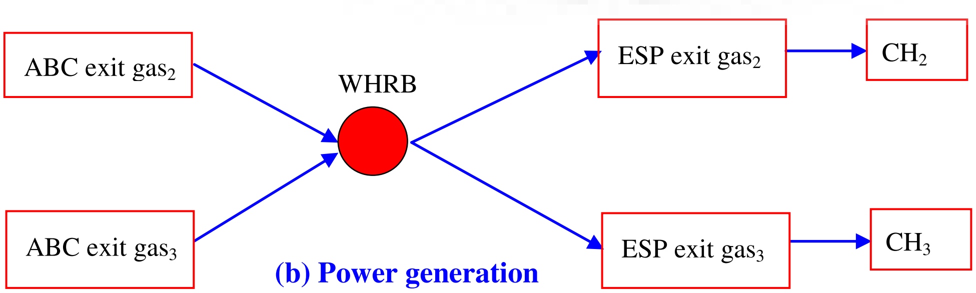


(b) Power generation

Strategy-8



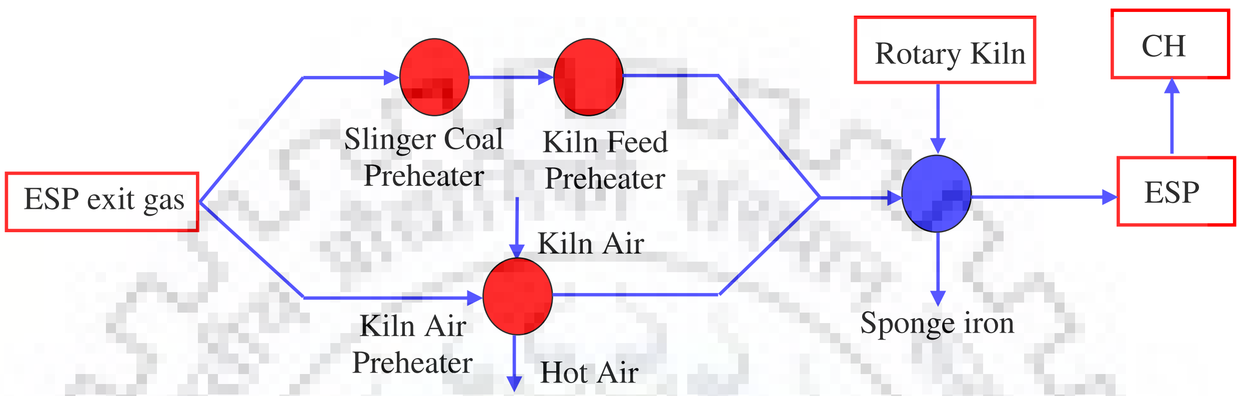
(a) Preheating kiln feed, slinger coal and kiln air



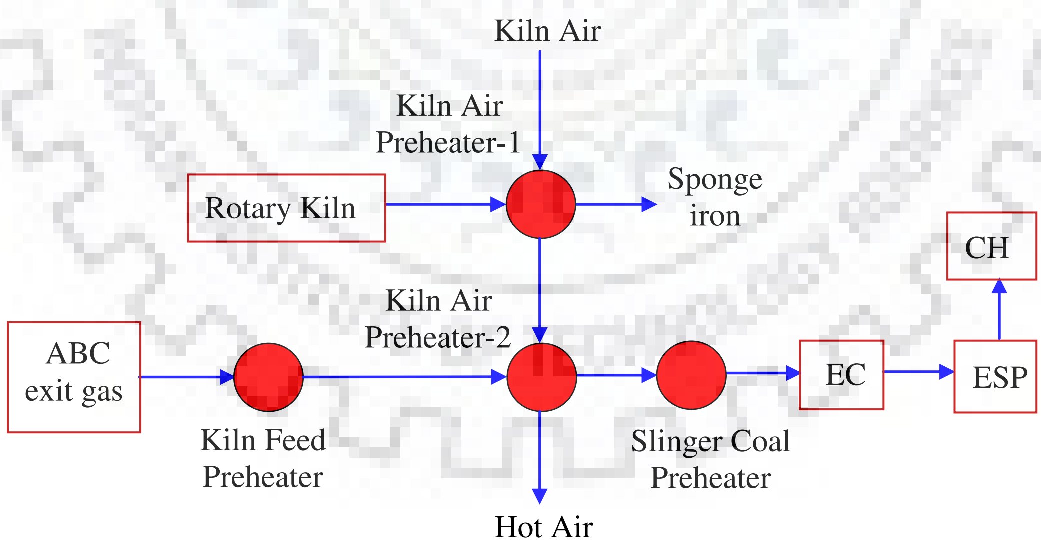
(b) Power generation

Schematic presentation of Scheme-1 and Schme-2 for energy conservation

Scheme-1



Scheme-2



APPENDIX E

EQUIPMENT REQUIRED FOR MODIFICATION

Energy conservation schemes and strategies developed in the present work require the use of new units such as solid-gas and gas-gas heat exchanger and gas carrying duct. The literature related to these exchangers is discussed in following sub-sections.

Solid-Gas heat exchanger

Many authors studied different types of heat exchangers over the period of time. For heat exchange between solid particles and fluid, Lovell and Karnofsky [99] presented a method that accounted conductive resistance heat transfer within solids. Conductive heat transfer in packed beds, subjected to flowing gases were studied analytically by Balakrishnan and Pei [9] based on more realistic assumption of a finite contact spot between spheres in the bed. The effect of parameters such as contact spot dimensions, packing geometry, Biot modulus, and radiation on the conduction mode was examined in their analysis, which incorporated convective effects of flowing gases as boundary conditions. Further, Pelagagge et al. [122] studied optimum criterion of heat recovery by air from solid beds.

Convective heat transfer properties of a porous metallic fibre material used in gas surface combustion burners were studied by Golombok et al. [62]. They predicted heat transfer coefficient between hot gas and metal fibre using a non-steady-state method based on cyclic counter flow heat regenerator theory and found rapid increase in heat transfer coefficient with increasing gas flow rate.

Jain et al. [76] studied air-solid heat transfer in a cyclone heat exchanger having 100 mm inside diameter. Effects of solid feed rate (0.5-7.5 g/s), cyclone inlet air velocity (9-22 m/s) and average particle sizes (163-460 pm) on the heat transfer rate, exit solid temperature and heat transfer coefficient were studied on direct contact heating of sand particles heating with air. They proposed an empirical correlation for the prediction of heal transfer coefficients experimentally which fitted well within an error of +10% to -15% with experimental data. Further, similar study for gas-solid heat transfer was performed by Shimizu et al. [152].

Simulation studies on gas-solid heat transfer during vertical pneumatic conveying, with hot gas supply before the solid feeder were carried out by Rajan et al. [140]. They used one-dimensional, two-fluid model and results showed higher heat recovery with increasing number of gas inlets and less gas to solid flow ratio. Since gas-solid heat transfer phenomenon was similar to heat transfer in a co-current heat exchanger, they identified that driving force of gas-solid heat transfer decreases with height. To overcome this, simulation of parallel pneumatic drying and short ducts, to maintain driving force, were used that resulted in 88.8% increase in heat transfer rate. Rajan et al. [140] also found that shorter parallel ducts could be utilized for drying of fine particles and to improve heat transfer and maintaining uniform driving force, distribution of gas flow could be located at more than one location in the duct. Further, Rajan et al. [139] studied the effect of solid loading ratio, particle size and their interactions on heat transfer rate, temperature profile and thermal effectiveness of gas. They found increase in heat transfer rate with increasing solid loading ratio and decreasing particle size. Higher heat recovery can be achieved for large particles at high solid loading ratios, while it can be achieved with wide range of solid loading ratios for small particles. Authors reported 66.6% improvement in heat transfer.

For the conventional coal based sponge iron industries Prasad et al. [131] designed a rotary drier to preheat feed materials for removing undesirable moisture with waste gas. The design parameters were evaluated while applying principles of drying kinetics and mass and energy balances. The design was based on the concept of Kern [79] and they also predicted the retention time based on the correlation of Friedman and Marshall [57,58].

Gas-Gas heat exchanger

Newey and Howard [120] developed novel gas-to-gas heat exchanger based on shallow fluidized bed technology. The novelty of heat exchanger was discussed in terms of distributor plate. It distributes the fluidized gas in such a way that bed is easily fluidized as well as entire bed flow pattern is parallel to the distributor by providing transverse forces. The heat exchanger is considered to be well suited to waste heat recovery applications where hot flue gases are used to pre-heat combustion air. It is shown that the new system can avoid some of the serious problems of existing gas-to-gas heat exchangers, especially at high temperatures.

Bier et al. [16] predicted overall heat transfer coefficients for gas to gas heat exchange in compact cross-flow micro heat exchangers considering nitrogen, argon and helium gases. They found that for low flow rates overall heat transfer coefficients reduced significantly.

Henk et al. [71] patented a recuperative heat exchanger for gas-gas heat exchange at temperatures above 700^{O}C. The invention particularly relates to a heat exchanger comprising of a refractory lined vessel having a vertically extending steel shell closed at its top and bottom ends by respective ends, wherein the space within the vessel is divided into respective top and bottom end chambers. The end chambers are connected by a plurality of substantially vertical tubes of refractory ceramic material extending between said plates.

Inagaki et al. [75] investigated experimentally flow-induced vibration, heat transfer and pressure drop of helically coiled tubes of an intermediate heat exchanger (IHX), using a full-size partial model and air as the fluid. The test model comprised of 54 helically coiled tubes separated into three layer bundles, surrounding the centre pipe. The vibration of tube bundles was mainly at the centre pipe, and individual vibrations of tube bundles were not significant during operations of IHX. Outside tube forced convection heat transfer was obtained as a function of $Re^{0.51}$ $Pr^{0.3}$ and friction factor, depending on the tube-arrangement, as a function of $Re^{-0.14}$. The gas-gas heat transfer is also possible in shell and tube heat exchanger though heat transfer coefficient is considerably less. Design of such heat exchangers is shown in the work of Sinnott and Towler [154].

Gas carrying duct

It is one of the main equipment. Prasad et al. [132] designed a waste gas carrying duct using first order differential equation based model by estimating heat transfer rates and temperature of the waste gas while flowing through the duct. This method is easily applicable to get the waste gas of desired temperature at different length. It also helped to find out the type of refractory to be used in gas carrying duct as overall heat transfer coefficient was related to thermal conductivity of refractories of respective refractory type [6].

APPENDIX F

LIST OF PUBLICATIONS

<u>*Published in International Journals*</u>

1. Ramanaiah V, Khanam S. Modified approach of total site integration for energy conservation: A case study of sponge iron cluster. Chemical Engineering Research and Design 2018;133:142–54. doi:10.1016/j.cherd.2018.03.014.
2. Ramanaiah V, Khanam S. Analyses of different modifications proposed for sponge iron process for best utilization of waste heat. Process Integration and Optimization for Sustainability 2018:2:365-381. https://doi.org/10.1007/s41660-018-0057-y

<u>*International Conferences*</u>

1. Ramanaiah B.V, Khanam Shabina, "Energy Integration in an Indian Sponge Iron Process", "ICAER-2017" held during December, 12-14, 2017 in IIT Bombay, India.