voestalpine Texas LLC

800 N. Shoreline Blvd, Suite 1600 S
Corpus Christi, TX 78401, USA
Office (361) 452-0746, Fax (361) 452-0840
office.texas@voestalpine.com
www.voestalpine.com/texas

November 7th, 2013

Ms. Melanie Magee
United States Environmental Protection Agency
Region 6
Multimedia Planning and Permitting Division
1445 Ross Ave
Suite 1200
Dallas, TX 75202

Re: Response to Application Completeness Determination for voestalpine Texas LLC Greenhouse Gas Prevention of Significant Deterioration Permit Application

Dear Ms. Magee,

Thank you for reviewing the referenced application and issuing the Completeness Determination dated Oct 22, 2013, from Wren Stenger, Director, Multimedia Planning and Permitting Division. We have prepared responses to your questions and they are included in the enclosure included with this cover letter. Included with the responses in the enclosure are also multiple attachments with requested and/or support data for our responses. Also referenced in Ms. Stenger's cover letter for the Completeness Determination is the necessity for the biological and cultural assessments. Those documents have been provided previously (when the application was originally submitted) and voestalpine has met with the US Fish and Wildlife Service and the National Oceanic and Atmospheric Administration regarding our assessments. voestalpine will be providing revisions in the near future encompassing any comments.

We appreciate the opportunity to provide these responses and if you or your staff has any questions, please contact me at +1-361-452-0746 or matthias.pastl@voestalpine.com or Mr. Johann Prammer at +43/50304/55-77171 or at johann.prammer@voestalpine.com, or Mr. Graham Donaldson at +1-919-855-2274 or graham.donaldson@erm.com.

Kind Regards

Matthias Pastl
Head of Communications and Public Relations

Enclosure

cc: Johann Prammer, voestalpine Stahl, GmbH
Graham Donaldson, ERM

ENCLOSURE

voestalpine Texas LLC Responses to EPA Completeness Comments
Application for Greenhouse Gas Prevention of Significant Deterioration Permit
Direct Reduction Iron/Hot Briquetted Iron Project
November 7, 2013

1. The process description does not appear to follow the process flow diagram that is provided, or identify all emission points with the associated emissions point number (EPN) that emit GHG emissions or have the potential to emit. On the process flow diagram there are several streams that have been identified as "CP". Please provide supplemental information that defines the meaning of this term. Please update the process flow diagram to include a representation of the equipment (compressor and mist eliminator) that is mentioned in the process description. The blocks that are used to represent several pieces of equipment make it difficult to understand the process. For example, on page 5 of the application it is stated that the spent reducing gas (top gas) exits the reduction zone of the reactor through the refractory lined top gas duct and enters the top gas scrubber to be cleaned and cooled. After scrubbing and cooling, approximately two-thirds of the clean top gas (now called process gas) flows through a second set of mist eliminators and then to the inlet of the first stage process gas compressor, followed by a second compressor. This equipment is not shown on the process flow diagram. It is suggested that voestalpine revise the current process flow diagram by delineating further the equipment/components that comprise the reduction and reformer process. If voestalpine finds it beneficial or necessary, it is suggested that additional pages be created and provided to EPA. For clarity purposes, it may be beneficial that voestalpine provide additional pages for each separate process that refers to the original process flow diagram.

   Response:

   CP represents criteria pollutants meant to represent the combustion emissions such as PM, SO_2, NO_x, etc.

   New Process Flow Diagrams including information requested in Item 1 and other Items described below are included in Attachment 1.

2. In addition to the previously mentioned comment, please provide supplemental information to the process flow diagram and/or process description:

2.A. On page 2 of the application, in the "Iron Oxide Storage and Handling" section, it is stated that the oxide coating station enables feeding of coating directly to the charge hopper of the shaft furnace. Is the oxide coating directly fed to the furnace charge hopper shown on Figure 4? This addition does not appear to be shown on the process flow diagram. It is stated that the furnace feed conveyor discharges through a riffler to the charge hopper at the top of the shaft furnace. Please explain the purpose of the riffler. Is it a separate piece of equipment or is a part of the charge hopper? Also, it is stated that the oxide screening operation is two-fold in the storage and handling section and unusable material is discarded. Please provide supplemental information that explains where and how off-spec material is discarded. On page 3, the application states that the storage pile and associated operations are controlled with fugitive suppressants. Is this a continuous operation? Is it automated?

<u>Response:</u>

The coating station is before the vertical feed conveyor and therefore the coating material is not fed directly into the charge hopper. Please see the revised Process Flow Diagrams included in Attachment 1.

Additional Description for oxide handling and off-specification material handling:
The day bins discharge to a screen to separate the -6mm fractions from the desired +6mm to -20mm oxide fractions. The desired oxide fractions are discharged on the oxide transfer conveyor. The -6mm fractions are screened in the oxide screening station to separate the -3mm fractions from the +3mm to -6mm fraction. The +3mm to -6mm fraction, called oxide undersize, is discharged into the oxide undersize bunker. Oxide undersize material will be fed onto the oxide transfer conveyor.

The remet material is screened to separate the +35mm (remet oversize) and the -5mm (remetfines) fractions. Remet material will be fed onto the oxide transfer conveyor. The oxide coating station enables feeding of coating to the oxide transfer conveyor. The material on the oxide transfer conveyor is weighed and then discharged onto the furnace feed conveyor which is located where the DRI/HBI Core Plant begins. The furnace feed conveyor is a vertical, pocket type conveyor with flexible side walls that delivers material to the top of the shaft furnace structure. The furnace feed conveyor discharges through a grizzly and riffler to the charge hopper.

The charge hopper located at the top of the shaft furnace is equipped with a weight indicator which allows the operator to know the quantity of feed in the charge hopper. Through controllers, the day bin discharge feeders, the oxide undersize feeder, and the remet feeder are automatically stopped and started to control the level in the charge hopper. In addition, a Grizzly will be provided above the charge hopper.

Explanation of riffler:
The riffler is a tool for distributing the charge material in the charge hopper over a wider area in order to avoid accumulation of material in a single location within the charge hopper and can be understood as a distribution chute on the top but within the charge hopper.

Operation with suppressant:
The suppressant delivery system will be an automated process that will be controlled as a function of meteorological conditions (rain, humidity, solar impact, etc…) and operational experience. voestalpine intends to monitor the suppressant delivery closely to maximize suppressant effects while limiting wastewater accumulation and water demand. voestalpine has not established the frequency of suppressant addition at this time.

2.B. On page 3 of the application, in the “Reduction Reactor” section, it is stated that the use of the small flow of inert seal gas into the furnace through the seal leg prevents the escape of furnace gases to the atmosphere, while still allowing the free flow of material by gravity into the furnace without the use of lockhoppers. Is there a benefit to GHG emission production or furnace operating efficiency that is afforded the furnace using the inert seal gas which eliminates the need to use lockhoppers? Also, it is stated that the reduction reactor is not directly vented to the atmosphere so it does not have a specific emission source associated with it. Seal gas is used to pressurize both the top and bottom of the reactor so that the system reducing gases do not vent to the atmosphere. The process flow diagram does not appear to show the connection between the top and bottom seal with the seal system. Please update the process flow diagram to show the correct tie-ins. What is the proposed compliance strategy for ensuring that the seal system is properly functioning? What operating parameters will be monitored to ensure the seal system is maintaining a positive pressure seal around the reactor and reducing gases are not vented to the atmosphere?

Response:

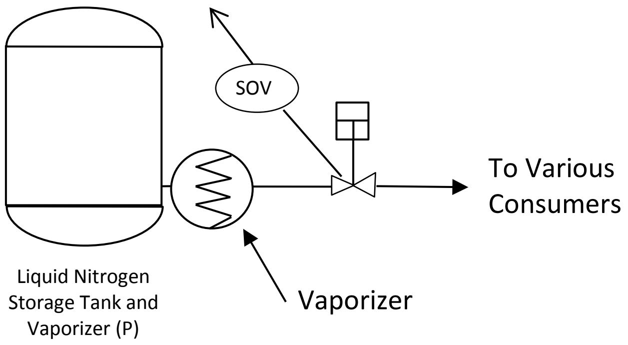
Explanation elimination of lockhopper with regard to GHG:
The free flow of material without lockhoppers is a unique characteristic of the MIDREX process to achieve homogenous and stable material flow from the charge hopper to the product discharge chamber and therefore important for operating efficiency. Stable process conditions minimize the use of natural gas and are therefore a benefit to minimize GHG emissions.

Description seal and purge gas system:
Inert seal gas for the plant, which is used primarily for sealing the top and bottom of the furnace, is provided by the seal gas generation system which takes a small partial flow of the hot reformer flue gas and prepares this partial flow for using it as “inert seal gas”. In case seal gas is not available (e.g. for the initial startup as well as after maintenance downtimes), a liquid nitrogen system will be used to supply purge gas.. voestalpine’s original application included an inert gas generation system instead of this liquid nitrogen system. With the inert gas generation system being replaced by a nitrogen supply system the following equipment will be eliminated: inert gas air compressor, inert gas generator, inert gas compressor, inert gas after cooler inert gas dryer, purge gas compressors, purge gas dryers, and purge gas tanks.
. Please see the revised Process Flow Diagrams included in Attachment 1.

Seal gas is necessary to separate prevailing process conditions from ambient air conditions. This is mainly to prevent the escape of furnace gases into atmosphere and to avoid contact of atmospheric oxygen and humidity with reduced metallic burden (i.e. product). The seal gas is compressed by a positive displacement type compressor and then cooled in a shell and tube aftercooler to remove the heat of compression. The cooled seal gas passes through a mist eliminator. Cooled seal gas is divided afterwards into wet seal gas network and via seal gas dryer into dry seal gas network. The seal gas dryer is a refrigerant type unit designed to remove moisture of the wet seal gas. The seal gas outlet temperature will be adjusted by an economizer to 10 – 15 °C above the due point (6°C) to avoid condensation. This dry seal gas is then distributed to the bottom seal or to the purge gas compressors.

In the case when seal gas may not be available (e.g. for the initial startup as well as after maintenance downtimes), a liquid nitrogen system will be used to supply purge gas. Purge gas is used to purge the combustibles out of the system when needed (such as for a maintenance shutdown). Also, the liquid nitrogen system will be the source for impulse purge for some instruments. The liquid nitrogen system additionally serves as a back-up supply source of some seal gas users, such as the bottom seal gas for the bottom seal leg.

Nitrogen Supply System:
The system comprises a liquid storage tank of 20,000 gallons liquefied nitrogen and an evaporating system (ambient conditions are believed to be adequate for nitrogen vaporization; no additional heat added) to transfer the nitrogen from liquid to gaseous status at a pressure of approximately 6.75 bar gauge



Description furnace bottom seal gas system:
The furnace bottom seal gas system consists of a compressor, a dilution hood, a dust collection scrubber, a fan, and a stack to supply and exhaust seal gas for sealing the bottom of the shaft furnace. The bottom seal gas compressor supplies dry seal gas to the lower seal leg of the shaft furnace at the required pressure. The compressor is a positive displacement type run dry to maintain the dry seal gas conditions. The bottom seal gas is vented through the Product Discharge Chamber (PDC) vent line, collected in the dilution hood, cleaned in the dust collection scrubber, and exhausted through the bottom seal dust collection fan and stack. The hood captures sufficient air to maintain a mixture of gases that remains below minimum explosive limits in the dust collection system. The scrubber is designed to remove the entrained dust particles exiting the PDC, before the seal gas / air mixture enters the fan and the stack.

The seal gas cooler is a packed bed type cooler, which is vessel of 3.5m diameter, which contains 9.5 m^3 Hypack #3 (or equivalent) packing rings out of stainless steel: cold process water is added on top of the scrubber and passed together with the hot gas over the surface of the packing, where the heat exchange from gas to water is performed which withdraws the energy from the hot gas. Inside the cooler the process water is saturated with CO_2 (according prevailing temperature and pressure). The water stream is then passed to the degassing units, where some CO_2 will be released which are already considered at that point. There will be no GHG emissions at the seal gas cooler itself.

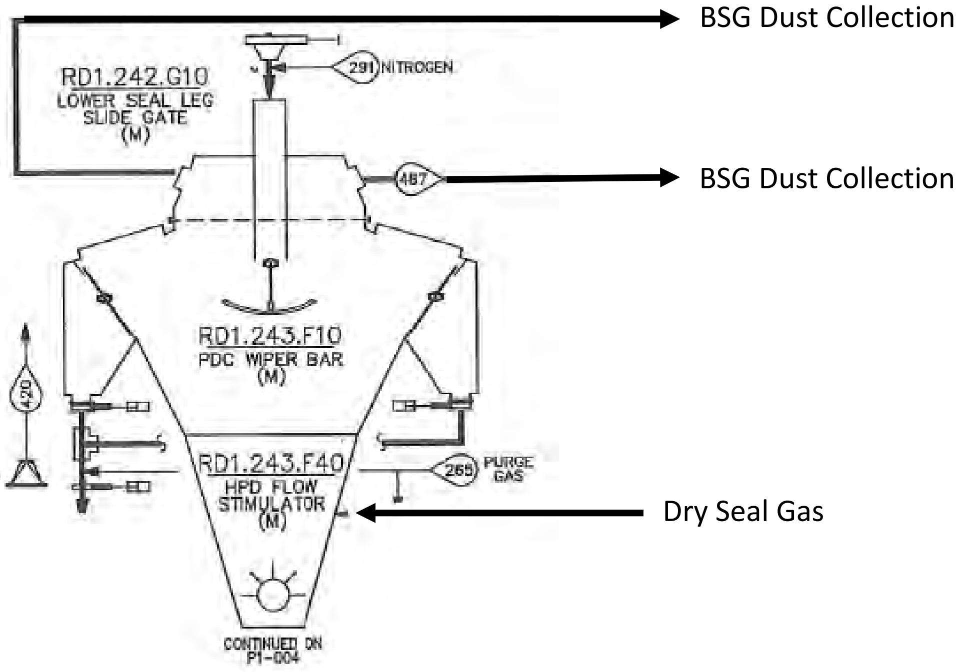
Compliance strategy for seal system: monitoring and operating parameters:
The oxygen content of the seal gas after the flue gas afterburner is automatically controlled; typical oxygen range is in the range of 0.3-0.5%. The seal gas outlet temperature will be adjusted by an economizer to 10 – 15 °C above the dew point (6°C) to avoid condensation. The pressure drop in the seal legs (which can be assumed as 0.1 bar per meter of material column) will be monitored by pressure sensors and the material flow by flow sensors (to be checked with PFD). In case of loss of pressure slide gates will automatically close the upper and lower seal leg.

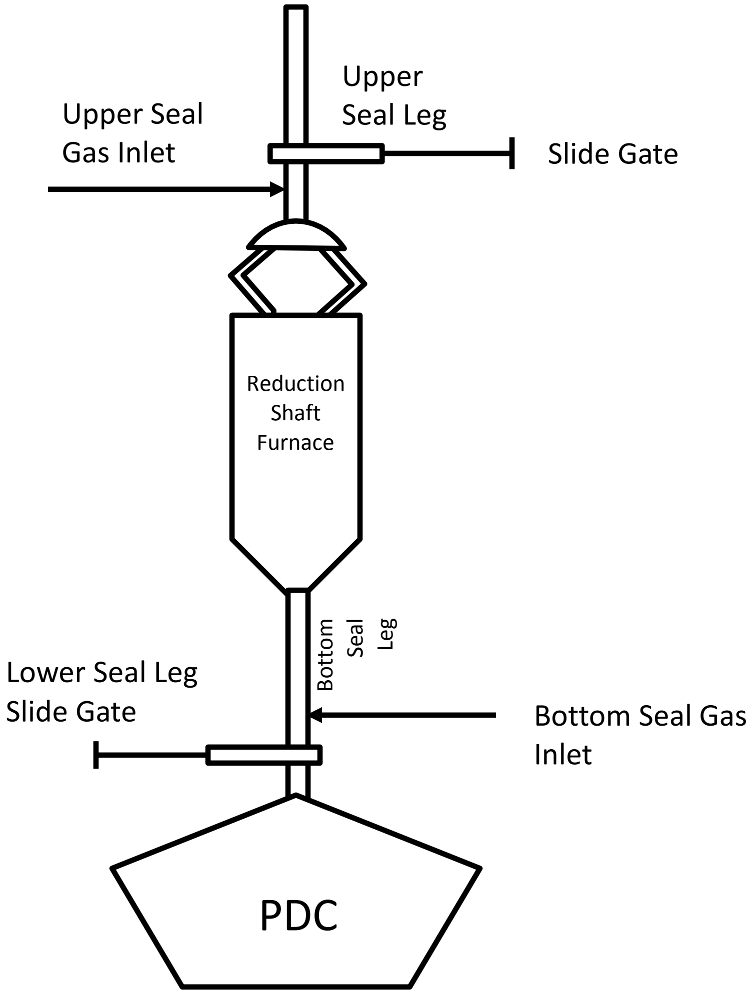
2.C. Beginning on page 3, in the “Hot Discharge System” section, it states that material is discharged from the furnace via a dynamic seal leg and a hydraulically driven wiper bar. This section also discusses material discharge from the lower cone to the lower seal leg, the use of a burden feeder, surge hopper and several feed legs. Is this equipment located internal to the furnace or is it external separate pieces of equipment? If possible, please show this equipment on the process flow diagram.

Response:

Location of equipment for discharge:
The burden feeder and the lower cone are part of the reduction shaft. The material flows through a pipe (i.e. lower seal leg) to the product discharge chamber with the whiper bar. The product discharge chamber is a unit directly connected via the seal leg with the reduction shaft. Then the material flow to the surge hopper and over seven feed legs to the briquetting machines. See below detail drawings for the discharge chamber.

Product Discharge Chamber




Upper Seal Leg
Upper Seal Gas Inlet
Slide Gate
Reduction Shaft Furnace
Bottom Seal Leg
Lower Seal Leg Slide Gate
Bottom Seal Gas Inlet
PDC

2.D. On page 4 of the application, in the "Hot Briquetting System" section, it is stated that off-specification product (remet) produced during plant start-up or process upset bypasses the briquette machines and is discharged through a bypass feed leg to the bypass discharge feeder and then to the hot briquetted iron (HBI) cooling system. Please show this bypass discharge feed on the process flow diagram. Also, this section includes an explanation of the dust collection system. It appears that this system has not been represented on the process flow diagram. Please update the process flow diagram to show the dust collection system.

Response:

Description hot discharge system:
The hot reduced material is discharged from the furnace via a dynamic seal leg and a hydraulically driven variable speed hot wiper bar. The design of the wiper bar allows it to be adjusted by shims up to ±20 mm. The speed of the lower burden feeder is controlled in ratio to the average discharge rate of the furnace to achieve a uniform flow of the material from the lower cone to the lower seal leg. The hot material flows across the wiper bar pan and then passes through a set of hydraulically driven screen sizers that limit the maximum size of the product passing into the surge hopper of the product discharge chamber. The oversize material can then be automatically discharged by means of a tramp chute to the ground level. The wiper bar pan contains slotted openings which can be opened by means of hydraulically operated gates. These openings allow the discharge of fines to help prevent buildup on the pan. The hot reduced material is discharged from the surge hopper into the feed legs. Seven feed legs are connected directly to the seven briquetting machines; one feed leg is connected to a hot rotary valve. The bypass hot rotary valve is designed for full production capacity. Each leg is isolated by a slide gate and ball valve for safety from its respective discharge device. Please see revised Process Flow Diagrams included in Attachment 1.

2.E. On page 5 of the application, in the "Process Gas System" section, it is stated that after scrubbing and cooling, approximately two-thirds of the clean top gas (now called process gas) flows through a second set of mist eliminators and two compressors. Please provide design efficiency data for the compressors. After compression the process gas is mixed with natural gas and preheated to form the feed gas for the reformer. Please update the process flow diagram to show the compressor. Also, please label this process gas line that is fed to the reformer. What heat transfer fluid is used to preheat the process gas in the heat recovery system? Please provide supplemental information to the process description pertaining to the heat transfer fluid used in the heat recovery system. The process flow diagram indicates that in addition to the preheating of the process gas, the top gas fuel is also preheated. Is this correct? If so, please update the process description with this information. Also, the process description states that the top gas fuel - after it is mixed with a small amount of natural gas - is passed through a mist eliminator to remove water droplets before fueling the reformer burners. Please update the process flow diagram to show this mist eliminator.

__Response:__

Design efficiency data for compressor:
The compressors in the process gas system are centrifugal machines specifically selected for the Midrex Direct Reduction Process. The process gas compressors are arranged with two (2) machines operated in series. These machines are designated the first stage and second stage process gas compressors. The flow rate is controlled by adjustable inlet guide vanes. The antisurge control for each compressor is done by a compressor surge protection valve. Each compressor can be operated stand alone for reduced plant capacity (approx. 60-75%) also. Overall Efficiency is 70 – 75 %, motor efficiency is 90 %; 8.4 MW output per compressor, differential pressure over both compressors is 1.6 bar. Please see revised Process Flow Diagrams included in Attachment 1.

Heat recovery system (preheating of process gas, burning air and top gas fuel):
In order to minimize the energy consumption of the Direct Reduction Plant, sensible heat contained in the hot flue gases coming from the reformer is recovered in the recuperator, or heat recovery unit. The flue gas from the reformer is used to preheat the reformer main burner combustion air, natural gas, top gas fuel and the feed gas streams. The total benefit of the heat recovery system is an increase in reformer capacity and a reduction in the net plant energy consumption. The system will consist of the bundles for preheating of the gas and air streams noted above. The recuperating system will consist of two identical units, each attached to one of the main flue gas headers running on side the reformer. Each recuperator unit will consist of a steel casing lined with insulation. Due to the flue gas inlet temperature of about 1170° C various types of insulating castables and block insulation are used. In idling condition when there is no flow to the main burners the maximum temperature in the hot air bundle is controlled by injection of cooling air into the in-coming flue gas stream.

Driven by a hot fan, the flue gases are exhausted from the reformer through the recuperator via the flue gas stack and are released into the atmosphere. The stack is a forced draft flue gas stack with a hot fan that draws the hot flue gas from the reformer through the preheaters situated in the recuperator. An inline fan is used to create the required suction. Due to process requirements the arrangement of tube bundles in relation to the direction of flue gas flow is as follows (Please see revised Process Flow Diagrams included in Attachment 1 and a new process flow diagram that focuses on heat recovery, Figure 4b):

Hot Air Tube Bundle:
The hot air preheaters are U-type heat exchangers made of high alloy heat resistant material suspended in the refractory lined recuperator. The airflow passes through the tube bundles. In the hot air bundles the Combustion Air for the burners will be heated up to the required temperature of approx. 600°C.

Feed Gas Bundles:
The feed gas preheaters are also U-type bundle heat exchangers, which are suspended in the refractory lined recuperator ducts and situated downstream from the combustion air preheater. The feed gas preheaters heat the process and natural gas in stages. Final preheat temperature is 600°C.

Top Gas Fuel Tube Bundle:
The bundles are of alloy steel material. These bundles preheat the total fuel gas to the main burners to 350°C.

Natural Gas Tube Bundle:
The bundles are of alloy steel material. These bundles preheat the natural gas to the bustle gas, transition zone natural gas to 350°C.

Cold Air Tube Bundle:
The bundles are of carbon steel material. This is the first step preheat of the combustion air.

2.F. On page 5 of the permit application, in the "Reformer" section, it is stated that natural gas-fired auxiliary burners maintain the reformer box temperature during plant idle conditions to minimize both restart time and thermal cycling of the reformer tubes. Are GHG emissions produced during these times? If so, have the GHG emissions been accounted for in the proposed GHG emission rates? How often will these plant idle conditions occur? Also, continuing on page 6 of the same section, it is stated that the flue gas exiting the reformer box via the flue gas headers flows to the heat recovery system where the waste heat is recovered. Please update the process flow diagram to show this heat recovery from the reformer flue gas.

Response:

Contribution auxiliary burners to GHG-emission:
The auxiliary burners are included in the overall emission calculations for the reformer. The auxiliary burners operate on a near continuous operation with the main burners and also have the use of being the primary heat source during idle modes to ensure that reformer maintains adequate temperature to reduce thermal cycling.

How often plant idle condition will occur:
voestalpine's intended (i.e. design aspects) operation is as follows
Total annual time: 8760 hours
Production hours: 7800 hours/a
Annual shutdown time: 480 hours
Idling time: max. 480 hours per year
Initially, the idling time can be assumed with 3 – 5 % of the total annual time. Especially during startup of the plant and during initial operation, the idling mode will happen more often due to unexpectable reasons. After gaining experience with the plant and stabilized operation, the idling mode can be avoided and the gained time will be shifted towards production hours.

Please see revised Process Flow Diagrams included in Attachment 1 and a new process flow diagram that focuses on heat recovery.

2.G. On page 6 of the permit application, in the "Heat Recovery System" section, it is stated that the flue gas exits on both sides of the reformer and enters the parallel train heat recovery system. Each parallel system contains combustion air pre-heaters and feed gas pre-heaters. Please update the process flow diagram to show this heat recovery system as explained in the process description. Also, it is stated that the heat recovery system increases the reformer capacity and reduces the net plant energy consumption by approximately 25-30 % over the first generation designed in the 1960's. Please provide technical literature, resources and/or calculations to substantiate this energy consumption claim. Please provide supplemental technical data that includes the design efficiency of the combustion air and feed gas pre-heaters. What parameters will be monitored and recorded to ensure the pre-heaters are operating as designed? What is the proposed compliance strategy for the heat recovery system? If possible, please provide benchmark data that compares the energy consumption of voestalpine's facility to similar sources, nationwide or international, in the direct reduction iron (DRI) industry.

Response:

Heat Recovery System:
Please see the detailed description provided in 2.E above and the revised process flow diagrams included in Attachment 1. Additional information on Process Control is provided below.

Gas quality is controlled by gas analyzers, thermocouples and flow meters. Mass and energy balance will be calculated by latest state of the art "Level 2" automation system.

Process Control
The process control is based on the one hand by indication and operation of the basic process automation system (Level 1 Basic Automation System) based on operator input and reaction. On the other hand a Level 2 Process Control System including process models and an expert system will be installed, which will monitor actual operation, calculate and balance current situation and deliver proposals for operator actions to be taken in advance to fulfill targeted operation results. The following basic process controls are fundamental to direct reduction:

Reductant-to-Oxidant Ratio (Quality of Reducing Gas)
This is controlled by a CO_2 analyzer-controller that uses the reformed gas analysis to regulate the addition of process natural gas to the process gas. The normal set point is about 2.5% CO_2 on a dry basis. The controller adds natural gas if the reformed gas CO_2 is too high. A lower CO_2 percentage means higher gas quality.

H2/CO Ratio
This is controlled at 1.5-1.6 by a virtual water analyzer-controller recorder that regulates the water temperature to the process gas side of the top gas scrubber. A higher water temperature will increase the H_2/CO ratio.

Bustle Gas CH_4

This is controlled by a CH_4 analyzer-controller that regulates the flow of enrichment natural gas to the reformed gas. The bustle gas CH_4 concentration is typically maintained at about 4.0%, with a low alarm at 2.0% and a high alarm at 5.0%.

Bustle Gas Temperature

If the bustle gas temperature is higher than desired, it is lowered by a temperature controller that regulates the flow of reformed gas through the reformer gas cooler. The cooled reformer gas is reintroduced to the reformer gas as necessary to regulate the bustle temperature for the chosen feed. Enrichment natural gas is also injected into the cooled reformer gas line just upstream of the reformed gas mixer. This mixture (with controlled temperature and methane content) is called bustle gas.

Control of Metallization

The operator compares the flow of process gas with the product discharge rate. For example, the operator may choose 900 Nm^3 or 1000 Nm^3 of process gas per metric ton (ton) of DRI depending on the desired metallization. The process gas flow controller holds a preset process gas flow. More precise control can be maintained by using the flow of reformed gas or bustle gas. However, the flow of hot reformed gas cannot be directly measured and must be calculated. Regardless of the gas used, the ratio itself is not directly controlled. Also, the change in metallization for a change in process gas flow/ton is not linear.

2.H. On page 6 of the permit application, in the “Seal Gas and Purge Gas System” section, it is stated that inert seal gas for the plant, which is used primarily for sealing the top and bottom of the furnace, is provided by the seal gas generation system. This system takes hot reformer flue gas and cools it in the seal gas generation system. The seal gas cooler is a packed bed, direct contact type cooler which cools the reformer flue gas to near ambient temperature. Please provide supplemental information on the operation of the direct contact type cooler. What type of heat transfer fluid is used? Is cooling water the heat transfer fluid? If so, the process flow diagram does not show a cooling tower as part of the proposed project. Please update the process flow diagram to show the cooling tower. Will the cooling tower be a potential source for GHG emissions? Since CO_2 emissions are associated with combustion pollutants and CH4 pollutant is associated with VOC pollutants, will the direct contact type cooler be a potential source for GHG emissions to enter the cooling water system? Are there other sources where cooling water is used for heat exchangers and due to a process leak GHG emission could potentially enter the cooling tower system? If voestalpine feels that such streams do not have GHG pollutants an explanation is required. If there is a possibility for GHG emissions, please supplement the BACT analysis with an evaluation of leak repair and monitoring technologies and a proposal of what voestalpine would implement as BACT. Part of the dry seal gas is compressed by the purge gas compressors and dried in a desiccant dryer. The dry purge gas is stored in tanks to be used for emergency plant shutdown situations and for small high pressure requirements during normal operation. The process flow diagram does not have a representation of the storage tanks. Please update the process flow diagram to show these tanks. Where will the storage tank vents be directed? If so, the combustion of the tank vapors might generate GHG emissions, therefore a BACT analysis should be developed for the tanks to be installed for the project. Please be sure to incorporate into the tank BACT analysis the factors that were considered for the design of the tanks. Please provide any other additional information for the tanks, including whether the applicant chose to have the tanks painted white or another color of high refractive index to reduce vapor production?

Response:

Seal gas – purge gas system:
With regard to seal gas and purge gas system please see description in Response 2.B above. Please see the revised process Flow Diagrams included in Attachment 1.

Operation of the direct contact type cooler:
The seal gas cooler is a packed bed type cooler, which is vessel of 3.5m diameter, which contains 9.5 m^3 Hypack #3 (or equivalent) packing rings out of stainless steel: cold process water is added on top of the scrubber and passed together with the hot gas over the surface of the packing, where the heat exchange from gas to water is performed which withdraws the energy from the hot gas. Inside the cooler the process water is saturated with CO_2 (according prevailing temperature and pressure). The water stream is then passed to the degassing units, where some CO_2 will be released which are already considered at that point. There will be no GHG emissions at the seal gas cooler itself.

Direct water cooling system for Spent Gas – description degassing system:
All partial flow rates from process water system (including direct cooling) are collected and treated at the process water treatment plant. Due to the required direct contact of process water with process gases (top gas scrubber in particular) some traces of gas constituents are dissolved in water at elevated process gas pressure. In order to reduce the fugitive emissions at the clarifier and to reduce scaling in process water ducts a forced degassing will be installed. The return water flow from the Top gas scrubber weir (after depressurization) is routed to a degasser vessel where air (coming from a separate air fan) flows countercurrent to the hot return water (see picture below of a process water degasser). The majority of the dissolved gas constituents (CO_2, CO, etc.) are collected in the degasser vent gas flow and released to atmosphere at safe location. The degassed water from the degasser is flowing to the clarifier. The emissions from the degasser are estimated and included with the revised emissions estimate included with Attachment 2. The emissions from the degasser are included in the revised GHG BACT submitted under separate cover.

Picture of process water degasser

2.I. On page 6 of the application, in the “Bottom Seal Gas System” section, it is stated that the furnace bottom seal gas system consists of a compressor, a dilution hood, a dust collection scrubber, a fan, and a stack to supply and exhaust seal gas for sealing the bottom of the shaft furnace. Is the “Bottom Seal Gas System” a part of the same seal gas system that is discussed in the previous section on page 6 (Seal Gas and Purge Gas System) or is it a different system? If so, please update the process flow diagram to show the different seal gas systems that exist for the top of the furnace and the bottom of the furnace. Please indicate the process line tie-ins on the process flow diagram to the seal gas system that provides seal gas to the top seal of the furnace. Please provide a supplemental process diagram that delineates the different sections to the seal gas system. Is the previously mentioned seal gas stack directed to the flare? If so, has this been accounted for in the emission calculations for the flare? If not, where is the stack exhaust directed? What is the design efficiency of the equipment that comprises the seal gas system (i.e., compressor, scrubber, fan, dilution hood)? What is the proposed compliance strategy for the seal gas system? What parameters will be monitored and recorded to ensure this system is operating according to design? How will on-site personnel determine if the system(s) is working properly?

Response:

Description furnace bottom seal gas system:
With regard to seal gas and purge gas system please see description in Response 2.B above. Please see the revised process Flow Diagrams included in Attachment 1.

The furnace bottom seal gas system consists of a compressor, a dilution hood, a dust collection scrubber, a fan, and a stack to supply and exhaust seal gas for sealing the bottom of the shaft furnace. The bottom seal gas compressor supplies dry seal gas (nitrogen) to the lower seal leg of the shaft furnace at the required pressure. The compressor is a positive displacement type run dry to maintain the dryseal gas conditions. The bottom seal gas is vented through the PDC vent line, collected in the dilution hood, cleaned in the dust collection scrubber, and exhausted through the bottom seal dust collection fan and stack. The hood captures sufficient air to maintain a mixture of gases that remains below minimum explosive limits in the dust collection system. The scrubber is designed to remove the entrained dust particles exiting the PDC, before the seal gas / air mixture enters the fan and the stack.

*Seal gas stack to flare***:**
No seal gas is routed to the flare;

Design efficiency equipment seal gas system:
With regard to seal gas and purge gas system, please see description in Response 2.B above.. Furthermore, Response 4 provides some additional background to voestalpine’s procedures for equipment procurement. Please see the revised process Flow Diagrams included in Attachment 1.

Compliance strategy for seal system: monitoring and operating parameters:
Compliance strategy for the seal gas system is discussed in the response to 2B above.

How will onsite personnel will determine correct operation:
For plant operation there is a main control room where the plant operators are able to observe/determine correct operation conditions.
A PC-based Visualization System (HMI-System → Human Machine Interface) will be installed. The main features of the HMI system are:

- Menu based window system, which gives the same 'look and feel' as typical Windows applications.
- Fast access to frequently used windows via tool bars.
- Minimum of operator input by usage of selection boxes.
- Operation using mouse and/or keyboard.
- Presentation of process data and machine components in graphical form.
- Plant Area and Process Tasks oriented organization of the HMI applications displays.
- Full functionality available at every client operator station within a server domain,
- allowing to use alternate client operator stations in case of failure of a single station.
- Availability of Level 1 and Level 2 HMI functions on each client operator station.

The HMI-System allows:

- Equipment mode selection (Auto, Manual, Local) of equipment
- Plant Mode selection and start / stop of process
- Switching of all auxiliary systems and functions such as hydraulic system etc. for equipment that allows remote operation
- Control and monitoring of the production process
- Display of equipment status, interlock information and Plant Mode preconditions and their status
- Manual intervention when required

Safety related features of the HMI-System are:

- Operation of plant areas/machine components restricted to respective location of a client operator station
- Access level according to defined user groups; e.g.; operator, maintenance, metallurgist etc. For example, certain parameters may only be changed by the metallurgist, whereas some functions are only allowed to maintenance user
- Where additional safety is required, process oriented functions or operations require additional confirmation by the operator via buttons on the display
- Plausibility and limit checks on operator inputs

2.J. On page 7 of the application, in the “Inert Gas System” section, it is stated that an inert gas system supplies seal gas for the plant in the event that the reformer combustion system is not in operation. This system consists of an inert gas generator where natural gas and air are burned at close to the stoichiometric ratio so that the product of combustion yields a suitable inert gas with very low oxygen content. This generator doesn’t appear to be represented on the process flow diagram. Is this a potential GHG emission source? If so, please update the process flow diagram to show generator with associated EPN, as well as, providing supplemental emission calculations. Please provide a 5-step BACT analysis for this generator. Please provide the design efficiency for the generator. How will this system be monitored and controlled? In the event that seal gas is interrupted due to the reformer combustion system not operating, is the switch to inert gas automated? Is there continuous monitoring of certain operating parameters that alert on-site personnel to system problems? What is the proposed compliance strategy for this inert gas system? What parameters are monitored and recorded to ensure the mechanism that triggers the switch is operating and will operate properly? Is there a preventative maintenance schedule on these process controllers? Please update the process flow diagram to show the inert gas system and tie-ins to the seal gas system.

Response:

Inert gas system:
The inert gas generation system (system for providing seal gas in case of special process conditions - e.g. for the initial startup as well as after maintenance downtimes), and storage will be replaced by liquid nitrogen storage and vapor delivery system. Please see description in Response 2.B n*itrogen supply* above. With the inert gas generation system being replaced by a nitrogen supply system the following equipment will be eliminated: inert gas air compressor, inert gas generator, inert gas compressor, inert gas after cooler inert gas dryer, purge gas compressors, purge gas dryers, and purge gas tanks. Please see the revised process Flow Diagrams included in Attachment 1.

2.K. On page 7 of the permit application, in the "Machinery and Process Cooling Water System" section, it states that the water system consists of a machinery cooling water circuit and a process cooling water circuit. Please provide a separate process flow diagram that depicts both systems. The machinery cooling water is a closed circuit that supplies cooling water to all indirect coolers such as burden feeders, rotating equipment lubrication oil, heat exchangers, etc. Please provide supplemental information on the operation and design efficiency of the burden feeders proposed for the project and how cooling water is utilized. The process cooling water circuit supplies cooling water to the direct contact coolers and the process users, such as the top gas scrubber and the dust collection systems. It also provides the cooling water for the machinery cooling water heat exchangers. The process cooling water system consists of a sump, circulation pumps, process water cooling towers, and a clarifier system. Is it possible for GHG emissions to be present in the process water cooling towers due to process equipment leaks into the system or CO_2 entrainment?

<u>Response:</u>

Cooling water system:
The water system consists of machinery cooling water circuit and two process cooling water circuits, which are re-cooled with a secondary cooling circuit (non-conduct with primary circuits) by means of sea water cooling towers. Please see the Process Flow Diagrams for the water use at the facility. One is a plant-wide

<u>Sea water cooling System:</u>
The hot water returned from the heat exchangers is led through the sea water cooling towers and collected in the cold sea water basin. The sea water cooling tower drift will be according the required value from the BACT analysis. Different pump groups are installed depending on the process requirements, operation flexibility and plant maintenance requirements. Each pump group is additionally equipped with stand-by pump capacity. The cooled water with a temperature of approx. 32°C will be pumped by the cold water pumps to the heat exchangers, which are located at the water treatment plant area, in order to cool the different process cooling water circuits at the core plant area. Instrumentation for the indication of cooling-water temperatures, pressures and flow rates is installed.

<u>Machinery cooling water System:</u>
The machinery cooling water is designed as a closed circuit which supplies cooling water to all indirect coolers such as burden feeders, rotating equipment lubrication oil heat exchangers, etc. The machinery cooling water system basically consists of circulation pumps and heat exchangers. The circulation pumps circulate hot water from the machinery cooling water pond through the sea water cooled heat exchangers to the consumers in the core plant area with a cold water temperature of approx. 35°C.

<u>Contaminated Process Water System:</u>
The hot water returned from the process consumers is collected in the hot water pond. Different pump groups are installed depending on the process requirements, operation flexibility and plant maintenance requirements. Each pump group is additionally

equipped with stand-by pump capacity. The cold water stream of the contaminated process water is pumped by the process water pumps from the hot water pond via the sea water cooled heat exchangers to the consumers in the core plant area with a cold water temperature of approx. 35°C. The hot water stream of the contaminated process water is pumped by the process water pumps directly from the hot water basin to the consumers in the core area with a hot water temperature of approx. 70°C. Instrumentation for the indication of cooling-water temperatures, pressures and flow rates is installed.

The contaminated return water from the gas scrubbers in process area is cleaned via a classifier / clarifier-system. The contaminated process water stream with high solid content from the scrubbers and dedusting plant is conducted to the process classifier for an initial cleaning. The classifier will be designed for continuous processing to separate the coarse grain from slurry water. The settled coarse particles within the classifier will be removed by screw mechanism via a discharge chute. The slurry overflow of the classifier will be fed into the clarifier water collecting pond where the contaminated water from all process gas cleaning systems is collected and flows from there into the clarifier. In the clarifier the suspended solids as well as the precipitated carbonates are separated by gravity. The clarifier overflow is directed into the hot process water pond. From the clarifier underflow the slurry is pumped to the chamber filter press. The filter cake will be conveyed via belt conveyor to stock yard. In case of emergency the slurry settling pond will be used. The dewatered slurry is unloaded by a front loader onto trucks. Contaminated cold cooling water basin will have a bottom slope to a sump to ensure collecting of settled solids. The sump will be equipped with a sump pump.

Clean Process Water System:

The return water from the seal gas/reformed gas cooler and the inert gas generator will be directly routed to the clean warm water pond. The warm water is pumped by the clean process water pumps from the warm water pond via the sea water cooled heat exchangers, which are located at the central water treatment area, to the corresponding consumers of the core plant area with a cold temperature of approx. 35°C. Instrumentation for the indication of cooling-water temperatures, pressures and flow rates is installed. Any suspended solids are kept within the demanded limits by means of side-stream filtration. The backwash sequence depends on the filter pressure drop and is started automatically. The filters are backwashed with a combined air-water backwash program. Contaminated backwash water is conveyed to the contaminated process water system. The blow down water from the Clean Process Water System will be implemented to the Contaminated Process Water System. Water samples are to be taken and analysed for conformance. Make up water is automatically refilled to compensate for water losses caused by evaporation and blow down.

Degassing unit / GHG-discussion:

All partial flow rates from process water system (including direct cooling) are collected and treated at the process water treatment plant. Due to the required direct contact of process water with process gases some traces of gas constituents are dissolved in water at elevated process gas pressure. In order to reduce the fugitive emissions at the clarifier and to reduce scaling in process water ducts a forced degassing will be installed. See description in 2H above.

Provide technical data regarding operation and design efficiency of burden feeders and cooling water utilization:
The discharge zone consists of the refractory lined furnace cone equipped with hydraulically operated burden feeders and a flow aid insert, essential for uniform flow of the material within the furnace. Uniform material flow in the furnace ensures consistent, homogeneous product quality. In this hot discharge furnace, the reduced iron is discharged from the furnace cone at ideal briquetting temperatures of over 700 degrees C. Since the burden feeders get into direct contact with the hot DRI material, internal water indirect cooling is necessary in order to obtain longterm availability of the equipment. The cooling water supply is from the closed cooling water system, which mainly comprises a cooling water pond, some pumps, a heat exchanger system and an emergency tank, which is located inside the Midrex tower. The differential temperature of cooling water between entrance and outlet of the equipment is only approx. 10 °C.

2.L. Please update the process flow diagram by including the emergency generator and fire pump with the associated emission point number (EPN). Please provide design efficiency data for the emergency generator.

Response:

Process Flow Diagram:
Please see the revised Process Flow Diagrams included in Attachment 1.

Design efficiency of emergency generator:
The emergency engines have not been identified at this time. The respective use (electrical generation and mechanical energy) and EPA regulatory requirements such as the New Source Performance Standards (40 CFR Part 60) will guide voestalpine to an appropriate unit.

However, the general operations understanding are provided. To enable supply of electric power in case of failure of main power, an emergency diesel set is used for powering selected electrical consumers step by step in order to keep the plant and equipment in a safe condition. The diesel engine and alternator are direct coupled and mounted on a common base frame. Design technical data identified below:

General Characteristics	
Operation conditions	• Automatic start and automatic energizing of the emergency supply switchboard in case of mains failure. • Synchronizing of the emergency supply switchboard bus bar voltage with the mains power supply voltage once the mains voltage returned. • Manual stop of the emergency diesel set after the emergency loads have been transferred to the mains supplied bus bars. • Test and parallel operation of the emergency diesel set with the mains power supply.
Diesel engine	• Water cooled, four-stroke stationary type engine • For emergency operation • Electrical starting device and lead-acid battery • Day tank • Silencer, compensator and exhaust pipe • With preheating if required, to guarantee the automatic start and taking over of load in a short time • Air-water cooler with closed water circuit • With all required auxiliaries, speed control, monitoring and protection devices
Alternator	• Brushless, synchronous alternator • Direct coupled to engine • With anti-condensation heater and winding temperature protection • With built in current transformers for monitoring, protection and excitation
Electrical control cabinet	• With all control circuits for automatic start, synchronizing, manual stop, test

General Characteristics	
	• Battery charger for float and boost charging of starter battery • Indicating-, monitoring- and protection devices for diesel engine, alternator and auxiliaries • Automatic synchronizing equipment • Setting devices for speed and voltage • Voltmeters, amp-meters, watt-meters, power factor meter, synchronizing instruments, frequency meter, alternator temperature indicator

Technical Date - Alternator	
Alternator type	Brushless, synchronous alternator with damper winding for unbalanced load of 20%
Power Factor	0.8
Voltage	13.8kV, +/-5%, 3-phase
Protection	Suitable for indoor installation, depending on manufacturer’s standard

3. On page 2 of the permit application, it is stated an important quality of the reducing gas is the reductant/oxidant ratio, or “gas quality.” The quality is a measure of the potential for the gas to reduce iron oxide and is a ratio of reductants to oxidants contained in the gas:

   Quality = reductant/oxidant ratio = moles (H_2 +CO)/ moles (H_2O+ CO_2)

   Experience has found that the optimum gas quality for hot, fresh reducing gas should be 10 or higher. Also, to obtain essentially complete reduction, the quality of the spent reducing gas exiting the process should be at least 2. What operating parameters are controlled to ensure the gas quality is maintained at optimum levels? Another important property of the reducing gas is the H_2/CO ratio. Control of the H_2/CO ratio affords thermally balanced reduction reactions since reduction with carbon monoxide is exothermic, and reduction with hydrogen is endothermic. That is, the heat required by the hydrogen reaction is balanced by the heat supplied by the carbon monoxide reaction. Therefore, proper reduction temperatures can be maintained without significant additional heat input from fuel combustion. The typical H_2/CO ratio produced by the reformer is about 1.55:1. What is the proposed compliance strategy for the H_2/CO ratio? What parameters are monitored and controlled to maintain the optimum H_2/CO ratio for fuel usage efficiency?

   <u>Response:</u>

   Operating parameters for optimized gas quality, Compliance strategy, Monitoring/Control parameters for fuel efficiency:

   The response provided for Question 2.G is also applicable to this response. Please see Response 2.G above.

4. On page 15 and 16 of the application voestalpine included tables that list “possible” energy efficiency design improvements that can reduce fuel consumption and electricity usage. Please provide supplemental information detailing the anticipated percent efficiency gains and/or reduced GHG production with the implementation of the design attributes that “will” be employed by voestalpine from the tables provided. Also, provide a copy of any technical resources used to evaluate the design decisions for the voestalpine facility and any benchmark comparison data of similar sources existing nationally or internationally, that may have been utilized in the design selection strategy. Please provide technical resources, literature and calculations to substantiate the claimed efficiencies.

Response:

Energy efficiency design improvements:
Efficiency data for mentioned measures are dependent on available equipment on the market. There are design standards in place what regulate criteria regarding efficiencies etc. to use state-of-the-art equipment. The equipment purchasing process (and requests for vendor bid on equipment) will be implemented in a way that the suppliers will be bound by contracts to be in compliance with the relevant standards.
The supplies and services are in principle based on some of the following standards:

- DIN (Deutsche Institut für Normung e.V.)
- ISO (International Standard Organisation)
- VDE (Verein deutscher Elektrotechniker)
- IEC (International Electrotechnical Commission)
- FEM (Federation Européenne de la Manutention)
- IEEE (Institute of Electrical and Electronic Engineers)
- Mandatory Local Standards for local supplies and services
- Mandatory Country Regulation and Safety Laws
- ASME (American Society of Mechanical Engineer)
- ANSI (American National Standards Institute)
- API (American Petroleum Institute)
- ASTM (American Society for Testing Mechanical)
- ANSI (American National Standard Institute)
- NFPA (National Fire Protection Association)
- AGMA (American Gear Manufacture Association)
- AMCA (FAN Design Construction)
- IBC (International Building Code)
- NEMA National Electric Manufacturers Association
- OSHA Occupational Safety& Health Administration
- NEC National Electric Code

It is not practicable to quantify in detail the percent efficiency gains and/or reduced GHG production with regard to each in the tables listed measures at the current project phase due to above mentioned reasons. Generally accepted engineering ranges for performance of the state-of-the-art equipment over previous generation can be summarized as follows:

- The potential energy savings for insulating estimated to range from 2 to 5 percent
- The heat recovery system (recuperative preheating of air, fuel etc.) reduces the net plant energy consumption by approximately 25-30%.
- High-efficiency alternating current (AC) motors can save 1 or 2 percent of the electricity consumption of conventional AC drives
- Training programs and good housekeeping programs help to decrease energy consumption throughout the plant. Some estimates place the energy savings at 2 percent of total energy use
- Energy monitoring and management systems help provide for optimal energy recovery. These systems may reduce energy consumption by 0.5 percent

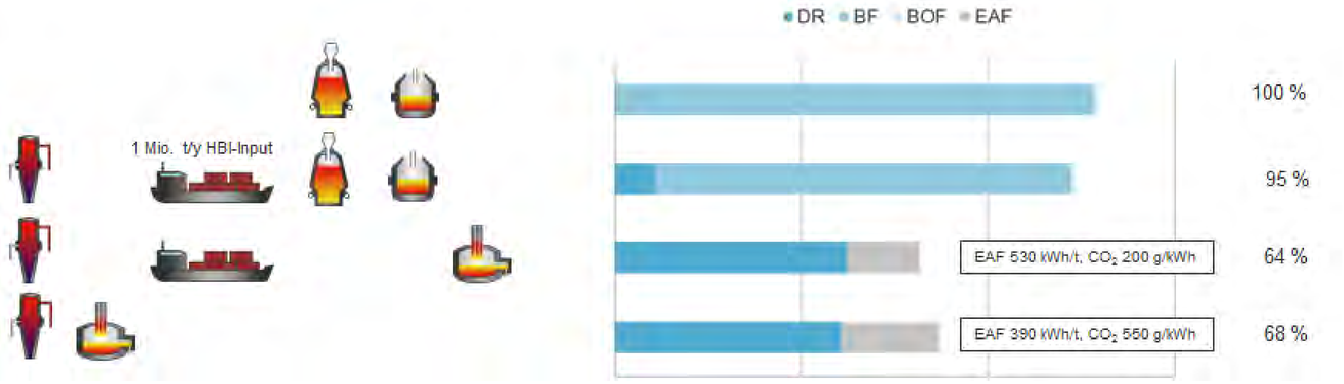
At the best case steady state consumption is 2.5 Gcal per metric ton (equivalent to ~10.4 decatherms per short ton). However, this value does not consider startup, shutdown, off-specification material re-processing, or adjustments for different ores, etc. It is voestalpine's intent to meet this steady state consumption at steady state operations, but to account of the greenhouse gases released during non-steady state operations, emissions are estimated with a value of 13 decatherms per short ton. This value is a previous BACT for another direct reduced iron facility.

Informational resource:
As stated in the USEPA Office of Air and Radiation document entitled *"Available and Emerging Technologies for Reducing Greenhouse Gas Emissions from the Iron and Steel Industry"* (can be provided at EPA's request) steelmaking by using DRI is a so called "Near-Term Technology." voestalpine believes with regards to GHG emissions that this is an invaluable in-house developed resource for the EPA. For example the DRI/EAF integrated steel-making route requires less energy and produces lower emissions than traditional integrated iron and steelmaking (i.e., coke battery, blast furnace, BOF). The US EPA document references a USDOE (2008) report claiming the reduction in emissions relative to traditional steelmaking for CO_2 up to 41 %. voestalpine is operating integrated steel mills in Europe. To transfer voestalpine's steelmaking into a low-carbon-based one in a first step a DRI/HBI – plant will be erected to follow the European Union's Roadmap for a low-carbon-economy in the steel industry. Additional, energy related information is provided in Attachment 3 with regards to the vendor supplied information.

Based on the road map, potential reductions using Direct Reduced Iron / Hot Briquetted Iron can be seen as follows:

▪ Input HBI to Blast furnace (BF) – LD-Steelmaking route (BOF)	▪ *Max. achievable CO_2-Reduction*.............................. 5 %
▪ Input HBI/DRI to EAF	▪ *Max. achievable CO_2-Reduction* .... 36 %

Using the base case of Blast Furnace (BF) and Basic Oxygen Furnace (BOF), the reduction potentials are summarized in the graphic, with detailed calculations in support of the basis provided below (support data is included in Attachment 4):



- **Basic Data:**

Power requirement O2-production	kWh/m3	0.56
CO2-emissions of generation of Electricity	kg/kWh	0.55 USA
CO2-emissions of generation of Electricity	kg/kWh	0.20 Austria
Calorific value natural gas	kcal/m3	8669
Input hot metal into BOF	kg/t	0.85
Input HBI into BF	kg/t	max .200
C-content coke (BF)	%	89
C-content coal (EAF)	%	85

- **Baseline BF** (basis: voestalpine consumption figures)

Coke	kg/$t_{hot\ metal}$	458
O2	m3/$t_{hot\ metal}$	70
Electricity	kWh/$t_{hot\ metal}$	30
$CO2_{BF}$	kg/$t_{hot\ metal}$	1508 (=$CO2_{coke}$ + $CO2_{O2}$ + $CO2_{Electricity}$)

- **Optimization BF : Input 200 kg/t HBI into BF – Austria** (basis: voestalpine consumption figures)

Coke	kg/$t_{hot\ metal}$	398
O2	m3/$t_{hot\ metal}$	70
Electricity	kWh/$t_{hot\ metal}$	30
$CO2_{BF\text{-}HBI}$	kg/$t_{hot\ metal}$	1313 (=$CO2_{coke}$ + $CO2_{O2}$ + $CO2_{Electricity}$)

- **Baseline BOF** (basis: voestalpine consumption figures)

O2	m3/t_{steel}	50
Electricity	kWh/t_{steel}	25
$CO2_{BOF}$	kg/t_{steel}	11 (=$CO2_{O2}$ + $CO2_{Electricity}$)

- **Baseline HBI-production** (basis: MIDREX/ENERGIRON)

Natural gas	Gcal/t_{HBI}	2.4
O2	m3/t_{HBI}	57
Electricity	kWh/t_{HBI}	115
$CO2_{HBI}$	kg/t_{HBI}	635 (=$CO2_{natural\ gas}$ + $CO2_{O2}$ + $CO2_{Electricity}$)

- **Baseline DRI-production** (basis: MIDREX/ENERGIRON)

Natural gas	$\text{Gcal/t}_{\text{HBI}}$	2.35
O2	m3/t_{HBI}	53
Electricity	$\text{kWh/t}_{\text{HBI}}$	105
CO2_{DRI}	kg/t_{HBI}	616 (=$\text{CO2}_{\text{natural gas}} + \text{CO2}_{\text{O2}} + \text{CO2}_{\text{Electricity}}$)

- **EAF-HBI - Austria** (basis: Table 5 consumption figures/cold DRI)

Coal	$\text{kg/t}_{\text{steel}}$	27
O2	$\text{m3/t}_{\text{steel}}$	33
Electricity	$\text{kWh/t}_{\text{steel}}$	530
Total $\text{CO2}_{\text{EAF-HBI}}$	kg/t_{HBI}	194 (=$\text{CO2}_{\text{coal}} + \text{CO2}_{\text{O2}} + \text{CO2}_{\text{Electricity}}$)

- **EAF-DRI = hot link DRI-EAF - USA** (basis: Table 5 consumption figures/hot DRI)

Coal	$\text{kg/t}_{\text{steel}}$	13
O2	$\text{m3/t}_{\text{steel}}$	35
Electricity	$\text{kWh/t}_{\text{steel}}$	390
Total $\text{CO2}_{\text{EAF-DRI}}$	kg/t_{HBI}	266 (=$\text{CO2}_{\text{coal}} + \text{CO2}_{\text{O2}} + \text{CO2}_{\text{Electricity}}$)

- **Discussion different Scenarios – CO2-emissions per t Steel-product:**
  - **Scenario 1: Steel production via BF-BOF-Route in Austria**
    - $\text{CO2}_{\text{scenario 1}} = \text{CO2}_{\text{BF}}$ X Input hot metal into BOF + CO2_{BOF} = 1508 X 0.85 + 11 = **1293 $\text{kg/t}_{\text{steel}}$**
    - $\text{CO2}_{\text{scenario 1}}$ = base line → **100 %**
  - **Scenario 2: Steel production via BF-BOF-Route in Austria using HBI**
    - $\text{CO2}_{\text{scenario 2}}$ = ($\text{CO2}_{\text{BF-HBI}}$ + X Input hot metal into BOF) + CO2_{BOF} + (CO2_{HBI} X Input HBI into BF X Input hot metal into BOF) = 1313 X 0.85 + 11 + 635 X 0.2 X 0.85 = 1235 $\text{kg/t}_{\text{steel}}$
    - $\text{CO2}_{\text{scenario 2}}$ compared to Scenario 1 : 1235 / 1293 = 0.95 → **95 %**
  - **Scenario 3: Steel production via EAF-Route in Austria using HBI**
    - $\text{CO2}_{\text{scenario 3}} = \text{CO2}_{\text{EAF-HBI}} + \text{CO2}_{\text{HBI}}$ = 635 + 194 = 829 $\text{kg/t}_{\text{steel}}$
    - $\text{CO2}_{\text{scenario 3}}$ compared to Scenario 1 : 829 / 1293 = 0.64 → **64 %**
  - **Scenario 4: Steel production via EAF-Route in USA using DRI (hot link)**
    - $\text{CO2}_{\text{scenario 4}} = \text{CO2}_{\text{EAF-DRI}} + \text{CO2}_{\text{DRI}}$ = 616 + 266 = 882 $\text{kg/t}_{\text{steel}}$
    - $\text{CO2}_{\text{scenario 4}}$ compared to Scenario 1 : 882 / 1293 = 0.68 → **68 %**

5. Beginning on page 17 of the permit application, it states that the DRI process results in far lower CO_2 emissions than conventional methods (compared to a blast furnace or other traditional approach). The main reason given for this reduction is that a DRI plant uses natural gas as a fuel instead of coke. In addition, on page 18 of the application, it is stated that the most common technologies used for natural gas-based DRI production are Midrex and HYL III and at this time, voestalpine has not selected a reformer supplier. Brief summaries of the Midrex DR and HYL DR processes were included in the application. Also, voestalpine did provide EPA supplemental information dated February 1, 2013 that states the Midrex and HYL processes are technologically and economically similar and that there are only plant-specific differences. It is not clear which technology will be used by voestalpine for the proposed project. Please provide supplemental information that informs EPA of voestalpine's intent. Did voestalpine perform a technical assessment that evaluates the two technologies of reformer-based DR to the reformer-less DR? Please provide supplemental information that includes this technical assessment. If possible, please provide any data that compares actual energy efficiency (fuel consumption and electricity consumption), actual GHG emissions and non-GHG emissions from the two technologies, and the technical resources that were used to perform the evaluation. Are there unique reasons to voestalpine for choosing one technology instead of another or selecting design options that aren't inherently more efficient or lower polluting than another (i.e., available feedstock, customer product purity specification)? If so, please provide detailed discussions on voestalpine's business purpose and objectives that affected design selections. If applicable, include a discussion on where GHG control strategies affect emissions of other regulated pollutants.

Response:

Comparisons HYL vs. MIDREX:
voestalpine provided detailed information comparing HYL and Midrex to the US EPA on June 10, 2013, and September 9, 2013, via email from our environmental consultant at Environmental Resources Management. Internal technological assessment was performed for both technologies which resulted in neither technology being inherently superior with regards to GHG emissions. This was not surprising to voestalpine, as both technologies use the same chemistry and stoichiometry within their respective technologies. Hence, the decision for MIDREX technology was amongst other issues an economically driven arbitration.

6. Beginning on page 24 of the application, Table 4-4 is a summary of the proposed GHG BACT determinations for voestalpine. The proposed BACT facility-wide emission limit is 1,814,144 tons/year. The proposed BACT emission limit for the reformer main flue ejector stack is no more than 13 MMBtu (decatherms) of natural gas/tonne HBI. Compliance for the reformer main flue stack is based on total natural gas consumption divided by total production (including regular and off spec DRI product) of the facility on a 12-month rolling total. In addition, the seal gas vent and the hot pressure relief vent (flare) are both included in the proposed limit of 13 MMBtu (decatherms)/tonne HBI produced. Please update the proposed emission limit by providing all emission calculations in short tons. Please provide any supplemental calculations that show how the proposed emissions will be calculated to determine compliance. What formulas will be used? What measurement indicators will be used? (i.e., natural gas flow for feed and fuel, flare vent flow, seal gas vent flow, weight scales for feed, product and off-spec product). Will these measurements be monitored and recorded continuously? The proposed emission limit is based on what design HBI production for the facility? Please provide the calculations and a basis for the rationale used to derive the proposed BACT limit. Also, the table identifies each emission source along with a source number. Please update the process flow diagram by identifying the emission sources listed and include the associated EPN.

Response:

GHG Emissions and Calculations:
The requested 13 MMBtu of natural gas per short ton of HBI includes the startup and shutdown emissions (routed to the flare) and any off specification materials. Revised calculations are included in Attachment 2 and will be the basis for determining CO_2 emissions from the facility. The calculations have a slight conservancy on a stack emissions basis as the emissions from the process water degasser and the flare are considered outside of the 13 MMBtu/ton even though mass balance requires that these emissions reduce the net emissions from the reformer stack.

The proposed BACT limit is based on previous and recent BACT determinations for a similar source. As discussed in the Response to Question 5, the technologies are not different in chemistry or stoichiometry; therefore, the rationale is the same.

Process Indicators:
voestalpine will propose that the main line natural gas will be monitored as this will indicate the entire carbon flow into the plant, save for the two emergency generators which can be monitored independently. Further monitoring will be on the production of direct reduced iron including off specification material.

7. On page 28 of the permit application, it is stated that one means of reducing natural gas consumption is to remove the oxygen that is being freed from the iron oxide ore from the process gas loop of the reducing gas. This oxygen, in the form of CO_2 and water vapor, inhibits the reaction of CO and H_2 with the oxygen of the ore when either or both are present at high levels. While some CO_2 and water vapor are necessary in the reactions of the reformer, the removal of excess CO_2 and water vapor in the system will improve overall efficiency. Current DRI process designs release CO_2 and water vapor from the process gas loop by off-taking a stream of spent reducing gas (prior to recycle back to the reformer) and using this stream as fuel in the reformer. Is there a difference in the amount of CO_2 produced in the furnaces of the two DR technologies (reformer-based vs. reformer-less) which results in more or less excess CO_2 that needs to be removed via some type of process gas recycle loop that ultimately improves the performance of the furnace operation?

Response:

Explanation regarding difference reformer-based vs. reformer-less technology with regards to CO_2:
There is no difference between the two processes. In the MIDREX case, the reformer off gas is the final sink for CO_2, in case of HYL the same amount of CO_2 comes out of the MEA and the heater off gas.

Comparison gas consumption/CO_2-emissions MIDREX/HYL:

- Goal of both processes is to reduce iron-oxide to iron:
  - $FeO_n + nCO(H_2) = Fe + nCO_2(H_2O)$
- that means for the reduction process demand of natural gas for producing the reducing gas (CO_2H_2) is the same (MIDREX, HYL)
- there is only a difference in the way of producing the reducing gas
  - MIDREX:
    Reducing gas production via external Reformer (endothermic process)
    $CH_4 + CO_2 = 2CO + 2H_2$
  - HYL-ZR:
    Partial oxidation of natural gas and Reforming in the reduction shaft furnace
    $CH_4 + 1/2O_2 = CO + 2H_2$
- From this point of view it looks like MIDREX could produce a higher rate of reducing gas than HYL-ZR (different mol-ratio)
  - MIDREX : 1 mol CH4 → 4 mol reducing gas
  - HYL-ZR : 1 mol CH4 → 3 mol reducing gas
- Due to the endothermic Reforming process MIDREX needs more energy to produce reducing gas → at the end the specific gas consumption for both processes are the same and the resultant CO_2-emissions.

Demand of natural gas:
In a first view the gas consumption during pre-basic-design of the plant configuration (this was at the time when voestalpine developed the application documents for AIR- and GHG-permits) was estimated to 380 m^3 per ton of product HBI (including all process conditions like idle condition, start up, shut down etc.; different product quality parameters). In the meantime the

level of detail engineering for the current plant increased in a way that we can assume a lower gas consumption compared to primarily assumptions. Please see included in Attachment 3 informational brochure regarding the two technologies. This information is not guaranteed by the vendors.

Decision criterion:

- Following criteria has to be considered for technology decision
  - Plantspecific details for design
  - Plant-size-dependent operating results on the strength of past experience
  - Quality-dependent operating results on the strength of past experience
- Based on the assessment for the 2 process technologies which have been done the MIDREX-technology have been selected as "ideally suited" to be in compliance with quality-requirements for the intended high quality HBI in a reliable manner.

8. On page 32 of the permit application, Table 4-6 presents approximate costs for construction and operation of a post-combustion carbon capture and sequestration system at voestalpine. The estimated cost to install, operate and maintain CCS is $142.3 million per year at the voestalpine facility at $87 per tonne of CO_2 controlled. The supporting calculations that were used to derive this estimate were not included in the application. Please provide the site-specific parameters that were used to evaluate and eliminate CCS from consideration. This material should contain detailed information on the quantity and concentration of CO_2 that is in the waste stream and the specific equipment to be used. This site-specific cost calculations should include, but are not limited to, size and distance of pipeline to be installed, pumps, compressors, the amine solution to be used and the equipment necessary to employ the chosen post-combustion technology. Please include cost of construction, operation and maintenance, cost per ton of CO_2 removed by the technologies evaluated and include the feasibility and cost analysis for storage or transportation for these options. Please discuss in detail any site specific safety or environmental impacts associated with such a removal system. The heading in Table 4-6 indicates that cost numbers are “$/ton of CO_2 controlled”; however at the bottom of page 32, a cost of “$87 per tonne of CO_2 controlled” is given. Please ensure that all calculations, including emission calculations, are done in short tons.

   Response:

   The BACT Analysis has been updated to include carbon capture and storage costs to add to the costs presented in the original BACT analysis. The BACT will be submitted under separate cover.

9. On page 37 of the permit application, the proposed BACT for the reformer main flue ejector stack is the energy integration through the combustion of spent reducing gas, combined with natural gas combustion for supplemental energy needs. In addition, beginning on page 35 of the permit, it is stated that natural gas is only needed to supplement approximately 40% of the total energy input to the reformer. Please provide supplemental supporting design calculations that were used to derive this percentage, including technical resources and literature. What is the proposed compliance strategy for this proposed BACT? What are the proposed monitoring and recordkeeping requirements to ensure compliance?

Response:

Supplemental Heat:
Based on information provided by Midrex, 247.3 MMBtu per hour will be used in the reformer to provide additional heat. The total heat rate is estimated at 1,591 MMBtu/hr into the reformer. This equates to 15.5% supplemental rate. Please see the emissions calculations included in Attachment 2.

The proposed compliance strategy with regards to CO_2 emissions would be a combination of natural gas monitoring and production monitoring (quality and off-speculation). Natural gas will be continuously monitoring at the primary connection to the provider pipeline and will be the basis for estimating CO_2 emissions. Production monitoring will occur continuously after the briquetter machines.

10. voestalpine did not propose to implement a fugitive emission monitoring program for piping components. Please provide supplemental data to the 5-step BACT analysis for fugitives that include a comprehensive evaluation of the technologies considered to reduce fugitive emissions and a basis for elimination, or information detailing why fugitive emissions will not be emitted from this project. The technologies could include, but are not limited to, the following:
    - Installing leakless technology components to eliminate fugitive emission sources;
    - Implementing an alternative monitoring program using a remote sensing technology such as infrared camera monitoring;
    - Designing and constructing facilities with high quality components and materials of construction compatible with the process known as the Enhanced LDAR standards;
    - Monitoring of flanges for leaks;
    - Using a lower leak detection level for components; and
    - Implementing an audio/visual/olfactory (AVO) monitoring program for compounds.

    The BACT analysis should include for the proposed monitoring program a compliance strategy. (i.e., frequencies of inspections, maintenance repair strategy, recordkeeping, etc.).

Response:

Fugitive emissions:
Fugitive emissions from the system are not expected from the reformer and reactor. The design of the system is that the system operates at a slight negative pressure allowing for the effective use of a seal and purge gas system. This system provides a net inflow into the reformer loop. However, the natural gas delivery system may have fugitive emissions; consequently, an emissions estimate and BACT analysis will be added to the GHG BACT analysis. The BACT analysis will be provided under separate cover.

Fugitive loss data for mentioned measures are dependent on available equipment on the market. The equipment purchasing process (and requests for vendor bid on equipment) will be implemented in a way that the suppliers will be bound by contracts to be in compliance with the relevant performance requirements.

11. On page 41 of the permit application, the proposed BACT for the flare is good combustion practices, proper maintenance and use of clean fuel, installation of a natural gas flare tip, periodic fuel sampling and analysis where composition could vary. Please provide supplemental data that discusses the design of the flare, i.e., percent combustion efficiency, specific monitoring and recordkeeping strategy, maintenance schedule, etc. What will be frequency of the sampling of fuel to determine quality? Will it be computer controlled? If so, will there be manual overrides? If possible, please provide benchmark comparison data of the new flare system to similar or existing flares in the DRI industry. What will the visible emissions monitoring entail? How often will visible emissions observations be conducted?

Response:

Flare Data:
The following codes and standards shall apply in all stages of design, manufacture, inspection and testing. The Vendor shall indicate in his proposal the detailed codes and standards which will be followed. All codes and standards shall be latest editions.

Materials	ASTM, ASME, AISI, API 537
Noise and radiation	OSHA, API 521, API 537
Pipe connections	ASME B31.3
Piping (General)	ASME B31.3
Safety	OSHA
Welding Qualifications	ASME Section IX, latest Ed.
Welding (General)	AWS
Electrical & Instrumentation	NEMA, NEC, ISA

Technical Process Data:

Parameter	Value	Units
Destruction, nominal	98	% against CO, CH4
Flow rate (Qn)	~13	M^3 / hr (Pilot natural gas)
	~30,000	M^3 / hr (Shutdown)
	Max: 150,000	M^3 / hr (~30 seconds)
Pressure, nominal	0.091	Mpa(g)
Pressure, maximum	0.125	Mpa(g)
Molecular weight	18.68	g/g-mole
Specific gravity	0.645	

Visible Emissions:
Since the flare is combusting hydrogen, carbon monoxide and methane, visible emissions above 20% opacity is not expected. voestalpine will conduct visible observations as per TCEQ requirements as part of 40 CFR 64 compliance aspects that will be incorporated into voestalpine's TCEQ PSD Permit.

12. On page 42 of the permit application, the BACT analysis for the seal gas vent states that in order to prevent the reducing gas from escaping the furnace, a higher pressure gas called seal gas is applied at both the charging and discharging opening. The seal gas is allowed to escape the furnace while the reducing gas is retained. Due to the higher seal gas pressure, a portion is also entrained into the reactor and combined with the spent reducing gas which travels back to the reformer. This seal gas is a small amount of cooled flue gas from the reformer combustion side, and primarily consists of atmospheric nitrogen, CO_2 and water vapor. Where is this seal gas allowed to escape? Is it re-captured back into the seal gas system or does it emit to atmosphere? If so, has this amount of CO_2 that is allowed to escape accounted for in the emission calculations? Is this seal system common to both DRI technologies?

Response:

Seal gas release;
The seal gas is released in the following manner. A part of the upper seal gas goes via the upper seal leg through the charging hopper to the atmosphere. A part of the lower seal gas goes via the PDC and the BSG system to atmosphere (at the GCS Scrubber and Briquetter scrubber). A detailed description of the seal gas system is provided in Response 2.B. The free flow of material without lockhoppers is a unique characteristic of the MIDREX process to achieve homogenous and stable material flow from the charge hopper to the product discharge chamber and therefore important for operating efficiency. Stable process conditions minimize the use of natural gas and are therefore a benefit to minimize GHG emissions

13. The emission calculations for the emergency generator located in Appendix A doesn't appear to include CO_2e emissions. Please explain the omission.

Response:

Emissions for the emergency generator:
CO_2 emission estimates have been added to the emergency generator calculations. A revised emissions estimate is included Attachment 2. A summary of the emissions were included in the BACT based on 40 CFR 60 compliant machines.

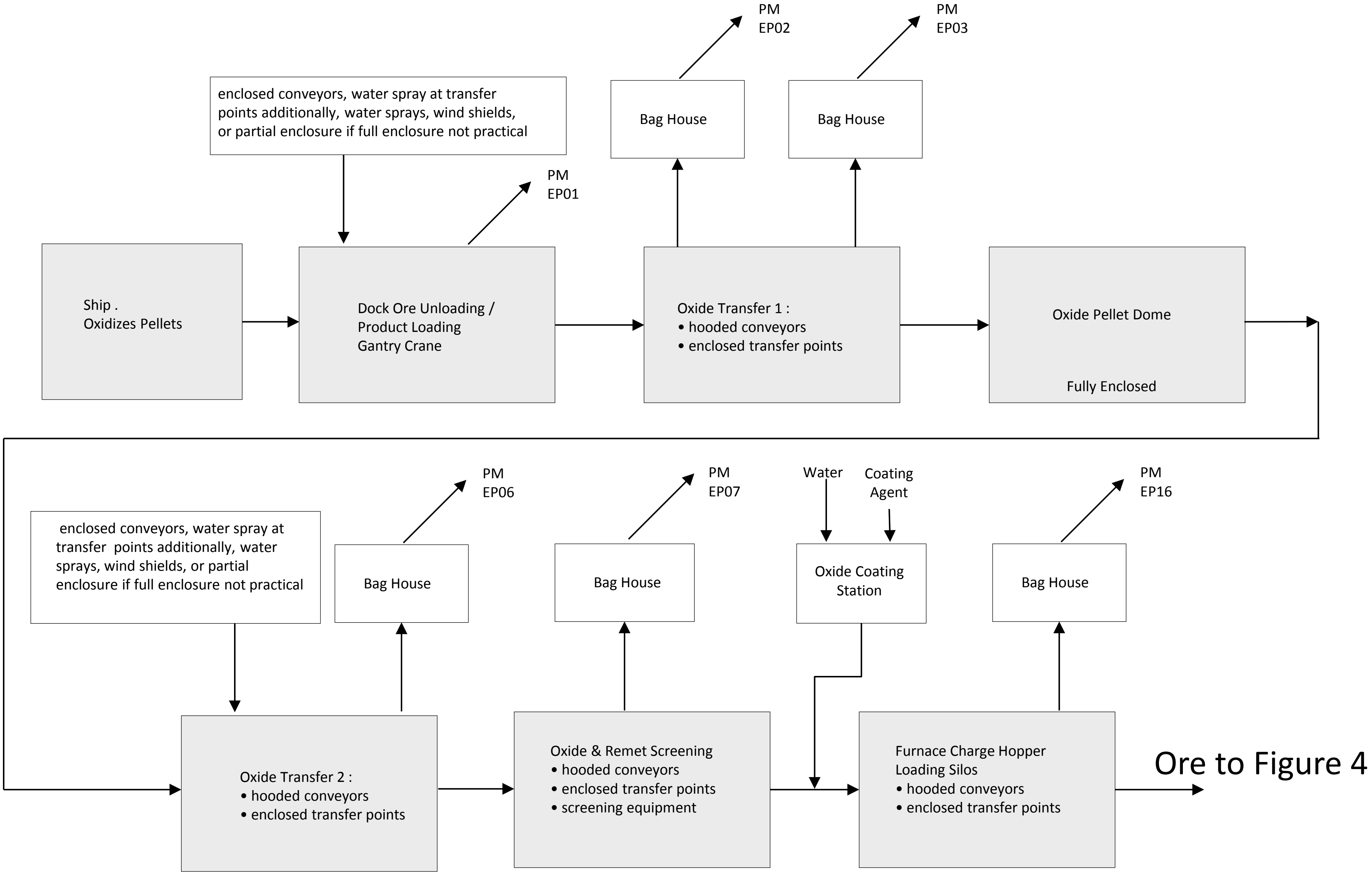
14. On page 24 of the permit application, the proposed GHG annual emission rates for the voestalpine is 1,814,144 tons/year. However, the total annual CO_2e emissions on page 4 of 9 of the Form PI-1 are given as 1,811,862 tons per year. Please provide an emission summary table of the emission sources, associated EPNs, the emissions for each source and the total annual emissions proposed for the voestalpine project.

Response:

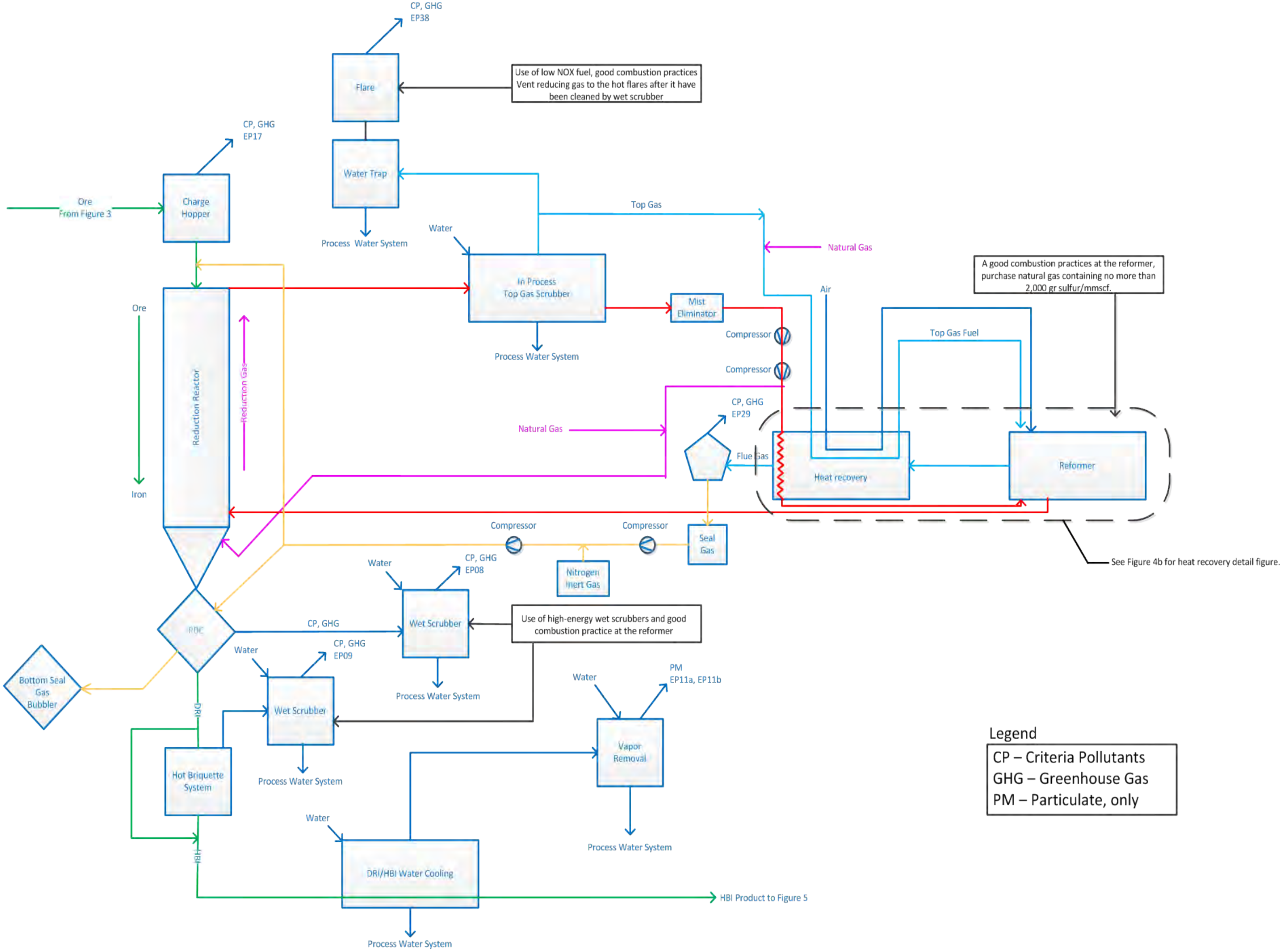
CO_2 Emissions Estimates:
Revised calculations are presented in Attachment 2.

Attachment 1

Revised Process Flow Diagrams



Process Flow Diagram Ore Handling / Figure 3



Use of low NOX fuel, good combustion practices
Vent reducing gas to the hot flares after it have
been cleaned by wet scrubber

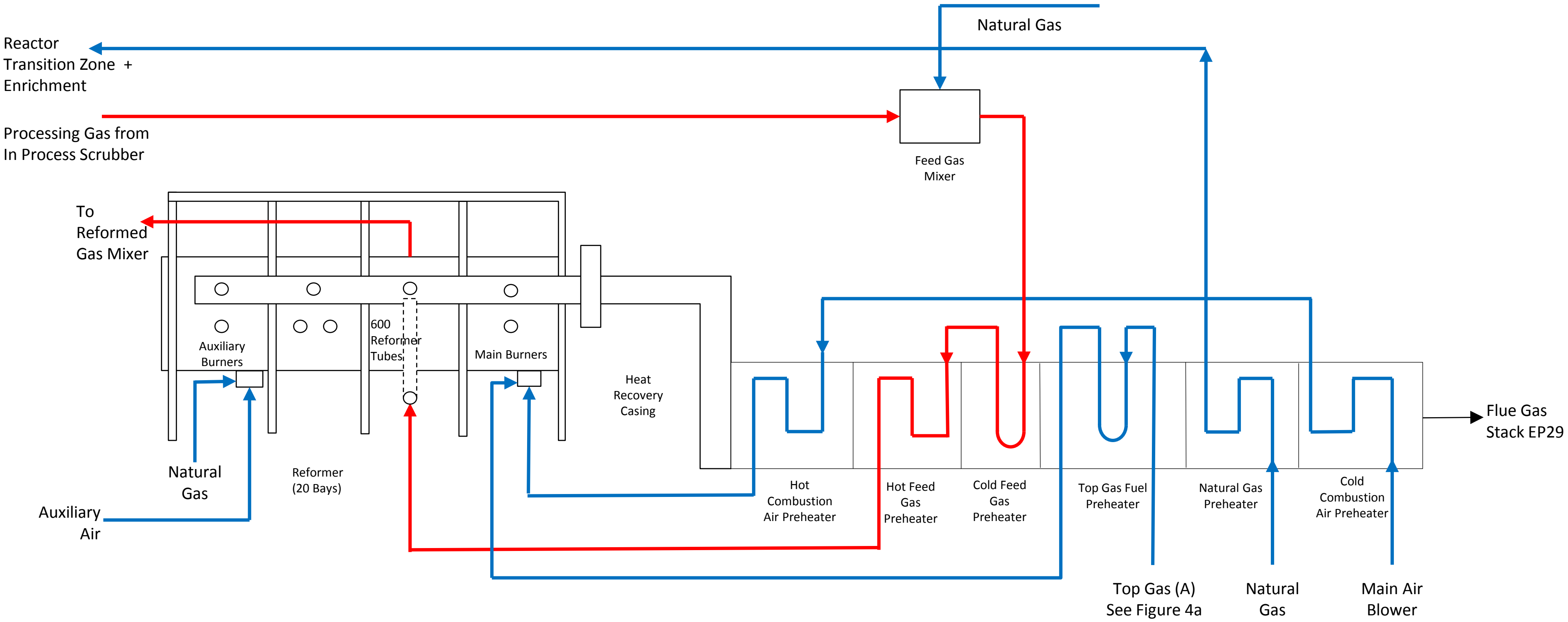
A good combustion practices at the reformer,
purchase natural gas containing no more than
2,000 gr sulfur/mmscf.

See Figure 4b for heat recovery detail figure.

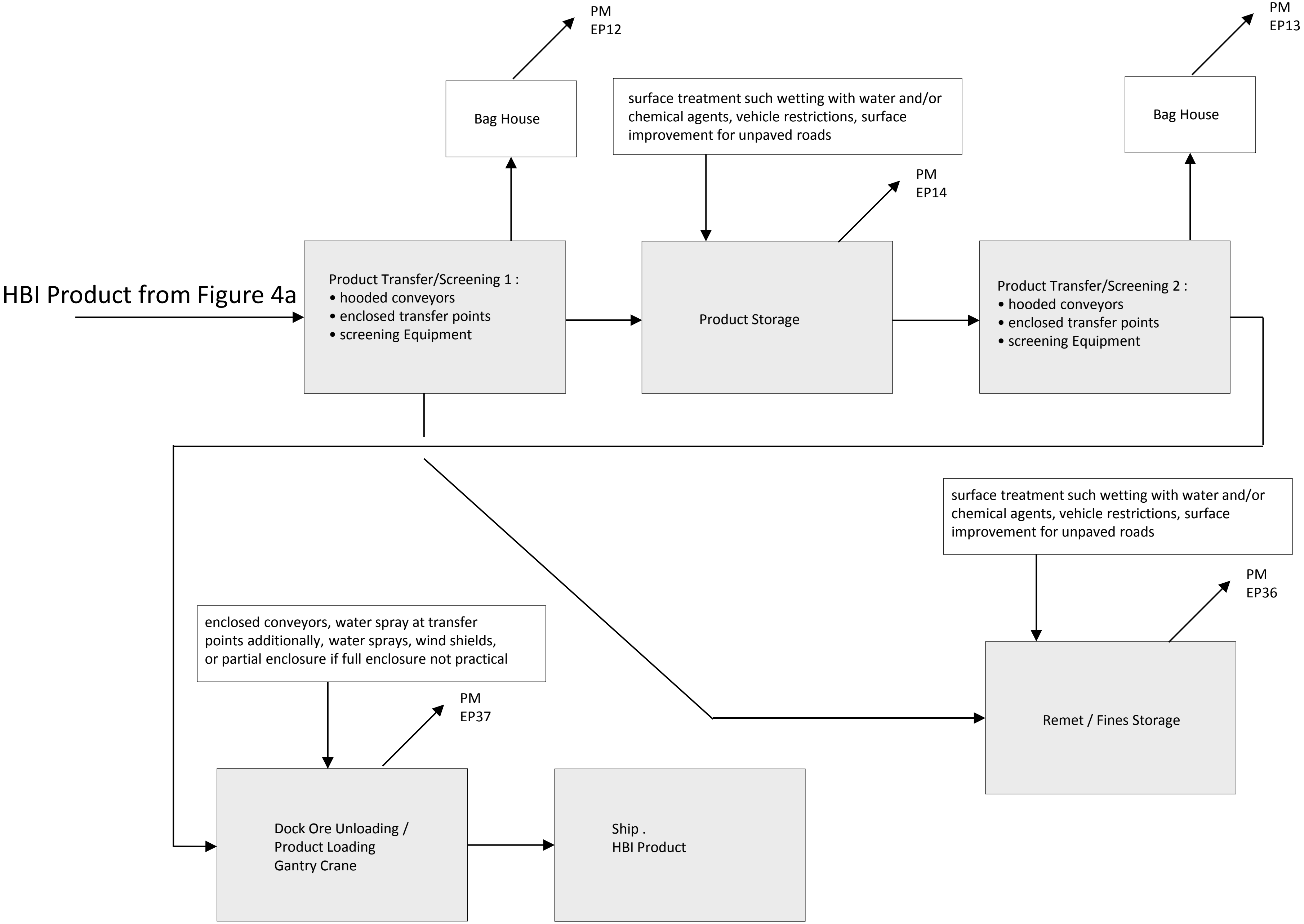
Use of high-energy wet scrubbers and good
combustion practice at the reformer

Legend
CP – Criteria Pollutants
GHG – Greenhouse Gas
PM – Particulate, only

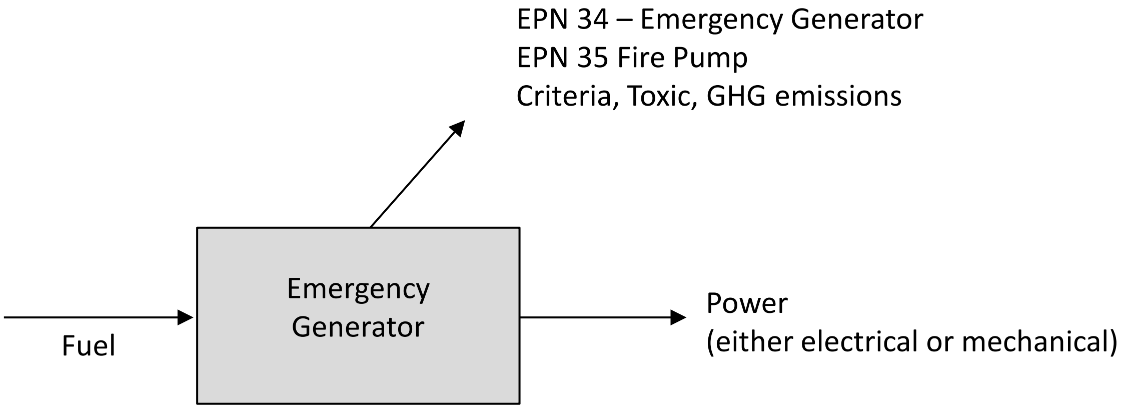
Process Flow Diagram Reduction, Reformer, Hot Briquetting Figure 4a



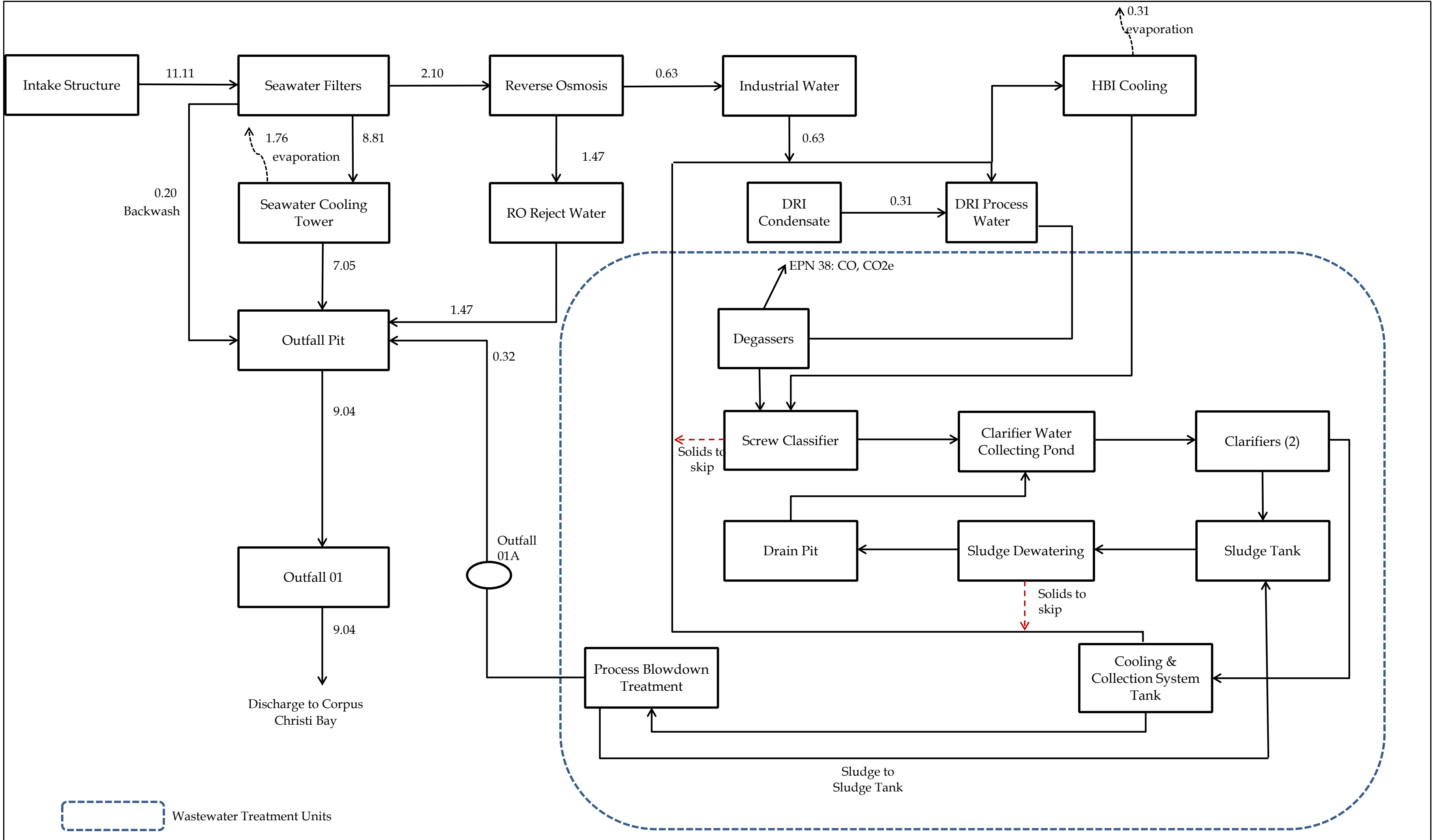
Process Gas & Reformer to Heat Recovery Figure 4b



Process Flow Diagram Product Handling Figure 5



Process Flow Diagram Figure 6



* values represent water flow in MGD

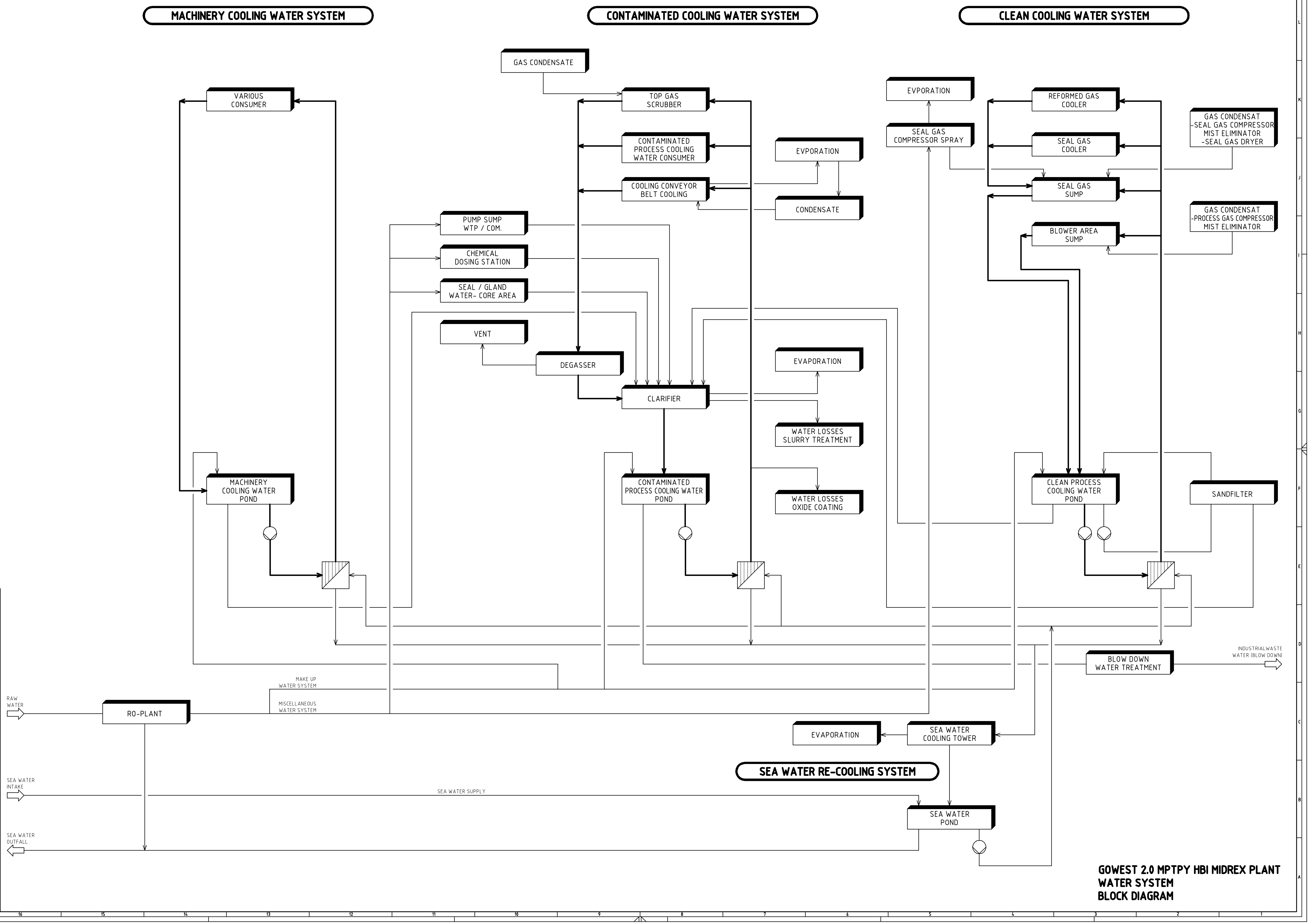
Environmental Resources Management

DESIGN: K. MARSH	DRAWN: K. MARSH	CHKD.: P. FLAHERTY
DATE: 8/8/2013	SCALE: N/A	REVISION: 3

File: P:\voestalpine\Water Permitting\Permit Application

ATTACHMENT 6
PROCESS FLOW DIAGRAM AND WATER BALANCE
LaQuinta Plant - TPDES Permit Application
voestalpine Texas LLC
San Patricio and Nueces Counties, TX

ERM



GOWEST 2.0 MPTPY HBI MIDREX PLANT
WATER SYSTEM
BLOCK DIAGRAM

Attachment 2

Revised Emission Calculations

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

			TSP Emissions		PM_{10} Emissions		$PM_{2.5}$ Emissions		NO_x Emissions	
Emission Point ID	**Modeling Stack No.**	**Air Emission Source Description**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**
1	1	Dock Ore Unloading / Product Loading Gantry Crane	0.28	1.22	0.13	0.58	0.02	0.09		
4	4	Oxide Unloading Bin & Dedusting	0.11	0.49	0.11	0.49	0.08	0.36		
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)	0.25	1.10	0.25	1.10	0.19	0.83		
6	6	Oxide Transfer & Dedusting (Post Domes)	0.11	0.49	0.11	0.49	0.08	0.36		
7	7	Oxide & Remet Screening & Dedusting	0.71	3.09	0.71	3.09	0.53	2.32		
16	16	Furnace Charge Hopper Loading Silos	0.01	<0.01	0.01	<0.01	0.01	<0.01		
17	17	Charge Hopper	0.10	0.44	0.10	0.44	0.08	0.33	2.02	8.84
29	29	Reformer Main Flue Ejector Stack	8.06	35.30	8.06	35.30	8.06	35.30	88.64	388.25
8	8	Furnace Dedusting (BSG Dust Collection)	2.50	10.95	2.50	10.95	1.25	5.48	2.02	8.84
9	9	Briquetter Dedusting	3.97	17.38	3.97	17.38	1.98	8.69	1.01	4.42
38	38	Hot Pressure Relief Vent (Flare)	0.51	0.08	0.51	0.08	0.51	0.08	20.26	2.96
11a	11a	HBI Cooling Conveyor 1	1.90	8.34	1.90	8.34	0.95	4.17		
11b	11b	HBI Cooling Conveyor 2	1.90	8.34	1.90	8.34	0.95	4.17		
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)	0.40	1.77	0.40	1.77	0.36	1.59		
13	13	Transfer & Product Screening Station No. 2 (Post Pile)	0.40	1.77	0.40	1.77	0.36	1.59		
14	14	HBI Pile	0.32	1.39	0.13	0.56	0.04	0.16		
36	36	Remet / Fines Storage	0.17	0.74	0.04	0.20	0.01	0.02		
30	30	Process Water Degasser								
33	33	Salt Water Cooling Tower	2.61	11.44	0.08	0.34	0.08	0.34		
39	39	Paved Road Fugitive Dust	0.25	1.08	0.05	0.21	0.01	0.03		
34	34	Emergency Generator	0.47	0.02	0.47	0.02	0.47	0.02	32.09	1.60
35	35	Fire Pump	0.03	<0.01	0.03	<0.01	0.03	<0.01	2.25	0.11
FugGHG	-	Fugitive Natural Gas								
		TOTAL EMISSIONS	**25.07**	**105.42**	**21.88**	**91.44**	**16.05**	**65.94**	**148.28**	**415.03**

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

			SO_2 Emissions		CO Emissions		VOC Emissions		CO_2 Emissions		CO_2e Emissions	
Emission Point ID	**Modeling Stack No.**	**Air Emission Source Description**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**
1	1	Dock Ore Unloading / Product Loading Gantry Crane										
4	4	Oxide Unloading Bin & Dedusting										
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)										
6	6	Oxide Transfer & Dedusting (Post Domes)										
7	7	Oxide & Remet Screening & Dedusting										
16	16	Furnace Charge Hopper Loading Silos										
17	17	Charge Hopper	0.24	1.05	2.82	12.36	0.18	0.81	12,486	54,689	12,489	54,702
29	29	Reformer Main Flue Ejector Stack	10.51	32.20	123.97	542.98	8.12	35.55	548,516	1,679,829	548,639	1,680,207
8	8	Furnace Dedusting (BSG Dust Collection)	0.24	1.05	2.82	12.36	0.18	0.81	12,486	54,689	12,489	54,702
9	9	Briquetter Dedusting	0.12	0.52	1.41	6.18	0.09	0.40	6,243	27,345	6,244	27,351
38	38	Hot Pressure Relief Vent (Flare)	0.02	<0.01	239.36	31.76	0.04	0.02	15,319	2,235	15,319	2,235
11a	11a	HBI Cooling Conveyor 1										
11b	11b	HBI Cooling Conveyor 2										
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)										
13	13	Transfer & Product Screening Station No. 2 (Post Pile)										
14	14	HBI Pile										
36	36	Remet / Fines Storage										
30	30	Process Water Degasser			24.26	106.24			252.25	1,104.86	257	1,126
33	33	Salt Water Cooling Tower										
39	39	Paved Road Fugitive Dust										
34	34	Emergency Generator	0.04	<0.01	3.80	0.19	0.99	0.05	3,930.81	196.54	3,931	197
35	35	Fire Pump	<0.01	<0.01	0.27	0.01	0.07	<0.01	256.61	12.83	257	13
FugGHG	-	Fugitive Natural Gas									19	84
		TOTAL EMISSIONS	**11.18**	**34.82**	**398.70**	**712.08**	**9.68**	**37.64**	**599,490**	**1,820,102**	**599,645**	**1,820,616**

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

			Ammonia Emissions		Benzene		Dichlorobenzene		Formaldehyde		n-Hexane	
Emission Point ID	**Modeling Stack No.**	**Air Emission Source Description**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**
1	1	Dock Ore Unloading / Product Loading Gantry Crane										
4	4	Oxide Unloading Bin & Dedusting										
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)										
6	6	Oxide Transfer & Dedusting (Post Domes)										
7	7	Oxide & Remet Screening & Dedusting										
16	16	Furnace Charge Hopper Loading Silos										
17	17	Charge Hopper			<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.04
29	29	Reformer Main Flue Ejector Stack			<0.01	<0.01	<0.01	<0.01	0.02	0.08	0.41	1.81
8	8	Furnace Dedusting (BSG Dust Collection)			<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.04
9	9	Briquetter Dedusting			<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.02
38	38	Hot Pressure Relief Vent (Flare)										
11a	11a	HBI Cooling Conveyor 1										
11b	11b	HBI Cooling Conveyor 2										
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)										
13	13	Transfer & Product Screening Station No. 2 (Post Pile)										
14	14	HBI Pile										
36	36	Remet / Fines Storage										
30	30	Process Water Degasser										
33	33	Salt Water Cooling Tower										
39	39	Paved Road Fugitive Dust										
34	34	Emergency Generator										
35	35	Fire Pump										
FugGHG	-	Fugitive Natural Gas										
		TOTAL EMISSIONS	**0.00**	**0.00**	**<0.01**	**<0.01**	**<0.01**	**<0.01**	**0.02**	**0.08**	**0.44**	**1.91**

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

			Naphthalene		Toluene		PAH		Lead		Mercury	
Emission Point ID	**Modeling Stack No.**	**Air Emission Source Description**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**
1	1	Dock Ore Unloading / Product Loading Gantry Crane							<0.01	<0.01	<0.01	<0.01
4	4	Oxide Unloading Bin & Dedusting							<0.01	<0.01	<0.01	<0.01
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)							<0.01	<0.01	<0.01	<0.01
6	6	Oxide Transfer & Dedusting (Post Domes)							<0.01	<0.01	<0.01	<0.01
7	7	Oxide & Remet Screening & Dedusting							<0.01	<0.01	<0.01	<0.01
16	16	Furnace Charge Hopper Loading Silos										
17	17	Charge Hopper	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
29	29	Reformer Main Flue Ejector Stack	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01		
8	8	Furnace Dedusting (BSG Dust Collection)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
9	9	Briquetter Dedusting	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
38	38	Hot Pressure Relief Vent (Flare)							<0.01	<0.01		
11a	11a	HBI Cooling Conveyor 1							<0.01	<0.01	<0.01	<0.01
11b	11b	HBI Cooling Conveyor 2							<0.01	<0.01	<0.01	<0.01
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)							<0.01	<0.01	<0.01	<0.01
13	13	Transfer & Product Screening Station No. 2 (Post Pile)							<0.01	<0.01	<0.01	<0.01
14	14	HBI Pile							<0.01	<0.01	<0.01	<0.01
36	36	Remet / Fines Storage							<0.01	<0.01	<0.01	<0.01
30	30	Process Water Degasser										
33	33	Salt Water Cooling Tower										
39	39	Paved Road Fugitive Dust										
34	34	Emergency Generator										
35	35	Fire Pump										
FugGHG	-	Fugitive Natural Gas										
		TOTAL EMISSIONS	**<0.01**	**<0.01**	**<0.01**	**<0.01**	**<0.01**	**<0.01**	**<0.01**	**0.01**	**<0.01**	**<0.01**

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

Emission Point ID	Modeling Stack No.	Air Emission Source Description	Cadmium		Chromium		Magnesium (as MgO)		Manganese	
			Average lb/hr	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**	**Average lb/hr**	**Average tons/yr**
1	1	Dock Ore Unloading / Product Loading Gantry Crane	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
4	4	Oxide Unloading Bin & Dedusting	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
6	6	Oxide Transfer & Dedusting (Post Domes)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
7	7	Oxide & Remet Screening & Dedusting	<0.01	<0.01	<0.01	<0.01	<0.01	0.02	<0.01	<0.01
16	16	Furnace Charge Hopper Loading Silos								
17	17	Charge Hopper	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
29	29	Reformer Main Flue Ejector Stack								
8	8	Furnace Dedusting (BSG Dust Collection)	<0.01	<0.01	<0.01	<0.01	0.02	0.07	<0.01	0.01
9	9	Briquetter Dedusting	<0.01	<0.01	<0.01	<0.01	0.02	0.10	<0.01	0.01
38	38	Hot Pressure Relief Vent (Flare)								
11a	11a	HBI Cooling Conveyor 1	<0.01	<0.01	<0.01	<0.01	0.01	0.05	<0.01	<0.01
11b	11b	HBI Cooling Conveyor 2	<0.01	<0.01	<0.01	<0.01	0.01	0.05	<0.01	<0.01
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
13	13	Transfer & Product Screening Station No. 2 (Post Pile)	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
14	14	HBI Pile	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
36	36	Remet / Fines Storage	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
30	30	Process Water Degasser								
33	33	Salt Water Cooling Tower								
39	39	Paved Road Fugitive Dust								
34	34	Emergency Generator								
35	35	Fire Pump								
FugGHG	-	Fugitive Natural Gas								
TOTAL EMISSIONS			**<0.01**	**<0.01**	**<0.01**	**<0.01**	**0.08**	**0.35**	**0.01**	**0.03**

SUMMARY TABLE: voestalpine Direct Reduction Iron Facility Emissions Totals

			Nickel		Copper		Zinc		Total HAPs	
Emission Point ID	Modeling Stack No.	Air Emission Source Description	Average lb/hr	Average tons/yr	Average lb/hr	Average tons/yr	Average lb/hr	Average tons/yr	Average lb/hr	Average tons/yr
1	1	Dock Ore Unloading / Product Loading Gantry Crane	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
4	4	Oxide Unloading Bin & Dedusting	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
5	5	Oxide Pellet Pile Transfer & Dedusting (Pre Domes)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
6	6	Oxide Transfer & Dedusting (Post Domes)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
7	7	Oxide & Remet Screening & Dedusting	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
16	16	Furnace Charge Hopper Loading Silos							0.00	0.00
17	17	Charge Hopper	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.04
29	29	Reformer Main Flue Ejector Stack							0.43	1.90
8	8	Furnace Dedusting (BSG Dust Collection)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.05
9	9	Briquetter Dedusting	<0.01	0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.03
38	38	Hot Pressure Relief Vent (Flare)							0.00	0.00
11a	11a	HBI Cooling Conveyor 1	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.01
11b	11b	HBI Cooling Conveyor 2	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.01
12	12	Transfer & Product Screening Station No. 1 (Pre Pile)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
13	13	Transfer & Product Screening Station No. 2 (Post Pile)	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
14	14	HBI Pile	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
36	36	Remet / Fines Storage	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.00	0.00
30	30	Process Water Degasser							0.00	0.00
33	33	Salt Water Cooling Tower							0.00	0.00
39	39	Paved Road Fugitive Dust							0.00	0.00
34	34	Emergency Generator							0.00	0.00
35	35	Fire Pump							0.00	0.00
FugGHG	-	Fugitive Natural Gas							0.00	0.00
		TOTAL EMISSIONS	**<0.01**	**0.02**	**<0.01**	**0.01**	**<0.01**	**0.01**	**0.46**	**2.05**

Emission Point Source Calculations

Case: **1**
Description: **Dock Ore Unloading / Product Loading Gantry Crane**

Inputs	Value	Units
Annual hours of operation	8,760	hrs/yr
Maximum Loading Rate (Conveyor Capacity)	1,500	tons/hr
Average wind speed	12.0	mph
Loading control efficiency	90.0%	%
Unloading control efficiency	90.0%	%
Annual throughput - iron ore	3,197,250	tons/yr
Excess Receiving Capacity (Annual)	110%	%
Conveyor System Service Factor (Hourly)	115%	%
Material moisture content-iron ore	3.20	%

Emissions Summary

Source Description	**Pollutant**	**Average Emission Rate**	**Annual Emission Rate**
		(lbs/hr)	**(tpy)**
Barge/Ship Unloading (Dock 1)	TSP*	0.279	1.22
	PM_{10}	0.132	0.579
	$PM_{2.5}$	0.0200	0.088
	Hg	2.79E-08	1.22E-07
	Cd	1.40E-07	6.12E-07
	Cr	1.40E-05	6.12E-05
	Mg (as MgO)	1.68E-03	7.34E-03
	Mn	1.48E-04	6.49E-04
	Pb	1.40E-05	6.12E-05
	Ni	8.38E-05	3.67E-04
	Cu	2.79E-05	1.22E-04
	Zn	5.59E-05	2.45E-04

* Per AP-42 (Table 13.2.2-2), PM-30 is assumed equivalent to total suspended particle matter (TSP)

Calculation Basis:

Loading emissions: (1) conveyor to ship/barge (i.e., 1 drop)
Unloading emissions: (1) ship/barge via gantry grab crane to hopper, (2) hopper to receiving conveyor (2 drops)

Emission Factor (lb PM / ton handled)*

$$Emission\ factor(lb/ton) = k*(0.0032)*\frac{\left(\frac{U}{5}\right)^{1.3}}{\left(\frac{M}{2}\right)^{1.4}}$$

where U = Mean wind speed (mph) for the area [1]
M = Material moisture content (%)[2]
k = Particle size multiplier [3]

Aerodynamic Particle Size Multiplier (k)	
< 30 micrometers	**0.74**
< 15 micrometers	0.48
< 10 micrometers	**0.35**
< 5 micrometers	0.2
< 2.5 micrometers	**0.053**

Note:
1. Corpus Christi average wind speed (previous 60 years) http://www.ncdc.noaa.gov/oa/climate/online/ccd/avgwind.html
2. From AP-42 (11/06), Section 13.2.4.2, Table 13.2.4-1 and process experience
3. From AP-42 (11/06), Section 13.2.4.3

* Calculation methodology derived from Section 13.2.4 "*Aggregate Handling and Storage Piles*," AP-42 (11/06).

Emission Calculations

Average Controlled Emission Rate (tons/yr) = Material Throughput (tons/yr) * EF (lb PM/ton) * (Number of drops) * (1-Control efficiency)
Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Particulate Emissions from Barge Loading/Unloading @ Dock # 1

Pollutant	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Excess Receiving Capacity	Conveyor Service Factor	Average Emission Rate	Annual Emission Rate
	(tons/yr[1])	(lb/ton)		(lb/yr)	%	%	%	(lb/hr$_{avg}$)	(tpy)
TSP	3,197,250	0.00383	2.00	24,473	90.0%	110%	115%	0.279	1.22
PM10	3,197,250	0.00181	2.00	11,575	90.0%	110%	115%	0.132	0.579
PM2.5	3,197,250	0.000274	2.00	1,753	90.0%	110%	115%	0.0200	0.088

[1] Capacity converted to US tons

[2] Pig iron has no intrinsic silt content, and is assumed to be non-emitting.

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr$_{avg}$)	(tpy)
Hg	0.1	2.79E-08	1.22E-07
Cd	0.5	1.40E-07	6.12E-07
Cr	50	1.40E-05	6.12E-05
Mg (as MgO)	6000	1.68E-03	7.34E-03
Mn	530	1.48E-04	6.49E-04
Pb	50	1.40E-05	6.12E-05
Ni	300	8.38E-05	3.67E-04
Cu	100	2.79E-05	1.22E-04
Zn	200	5.59E-05	2.45E-04

Emission Point Source Calculations

Emission Point Identifier: **4**
Description: **Oxide Unloading Bin & Dedusting**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Iron Oxide Consumption Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	11,000	Nm^3/hr
Clean Gas Concentration	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.111	0.485
PM_{10}	0.111	0.485
$PM_{2.5}$	0.0831	0.364
Hg	1.11E-08	4.85E-08
Cd	5.54E-08	2.43E-07
Cr	5.54E-06	2.43E-05
Mg (as MgO)	6.65E-04	2.91E-03
Mn	5.87E-05	2.57E-04
Pb	5.54E-06	2.43E-05
Ni	3.32E-05	1.46E-04
Cu	1.11E-05	4.85E-05
Zn	2.22E-05	9.71E-05

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	11,000	4.57	100%	0.111	0.485
PM_{10}	11,000	4.57	100%	0.111	0.485
$PM_{2.5}$	11,000	4.57	75.0%	0.0831	0.364

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	1.11E-08	4.85E-08
Cd	0.5	5.54E-08	2.43E-07
Cr	50	5.54E-06	2.43E-05
Mg (as MgO)	6000	6.65E-04	2.91E-03
Mn	530	5.87E-05	2.57E-04
Pb	50	5.54E-06	2.43E-05
Ni	300	3.32E-05	1.46E-04
Cu	100	1.11E-05	4.85E-05
Zn	200	2.22E-05	9.71E-05

Emission Point Source Calculations

Emission Point Identifier: 5
Description: Oxide Pellet Pile Transfer & Dedusting (Pre Domes)

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Iron Oxide Consumption Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	25,000	Nm^3/hr
Clean Gas Concentration	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.25	1.10
PM_{10}	0.25	1.10
$PM_{2.5}$	0.19	0.83
Hg	2.52E-08	1.10E-07
Cd	1.26E-07	5.52E-07
Cr	1.26E-05	5.52E-05
Mg (as MgO)	1.51E-03	6.62E-03
Mn	1.33E-04	5.85E-04
Pb	1.26E-05	5.52E-05
Ni	7.56E-05	3.31E-04
Cu	2.52E-05	1.10E-04
Zn	5.04E-05	2.21E-04

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	25,000	4.57	100%	0.25	1.10
PM_{10}	25,000	4.57	100%	0.25	1.10
$PM_{2.5}$	25,000	4.57	75.0%	0.19	0.83

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	2.52E-08	1.10E-07
Cd	0.5	1.26E-07	5.52E-07
Cr	50	1.26E-05	5.52E-05
Mg (as MgO)	6000	1.51E-03	6.62E-03
Mn	530	1.33E-04	5.85E-04
Pb	50	1.26E-05	5.52E-05
Ni	300	7.56E-05	3.31E-04
Cu	100	2.52E-05	1.10E-04
Zn	200	5.04E-05	2.21E-04

Emission Point Source Calculations

Emission Point Identifier: 6
Description: Oxide Transfer & Dedusting (Post Domes)

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Iron Oxide Consumption Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	11,000	Nm^3/hr
Clean Gas Concentration	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.111	0.485
PM_{10}	0.111	0.485
$PM_{2.5}$	0.0831	0.364
Hg	1.11E-08	4.85E-08
Cd	5.54E-08	2.43E-07
Cr	5.54E-06	2.43E-05
Mg (as MgO)	6.65E-04	2.91E-03
Mn	5.87E-05	2.57E-04
Pb	5.54E-06	2.43E-05
Ni	3.32E-05	1.46E-04
Cu	1.11E-05	4.85E-05
Zn	2.22E-05	9.71E-05

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	11,000	4.57	100%	0.111	0.485
PM_{10}	11,000	4.57	100%	0.111	0.485
$PM_{2.5}$	11,000	4.57	75.0%	0.0831	0.364

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	1.11E-08	4.85E-08
Cd	0.5	5.54E-08	2.43E-07
Cr	50	5.54E-06	2.43E-05
Mg (as MgO)	6000	6.65E-04	2.91E-03
Mn	530	5.87E-05	2.57E-04
Pb	50	5.54E-06	2.43E-05
Ni	300	3.32E-05	1.46E-04
Cu	100	1.11E-05	4.85E-05
Zn	200	2.22E-05	9.71E-05

Emission Point Source Calculations

Emission Point Identifier: **7**
Description: **Oxide & Remet Screening & Dedusting**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	70,000	Nm^3/hr
Clean Gas Concentration	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.705	3.09
PM_{10}	0.705	3.09
$PM_{2.5}$	0.529	2.32
Hg	7.05E-08	3.09E-07
Cd	3.53E-07	1.54E-06
Cr	3.53E-05	1.54E-04
Mg (as MgO)	4.23E-03	1.85E-02
Mn	3.74E-04	1.64E-03
Pb	3.53E-05	1.54E-04
Ni	2.12E-04	9.27E-04
Cu	7.05E-05	3.09E-04
Zn	1.41E-04	6.18E-04

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	70,000	4.57	100%	0.705	3.09
PM_{10}	70,000	4.57	100%	0.705	3.09
$PM_{2.5}$	70,000	4.57	75.0%	0.529	2.32

Metal Emissions (based on ore content)

Pollutant	Concentration (ppm)	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
Hg	0.1	7.05E-08	3.09E-07
Cd	0.5	3.53E-07	1.54E-06
Cr	50	3.53E-05	1.54E-04
Mg (as MgO)	6000	4.23E-03	1.85E-02
Mn	530	3.74E-04	1.64E-03
Pb	50	3.53E-05	1.54E-04
Ni	300	2.12E-04	9.27E-04
Cu	100	7.05E-05	3.09E-04
Zn	200	1.41E-04	6.18E-04

Emission Point Source Calculations

Emission Point Identifier: **16**
Description: **Furnace Charge Hopper Loading Silos**

Inputs

Description	Value	Units
Annual Hours of Operation	364	hr/yr
Designed Production Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	1,400	Nm^3/hr
Gas Dust Loading	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.0141	0.00257
PM_{10}	0.0141	0.00257
$PM_{2.5}$	0.0106	0.00193

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean gas Dust Loading	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	1,400	4.57	100%	0.0141	0.00257
PM_{10}	1,400	4.57	100%	0.0141	0.00257
$PM_{2.5}$	1,400	4.57	75.0%	0.0106	0.00193

This source consists of one silo. This source only emits while being filled.

Emission Point Source Calculations

Emission Point Identifier: **17**
Description: **Charge Hopper**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	3,197,250	tons/yr
Design Volmetric Vent Rate	10,000	Nm^3/hr
Gas Dust Loading	4.57	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction (uncontrolled 15%, AP-42)	75.0%	%

Emissions Summary

Pollutant	Calculation	Process Emission Rate (lb/hr)	Process Emission Rate (tpy)
TSP	Below	0.10	0.441
PM_{10}	Below	0.10	0.441
$PM_{2.5}$	Below	0.076	0.331
NOx	Seal Gas	2.02	8.8
SO2	Seal Gas	0.24	1.05
CO	Seal Gas	2.82	12.4
CO2	Seal Gas	12486	54689
CO2e	Seal Gas	12489	54702
VOC	Seal Gas	0.18	0.81
Lead (Pb)	Seal Gas	5.71E-05	2.50E-04
Benzene	Seal Gas	1.10E-05	4.80E-05
Dichlorobenzene	Seal Gas	6.27E-06	2.74E-05
Formaldehyde	Seal Gas	3.92E-04	1.72E-03
n-Hexane	Seal Gas	9.40E-03	4.12E-02
Naphthalene	Seal Gas	3.19E-06	1.40E-05
Toluene	Seal Gas	1.78E-05	7.78E-05
PAH	Seal Gas	4.39E-07	1.92E-06
Hg	Below	1.01E-08	4.41E-08
Cd	Below	5.04E-08	2.21E-07
Cr	Below	5.04E-06	2.21E-05
Mg (as MgO)	Below	6.05E-04	2.65E-03
Mn	Below	5.34E-05	2.34E-04
Ni	Below	3.02E-05	1.32E-04
Cu	Below	1.01E-05	4.41E-05
Zn	Below	2.02E-05	8.83E-05

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean gas Dust Loading	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	10,000	4.57	100%	0.10	0.441
PM_{10}	10,000	4.57	100%	0.10	0.441
$PM_{2.5}$	10,000	4.57	75.0%	0.076	0.331

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	1.01E-08	4.41E-08
Cd	0.5	5.04E-08	2.21E-07
Cr	50	5.04E-06	2.21E-05
Mg (as MgO)	6000	6.05E-04	2.65E-03
Mn	530	5.34E-05	2.34E-04
Pb	50	5.04E-06	2.21E-05
Ni	300	3.02E-05	1.32E-04
Cu	100	1.01E-05	4.41E-05
Zn	200	2.02E-05	8.83E-05

Emission Point Source Calculations

Emission Point Identifier: **29**
Description: **Reformer Main Flue Ejector Stack**

Inputs

Description	Value	Units	Comments
Annual Hours of Operation	8,760	hr/yr	
Designed Production Rate, Annual	2,205,000	tons/yr	Manfacturer
Designed Production Rate, Hourly	360	ton HBI / hr	Manfacturer
Maximum Production Rate Ratio	110%	%	Design Maximum
Total Reformer Vented Rate, wet @ 1.9% O2	464,300	Nm^3/hr	Design Maximum
Seal Gas System Off-take	25,000	Nm^3/hr	Manfacturer
Normal Reformer Firing Rate	1,591	MMBtu/hr	Main and Aux burners plus Top Gas
Nominal Natural Gas Supplement Rate	15.5%	%	Maufacturer
Total PM Emission Factor, dry @3%O2	10.0	mg/Nm^3	Vendor Guarantee, total
Filterable PM Emission Factor	25.0%	%	AP-42, Chapter 1, Section 4
NO_X Uncontrolled Concentration, dry at 3% O2	110.0	mg/Nm^3	Vendor Guarantee, total
PM_{10} Filterable Mass Fraction	100%		Worst case
$PM_{2.5}$ Filterable Mass Fraction	100%		Worst case
CO Emission Factor	84.0	lbs / $mmft^3$	AP-42, Chapter 1, Section 4, Table 1.4-1
CO2 Emission Factor	120,000	lbs / $mmft^3$	AP-42, Chapter 1, Section 4, Table 1.4-2
SO_2 Emission Factor	2.3	lbs / $mmft^3$	Manufacturer
VOC Emission Factor	5.50	lbs / $mmft^3$	AP-42, Chapter 1, Section 4, Table 1.4-2
Natural Gas, fuel and reduction gas raw materials	380	Nm^3/ton HBI	Manfacturer, total

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	8.06	35.3
PM_{10}	8.06	35.3
$PM_{2.5}$	8.06	35.3
NO_x	88.6	388
SO_2	10.51	32.2
CO	124	543
CO2	548,516	1,679,829
CO2e	548,639	1,680,207
VOC	8.12	35.6
Lead (Pb)	2.29E-03	7.00E-03
Benzene	4.82E-04	2.11E-03
Dichlorobenzene	2.75E-04	1.21E-03
Formaldehyde	0.0172	0.0754
n-Hexane	0.413	1.81
Naphthalene	1.40E-04	6.13E-04
Toluene	7.80E-04	3.42E-03
PAH	1.93E-05	8.44E-05

Calculate Emissions
Via Outlet Concentrations (Manufacutrer)

Pollutant	Normal Vented Flue Flow Rate	Dry Vented Flue Flow Rate	Component Concentration @ 3% O2	Control Efficiency	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(dNm^3/hr)	(mg/dNm^3)	%	(lb/hr)	(tons/yr)
NO_x	439,300	344,367	110.0	0%	88.6	388
TSP	439,300	344,367	10.0	0%	8.06	35.3
PM_{10}	439,300	344,367	10.0	0%	8.06	35.3
$PM_{2.5}$	439,300	344,367	10.0	0%	8.06	35.3

Emission rate correction to 3% O2: dNm^3 * mg/dNm^3 / 453600 * ((20.9-1.9)/(20.9-3))

Via AP-42 Emission Factor, Chapter 1, Section 4, Tables 1.4-1 and 1.4-2

Pollutant	Normal Reformer Firing Rate	Emission Factor	Control Efficiency	Seal Gas Compensation	Average Emission Rate	Annual Emission Rate
	(MMBtu/hr)	(lbs/mmBtu NG)	%	-	(lb/hr)	(tons/yr)
CO	1,591	0.0824	0%	0.946	124.0	543
VOC	1,591	0.00539	0%	0.946	8.12	35.6

Via Reducing Gas and Fuel Maximum

Pollutant	Total NG as Fuel and Raw Material	Max HBI output		Emission Factor	Control Efficiency	Seal Gas Compensation	Average Emission Rate	Annual Emission Rate
	(Nm^3/ton)	(ton HBI/hr)	(ton HBI/yr)	(lbs/mmcf NG)	%	-	(lb/hr)	(tons/yr)
SO2	380	360	2,205,000	2.30	0%	0.946	10.51	32.2
CO2	380	360	2,205,000	120,000	0%	0.946	548,516	1,679,829
CO2e	380	360	2,205,000	120,027	0%	0.946	548,639	1,680,207
Lead (Pb)	380	360	2,205,000	5.00E-04	0%	0.946	2.29E-03	0.0070

Calculate HAP Emissions

Pollutant	Normal Reformer Firing Rate	AP-42 Emission Factor	Nominal Natural Gas Supplement Rate	Seal Gas System Compensation	Average Emission Rate	Annual Emission Rate
	(MMBtu/hr)	(lb/MBtu)	%	(-)	(lb/hr)	(tons/yr)
Benzene	1,591	2.06E-06	15.5%	0.946	4.82E-04	2.11E-03
Dichlorobenzene	1,591	1.18E-06	15.5%	0.946	2.75E-04	1.21E-03
Formaldehyde	1,591	7.35E-05	15.5%	0.946	1.72E-02	7.54E-02
n-Hexane	1,591	1.76E-03	15.5%	0.946	4.13E-01	1.81E+00
Naphthalene	1,591	5.98E-07	15.5%	0.946	1.40E-04	6.13E-04
Toluene	1,591	3.33E-06	15.5%	0.946	7.80E-04	3.42E-03
PAH	1,591	8.24E-08	15.5%	0.946	1.93E-05	8.44E-05
Arsenic	1,591	1.96E-07	15.5%	0.946	4.59E-05	2.01E-04
Barium	1,591	4.31E-06	15.5%	0.946	1.01E-03	4.42E-03
Beryllium	1,591	1.18E-08	15.5%	0.946	2.75E-06	1.21E-05
Cadmium	1,591	1.08E-06	15.5%	0.946	2.52E-04	1.11E-03
Chromium	1,591	1.37E-06	15.5%	0.946	3.21E-04	1.41E-03
Cobalt	1,591	8.24E-08	15.5%	0.946	1.93E-05	8.44E-05
Copper	1,591	8.33E-07	15.5%	0.946	1.95E-04	8.54E-04
Manganese	1,591	3.73E-07	15.5%	0.946	8.72E-05	3.82E-04
Mercury	1,591	2.55E-07	15.5%	0.946	5.96E-05	2.61E-04
Molybdenum	1,591	1.08E-06	15.5%	0.946	2.52E-04	1.11E-03
Nickel	1,591	2.06E-06	15.5%	0.946	4.82E-04	2.11E-03
Selenium	1,591	2.35E-08	15.5%	0.946	5.51E-06	2.41E-05
Vanadium	1,591	2.25E-06	15.5%	0.946	5.28E-04	2.31E-03
Zinc	1,591	2.84E-05	15.5%	0.946	6.65E-03	2.91E-02

Emission Point Source Calculations

Source ID: **38**
Description: **Hot Pressure Relief Vent (Flare)**
Startup, Shutdown and Malfunction estimates

Inputs	Value	Units
Pilot Natural Gas Flowrate (Average)	459	scf/hr
Operation Hours	8,760	hr/yr
Higher Heating Value, Natural Gas	1,000	Btu/scf
Number of Startup/Shutdown Events	26	events/yr
Startup Venting Duration	8.00	hours
Startup Venting Volume	160,000	Nm^3
Startup Venting Heating Value	94	Btu/scf
Shutdown Venting Duration	0.500	hours
Shutdown Venting Volume	15,000	Nm^3
Shutdown Venting Heating Value	272	Btu/scf
Maximum Venting Rate	100%	%
Control Efficiency	98%	%

Emissions Summary

Pollutant	Max Hourly Emissions (lb/hr)	Annual Emission Rate (tpy)
TSP[1]	0.51	0.0763
PM_{10}[1]	0.51	0.0763
$PM_{2.5}$[1]	0.51	0.0763
NO_x	20.3	2.96
SO_2	0.0189	0.00360
CO	239	31.8
VOC	0.042	0.0152
CO2e	15,319	2,235
Lead (Pb)	2.30E-07	1.01E-06

[1] Assume PM_{10}=$PM_{2.5}$=TSP

Start Up Venting Emission Calculations

Calculate Flue Gas Components

Component	Concentration (ppmv)	Partial Volumetric Flow Rate* (Nm^3/hr)	Molar Vent Rate (kg-mol/hr)	Molecular Weight (kg/kg-mol)	Average Flow Rate (kg/hr)	Average Flow Rate (lb/hr)	Maximum Flow Rate (lb/hr)
CO_{2e}	46,300	890	39.7	44.0	1,748.37	3,854.48	3,854.48
H_2O	46,300	890	39.73	18.0	715.47	1,577.34	1,577.34
N_2	515,300	9,910	442	28.0	12,385.91	27,306.23	27,306.23
CO	60,800	1,169	52.2	28.0	1,461.25	3,221.50	3,221.50
H_2	263,300	5,063	226	2.02	455.41	1,004.00	1,004.00
CH_4	108,000	2,077	93	16.0	1,486.62	3,277.44	3,277.44
Total	**1,040,000**	**20,000**	**892**	**-**	**18,253.03**	**40,241.00**	**40,241.00**

* - Assumes ideal gas behavior.
Startup Concentrations estimate from Manufacturer

Calculate Flue Gas Emissions

Component	Emission Factor (lb/MMBtu)	Venting Heat Value (MMBtu/hr)	Average Flow Rate (lb/hr)	Maximum Flow Rate (lb/hr)	Control Efficiency	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
CO_{2e}	-	-	3,854.48	3,854.48	0%	3,854.48	401
CO			3,221.50	3,221.50	98%	64.4	6.70
CH_4	-	-	3,277.44	3,277.44	98%	65.5	6.82
TSP	0.00760	66	-	-	-	0.50	0.0525
NOx	0.0700	66	-	-	-	4.6	0.483
SO_2	0.0000648	66	-	-	-	0.00	4.47E-04
VOC	5.94E-04	66	-	-	-	0.04	4.10E-03

* - Assumes ideal gas behavior.

Shut Down Venting Emission Calculations

Calculate Flue Gas Components

Component	Concentration (ppmv)	Partial Volumetric Flow Rate* (Nm^3/hr)	Molar Vent Rate (kg-mol/hr)	Molecular Weight (kg/kg-mol)	Average Flow Rate (kg/hr)	Average Flow Rate (lb/hr)	Maximum Flow Rate (lb/hr)
CO_2	184,015	3,539	158	44.0	6,948.71	15,319.27	15,319.27
H_2O	70,632	1,358	60.6	18.0	1,091.47	2,406.28	2,406.28
N_2	5,576	107	4.78	28.0	134.03	295.49	295.49
CO	225,836	4,343	194	28.0	5,427.69	11,966.00	11,966.00
H_2	478,625	9,204	411	2.02	827.84	1,825.06	1,825.06
CH_4	35,316	679	30.3	16.0	486.13	1,071.72	1,071.72
Total	**1,000,000**	**19,231**	**858**	**-**	**14,915.87**	**32,883.83**	**32,883.83**

* - Assumes ideal gas behavior.
Shutdown Concentrations estimate from Manufacturer

Calculate Flue Gas Emissions

Component	Emission Factor (lb/MMBtu)	Venting Heat Value (MMBtu/hr)	Average Flow Rate (lb/hr)	Maximum Flow Rate (lb/hr)	Control Efficiency	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
CO_{2e}	-	-	15,319.27	15,319.27	0.00%	15,319.27	1,593
CO	-	-	11,966.00	11,966.00	98.0%	239	24.9
CH_4	-	-	1,071.72	1,071.72	98.0%	21.4	2.23
TSP	2.68E-04	288.2	0.00441	0.00441	0.00%	0.082	0.00850
NOx	7.00E-02	288.2	-	-	-	20.2	2.10
SO_2	6.48E-05	288.2	-	-	-	0.0187	0.00194
VOC	0.00	288.2	0.00	0.00	98.0%	0.00	0.00

* - Assumes ideal gas behavior.

Pilot Lights Emission Calculations

Flare pilot emission calculations are based on AP-42 Section 1.4 (07/98) Natural Gas Combustion, Table 1.4-1 and Table 1.4-2 factors for Large Wall-Fired Boilers, Uncontrolled (Post-NSPS)

(1) Calculate average/maximum heat input values (MMBtu/hr) based on natural gas flowrate

Fuel Component	**Natural Gas Flowrate (Average)[1]**	**Natural Gas Flowrate (Maximum)[1]**	**Gross Heating Value**	**Average Hourly Heat Input**	**Maximum Hourly Heat Input**
	(scf/hr)	**(kg/hr)**	**(BTU/scf)**	**(MMBtu/hr)**	**(MMBtu/hr)**
Natural Gas	459	459	1,000	0.459	0.459

(2) Calculate average (lb/hr and tons/yr) and maximum hourly (lb/hr) emissions

Pollutant	**Emission Factor[2]**	**Emission Factor**	**Average Hourly Heat Input**	**Average Emission Rate**	**Annual Emission Rate**
	(lb/ 10^6 scf)	**(lb/MMBtu)**	**(MMBtu/hr)**	**(lb/hr)**	**(tpy)**
TSP[1]	7.60	0.00760	0.459	0.00349	0.0153
NO$_x$	190	0.190	0.459	0.0872	0.382
SO$_2$[3]	0.600	6.00E-04	0.459	2.75E-04	1.21E-03
CO	84.0	0.0840	0.459	0.0386	0.169
VOC	5.50	5.50E-03	0.459	2.52E-03	1.11E-02
CO2e	120,027.00	1.20E+02	0.459	5.51E+01	2.41E+02
Lead (Pb)	5.00E-04	5.00E-07	0.459	2.30E-07	1.01E-06

[1] Assume PM$_{10}$=PM$_{2.5}$=TSP

[2] Emission Factors from AP-42 Table 1.4-1 and Table 1.4-2

[3] SO$_2$ AP-42 emission factor adjusted for typical natural gas Sulfur content

Emission Point Source Calculations

Emission Point Identifier: **NA**
Description: **Seal Gas Generator**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Designed Production Rate, Hourly	360	ton HBI / hr
Total Reformer Volumetric Vent Rate	464,300	Nm^3/hr
Seal Gas System Off-take	25,000	Nm^3/hr
Normal Reformer Firing Rate	1,591	MMBtu/hr
Nominal Natural Gas Supplement Rate	15.5%	lb/MMBtu
NO_X Uncontrolled Concentration, dry at 3% O2	110.0	mg/Nm^3
CO Emission Factor	84.0	lbs / mmft^3
SO_2 Emission Factor	2.300	Manufacturer
VOC Emission Factor	5.50	lbs / mmft^3
Natural Gas, fuel and reduction gas raw materials	380	Nm^3/ton HBI

NOTE:
Seal gas consists of Reformer Flue Gas. Emissions for theseal gas generator are calculated in the same manner as total emissions from the Reformer, and then adjusted by the ratio of flue gas diverted to the seal gas system. The Seal Gas Generator takes exhaust gas from the main ejector stack (Stack ID 29), cools and dehydrates the gas. This gas is then used to seal the tower. While most of the gas remains in the system (i.e. the tower) and exits via the Top Gas, it is assumed for worst-case permitting purposes to be emitted from the BSG Baghouse (Stack ID 8), the Briquetter Baghouse (Stack ID 9) and the Charge Hopper Baghouse (Stack ID 17). Emissions are calculated on this sheet then distributed by the proportions as follows: BSG Baghouse 40%; Charge Hopper Baghouse 40%; Briquetter 20%. Particulate Emissions are based on baghouse performance guarantees.

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)	BSG (Stk 8) 40% (lb/hr)	BSG (Stk 8) 40% (tpy)	Charge Hopper (Stk 17) 40% (lb/hr)	Charge Hopper (Stk 17) 40% (tpy)	Briquetter (Stk 9) 20% (lb/hr)	Briquetter (Stk 9) 20% (tpy)
NOx	5.04	22.1	2.02	8.8	2.02	8.8	1.01	4.4
SO_2	0.60	2.62	0.24	1.05	0.24	1.05	0.12	0.52
CO	7.05	30.9	2.82	12.4	2.82	12.4	1.41	6.2
CO2	31,215	136,723	12,486	54,689	12,486	54,689	6,243	27,345
CO2e	31,222	136,754	12,489	54,702	12,489	54,702	6,244	27,351
VOC	0.462	2.02	0.185	0.81	0.185	0.81	0.092	0.40
Lead (Pb)	1.30E-04	5.70E-04	5.20E-05	2.28E-04	5.20E-05	2.28E-04	2.60E-05	1.14E-04
Benzene	2.74E-05	1.20E-04	1.10E-05	4.80E-05	1.10E-05	4.80E-05	5.48E-06	2.40E-05
Dichlorobenzene	1.57E-05	6.86E-05	6.27E-06	2.74E-05	6.27E-06	2.74E-05	3.13E-06	1.37E-05
Formaldehyde	9.79E-04	4.29E-03	3.92E-04	1.72E-03	3.92E-04	1.72E-03	1.96E-04	8.58E-04
n-Hexane	2.35E-02	1.03E-01	9.40E-03	4.12E-02	9.40E-03	4.12E-02	4.70E-03	2.06E-02
Naphthalene	7.96E-06	3.49E-05	3.19E-06	1.40E-05	3.19E-06	1.40E-05	1.59E-06	6.98E-06
Toluene	4.44E-05	1.94E-04	1.78E-05	7.78E-05	1.78E-05	7.78E-05	8.88E-06	3.89E-05
PAH	1.10E-06	4.80E-06	4.39E-07	1.92E-06	4.39E-07	1.92E-06	2.19E-07	9.61E-07

Calculate Emissions
Via Outlet Concentrations (Manufacutrer)

Pollutant	Normal Vented Flue Flow Rate	Dry Vented Flue Flow Rate	Component Concentration @ 3% O2	Control Efficiency	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(dNm^3/hr)	(mg/dNm^3)	%	(lb/hr)	(tons/yr)
NO_x	25,000	19,598	110	0.00%	5.04	22.1

Emission rate correction to 3% O2: dNm^3 * mg/dNm^3 / 453600 * ((20.9-1.9)/(20.9-3))

Via AP-42 Emission Factor, Chapter 1, Section 4, Tables 1.4-1 and 1.4-2

Pollutant	Normal Reformer Firing Rate	Emission Factor	Seal Gas Compensation	Control Efficiency	Average Emission Rate	Annual Emission Rate
	(MMBtu/hr)	(lbs/mmBtu NG)	-	%	(lb/hr)	(tons/yr)
CO	1,591	0.0824	0.0538	0.00%	7.05	30.9
VOC	1,591	0.00539	0.0538	0.00%	0.462	2.02

Via Reducing Gas and Fuel Maximum

Pollutant	Total NG as Fuel and RM	Max HBI output	Emission Factor	Control Efficiency	Seal Gas Compensation	Average Emission Rate	Annual Emission Rate
	(Nm^3/ton)	(ton HBI/hr)	(lbs/mmcf NG)	%	-	(lb/hr)	(tons/yr)
SO2	380	360	2.3	0.00%	0.0538	0.598	2.621
CO2	380	360	120,000	0.00%	0.0538	31215	136723
CO2e	380	360	120,027	0.00%	0.0538	31222	136754
Lead (Pb)	380	360	5.00E-04	0.00%	0.0538	1.30E-04	5.70E-04

Calculate HAP Emissions (AP-42)

Pollutant	Normal Reformer Firing Rate	AP-42 Emission Factor	Nominal Natural Gas Supplement Rate	Seal Gas System Compensation	Average Emission Rate	Annual Emission Rate
	(MMBtu/hr)	(lb/MBtu)	%	(-)	(lb/hr)	(tons/yr)
Benzene	1,591	2.06E-06	15.5%	0.0538	2.74E-05	1.20E-04
Dichlorobenzene	1,591	1.18E-06	15.5%	0.0538	1.57E-05	6.86E-05
Formaldehyde	1,591	7.35E-05	15.5%	0.0538	9.79E-04	4.29E-03
n-Hexane	1,591	1.76E-03	15.5%	0.0538	2.35E-02	1.03E-01
Naphthalene	1,591	5.98E-07	15.5%	0.0538	7.96E-06	3.49E-05
Toluene	1,591	3.33E-06	15.5%	0.0538	4.44E-05	1.94E-04
PAH	1,591	8.24E-08	15.5%	0.0538	1.10E-06	4.80E-06
Arsenic	1,591	1.96E-07	15.5%	0.0538	2.61E-06	1.14E-05
Barium	1,591	4.31E-06	15.5%	0.0538	5.74E-05	2.52E-04
Beryllium	1,591	1.18E-08	15.5%	0.0538	1.57E-07	6.86E-07
Cadmium	1,591	1.08E-06	15.5%	0.0538	1.44E-05	6.29E-05
Chromium	1,591	1.37E-06	15.5%	0.0538	1.83E-05	8.01E-05
Cobalt	1,591	8.24E-08	15.5%	0.0538	1.10E-06	4.80E-06
Copper	1,591	8.33E-07	15.5%	0.0538	1.11E-05	4.86E-05
Manganese	1,591	3.73E-07	15.5%	0.0538	4.96E-06	2.17E-05
Mercury	1,591	2.55E-07	15.5%	0.0538	3.39E-06	1.49E-05
Molybdenum	1,591	1.08E-06	15.5%	0.0538	1.44E-05	6.29E-05
Nickel	1,591	2.06E-06	15.5%	0.0538	2.74E-05	1.20E-04
Selenium	1,591	2.35E-08	15.5%	0.0538	3.13E-07	1.37E-06
Vanadium	1,591	2.25E-06	15.5%	0.0538	3.00E-05	1.32E-04
Zinc	1,591	2.84E-05	15.5%	0.0538	3.79E-04	1.66E-03

Emission Point Source Calculations

Emission Point Identifier: **8**
Description: **Furnace Dedusting (BSG Dust Collection)**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Iron Oxide Consumption Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	63,000	Nm^3/hr
Clean Gas Concentration	18.0	mg/Nm^3
PM$_{10}$ Mass Fraction	100%	%
PM$_{2.5}$ Mass Fraction (Process Experience)	50%	%

Emissions Summary

Pollutant	Calculation	Emission Rate (lb/hr)	Emission Rate (tpy)
TSP	Below	2.50	11.0
PM$_{10}$	Below	2.50	11.0
PM$_{2.5}$	Below	1.25	5.5
NOx	Seal Gas	2.02	8.8
SO2	Seal Gas	0.24	1.05
CO	Seal Gas	2.82	12.4
CO2	Seal Gas	12486	54689
CO2e	Seal Gas	12489	54702
VOC	Seal Gas	0.18	0.81
Lead (Pb)	Seal Gas	1.77E-04	7.75E-04
Benzene	Seal Gas	1.10E-05	4.80E-05
Dichlorobenzene	Seal Gas	6.27E-06	2.74E-05
Formaldehyde	Seal Gas	3.92E-04	1.72E-03
n-Hexane	Seal Gas	9.40E-03	4.12E-02
Naphthalene	Seal Gas	3.19E-06	1.40E-05
Toluene	Seal Gas	1.78E-05	7.78E-05
PAH	Seal Gas	4.39E-07	1.92E-06
Hg	Below	2.50E-07	1.10E-06
Cd	Below	1.25E-06	5.48E-06
Cr	Below	1.25E-04	5.48E-04
Mg (as MgO)	Below	1.50E-02	6.57E-02
Mn	Below	1.33E-03	5.80E-03
Ni	Below	7.50E-04	3.29E-03
Cu	Below	2.50E-04	1.10E-03
Zn	Below	5.00E-04	2.19E-03

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean gas Dust Loading	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	63,000	18.00	100%	2.50	11.0
PM$_{10}$	63,000	18.00	100%	2.50	11.0
PM$_{2.5}$	63,000	18.00	50%	1.25	5.5

Metal Emissions (based on ore content)

Pollutant	Concentration (ppm)	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
Hg	0.1	2.50E-07	1.10E-06
Cd	0.5	1.25E-06	5.48E-06
Cr	50	1.25E-04	5.48E-04
Mg (as MgO)	6000	1.50E-02	6.57E-02
Mn	530	1.33E-03	5.80E-03
Pb	50	1.25E-04	5.48E-04
Ni	300	7.50E-04	3.29E-03
Cu	100	2.50E-04	1.10E-03
Zn	200	5.00E-04	2.19E-03

Emission Point Source Calculations

Emission Point Identifier: **9**
Description: **Briquetter Dedusting**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	100,000	Nm^3/hr
Clean Gas Concentration	18.0	mg/Nm^3
PM$_{10}$ Mass Fraction	100%	%
PM$_{2.5}$ Mass Fraction	50.0%	%

Emissions Summary

Pollutant	Calculation	Emission Rate (lb/hr)	Emission Rate (tpy)
TSP	Below	3.97	17.4
PM$_{10}$	Below	3.97	17.4
PM$_{2.5}$	Below	1.98	8.7
NOx	Seal Gas	1.01	4.4
SO2	Seal Gas	0.12	0.52
CO	Seal Gas	1.41	6.2
CO2	Seal Gas	6243.07	27344.63
CO2e	Seal Gas	6244.47	27350.78
VOC	Seal Gas	0.09	0.40
Lead (Pb)	Seal Gas	2.24E-04	9.83E-04
Benzene	Seal Gas	5.48E-06	2.40E-05
Dichlorobenzene	Seal Gas	3.13E-06	1.37E-05
Formaldehyde	Seal Gas	1.96E-04	8.58E-04
n-Hexane	Seal Gas	4.70E-03	2.06E-02
Naphthalene	Seal Gas	1.59E-06	6.98E-06
Toluene	Seal Gas	8.88E-06	3.89E-05
PAH	Seal Gas	2.19E-07	9.61E-07
Hg	Below	3.97E-07	1.74E-06
Cd	Below	1.98E-06	8.69E-06
Cr	Below	1.98E-04	8.69E-04
Mg (as MgO)	Below	2.38E-02	1.04E-01
Mn	Below	2.10E-03	9.21E-03
Ni	Below	1.19E-03	5.21E-03
Cu	Below	3.97E-04	1.74E-03
Zn	Below	7.94E-04	3.48E-03

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean gas Dust Loading	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	100,000	18.00	100%	3.97	17.4
PM$_{10}$	100,000	18.00	100%	3.97	17.4
PM$_{2.5}$	100,000	18.00	50%	1.98	8.7

Metal Emissions (based on ore content)

Pollutant	Concentration (ppm)	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
Hg	0.1	3.97E-07	1.74E-06
Cd	0.5	1.98E-06	8.69E-06
Cr	50	1.98E-04	8.69E-04
Mg (as MgO)	6000	2.38E-02	1.04E-01
Mn	530	2.10E-03	9.21E-03
Pb	50	1.98E-04	8.69E-04
Ni	300	1.19E-03	5.21E-03
Cu	100	3.97E-04	1.74E-03
Zn	200	7.94E-04	3.48E-03

Emission Point Source Calculations

Emission Point Identifier: **11a**
Description: **HBI Cooling Conveyor No. 1**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	48,000	Nm^3/hr
Clean Gas Concentration	18.0	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction	50.0%	%

Emissions Summary

Pollutant	Emission Rate (lb/hr)	Emission Rate (tpy)
TSP	1.90	8.3
PM_{10}	1.90	8.3
$PM_{2.5}$	0.95	4.2
Hg	1.90E-07	8.34E-07
Cd	9.52E-07	4.17E-06
Cr	9.52E-05	4.17E-04
Mg (as MgO)	1.14E-02	5.01E-02
Mn	1.01E-03	4.42E-03
Pb	9.52E-05	4.17E-04
Ni	5.71E-04	2.50E-03
Cu	1.90E-04	8.34E-04
Zn	3.81E-04	1.67E-03

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Maximum Clean gas Dust Loading	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	48,000	18.00	100%	1.90	8.3
PM_{10}	48,000	18.00	100%	1.90	8.3
$PM_{2.5}$	48,000	18.00	50%	0.95	4.2

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	1.90E-07	8.34E-07
Cd	0.5	9.52E-07	4.17E-06
Cr	50	9.52E-05	4.17E-04
Mg (as MgO)	6000	1.14E-02	5.01E-02
Mn	530	1.01E-03	4.42E-03
Pb	50	9.52E-05	4.17E-04
Ni	300	5.71E-04	2.50E-03
Cu	100	1.90E-04	8.34E-04
Zn	200	3.81E-04	1.67E-03

Emission Point Source Calculations

Emission Point Identifier: **11b**
Description: **HBI Cooling Conveyor No. 2**

Inputs

Description	**Value**	**Units**
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	48,000	Nm^3/hr
Clean Gas Concentration	18.0	mg/Nm^3
PM_{10} Mass Fraction	100%	%
$PM_{2.5}$ Mass Fraction	50.0%	%

Emissions Summary

Pollutant	**Emission Rate (lb/hr)**	**Emission Rate (tpy)**
TSP	1.90	8.3
PM_{10}	1.90	8.3
$PM_{2.5}$	0.95	4.2
Hg	1.90E-07	8.34E-07
Cd	9.52E-07	4.17E-06
Cr	9.52E-05	4.17E-04
Mg (as MgO)	1.14E-02	5.01E-02
Mn	1.01E-03	4.42E-03
Pb	9.52E-05	4.17E-04
Ni	5.71E-04	2.50E-03
Cu	1.90E-04	8.34E-04
Zn	3.81E-04	1.67E-03

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate (Nm^3/hr)	Maximum Clean gas Dust Loading (mg/Nm^3)	Mass Fraction of Total PM %	Average Emission Rate (lb/hr)	Annual Emission Rate (tons/yr)
TSP	48,000	18.00	100%	1.90	8.3
PM_{10}	48,000	18.00	100%	1.90	8.3
$PM_{2.5}$	48,000	18.00	50%	0.95	4.2

Metal Emissions (based on ore content)

Pollutant	Concentration (ppm)	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
Hg	0.1	1.90E-07	8.34E-07
Cd	0.5	9.52E-07	4.17E-06
Cr	50	9.52E-05	4.17E-04
Mg (as MgO)	6000	1.14E-02	5.01E-02
Mn	530	1.01E-03	4.42E-03
Pb	50	9.52E-05	4.17E-04
Ni	300	5.71E-04	2.50E-03
Cu	100	1.90E-04	8.34E-04
Zn	200	3.81E-04	1.67E-03

Emission Point Source Calculations

Emission Point Identifier: **12**
Description: **Transfer & Product Screening Station No. 1 (Pre Pile)**

Inputs

Description	**Value**	**Units**
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	40,000	Nm^3/hr
Cleaned Gas Particulate Concentration	4.57	mg/Nm^3
PM_{10} Mass Fraction	100.0%	%
$PM_{2.5}$ Mass Fraction	90.0%	%

Emissions Summary

Pollutant	**Average Emission Rate (lb/hr)**	**Annual Emission Rate (tpy)**
TSP	0.40	1.77
PM_{10}	0.40	1.77
$PM_{2.5}$	0.36	1.59
Hg	4.03E-08	1.77E-07
Cd	2.02E-07	8.83E-07
Cr	2.02E-05	8.83E-05
Mg (as MgO)	2.42E-03	1.06E-02
Mn	2.14E-04	9.36E-04
Pb	2.02E-05	8.83E-05
Ni	1.21E-04	5.30E-04
Cu	4.03E-05	1.77E-04
Zn	8.06E-05	3.53E-04

Notes:
Transfer & Product Screening Station No. 1 (Pre Pile)

Basis:

The Screening Scrubber captures particulate emissions from the screening operations. This includes two transfer operations.

Calculations:

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	40,000	4.6	100%	0.40	1.77
PM_{10}	40,000	4.6	100%	0.40	1.77
$PM_{2.5}$	40,000	4.6	90%	0.36	1.59

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	4.03E-08	1.77E-07
Cd	0.5	2.02E-07	8.83E-07
Cr	50	2.02E-05	8.83E-05
Mg (as MgO)	6000	2.42E-03	1.06E-02
Mn	530	2.14E-04	9.36E-04
Pb	50	2.02E-05	8.83E-05
Ni	300	1.21E-04	5.30E-04
Cu	100	4.03E-05	1.77E-04
Zn	200	8.06E-05	3.53E-04

Emission Point Source Calculations

Case: **14**
Description: **HBI Pile**

Inputs	Value	Units
Material storage time	8,760	hrs
Storage duration	365	days
Average wind speed	12.0	mph
Number of dry days per year	288	days
Percentage of time wind speed > 12 mph	50%	% (actual)
Annual Material Throughput	2,205,000	tons/yr
Excess Receiving Capacity (Annual)	110%	%
Conveyor System Service Factor (Hourly)	115%	%
Transfer rate-stacker	1,500	tons/hr
Transfer rate-reclaimer	1,500	tons/hr
Wind Erosion Chemical Suppression	95%	%
Control efficiency -stacker	90%	%
Control efficiency -reclaimer	75%	%
Material moisture content	6.5	%
Material silt content	2.5	%
Storage area ground surface silt content	6.0	%
Percentage PM 10 in material	13.0%	%
Pile maintenance/traffic hours	365	days
Dozer miles per day (only if HBI reoxidizes, unusal)	0.25	mi
Dozer average weight	5	tons
Front end loader miles per day	0.25	mi
Front end loader average weight	5	tons
Control efficiency - maintenance/traffic	70%	%

Corpus Christi average wind speed (previous 60 years) http://www.ncdc.noaa.gov/oa/climate/online/ccd/avgwind.html, 7/27/12
Corpus Christi mean number of days with precipitation .01" or more (previous 63 years) http://lwf.ncdc.noaa.gov/oa/climate/online/ccd/prcpdays.html, 7/27/12

Emissions Summary

TSP

Source Description	Average Emission Rate	Annual Emission Rate
	(lbs/hr)	(tpy)
Material Transfers In/Out of Storage	0.26	1.12
Equipment Traffic in Storage Area	0.03	0.11
Wind Erosion	0.04	0.15
TOTAL	**0.32**	**1.39**

PM$_{10}$

Source Description	Average Emission Rate	Annual Emission Rate
	(lbs/hr)	(tpy)
Material Transfers In/Out of Storage	0.12	0.51
Equipment Traffic in Storage Area	0.01	0.03
Wind Erosion	0.00	0.02
TOTAL	**0.13**	**0.56**

PM$_{2.5}$

Source Description	Average Emission Rate	Annual Emission Rate
	(lbs/hr)	(tpy)
Material Transfers In/Out of Storage	0.04	0.16
Equipment Traffic in Storage Area	0.00	0.00
Wind Erosion	0.00	0.00
TOTAL	**0.04**	**0.16**

[1] Per AP-42 (Table 13.2.2-2), PM-30 is assumed equivalent to toal suspended particle matter (TSP)

Metal Emissions (based on ore content)

Pollutant	Concentration	Average Emission Rate	Annual Emission Rate
	(ppm)	(lb/hr)	(tpy)
Hg	0.1	4.83E-08	2.12E-07
Cd	0.5	2.42E-07	1.06E-06
Cr	50	2.42E-05	1.06E-04
Mg (as MgO)	6000	2.90E-03	1.27E-02
Mn	530	2.56E-04	1.12E-03
Pb	50	2.42E-05	1.06E-04
Ni	300	1.45E-04	6.35E-04
Cu	100	4.83E-05	2.12E-04
Zn	200	9.67E-05	4.23E-04

Material Stacking and Reclaiming

Calculation Basis:

Calculation methodology derived from Table 12.5-4 "*Pile Formation, Staacker, Pellet Ore,*" AP-42 (10/86).
Transfer Locations: (1) stacker/dump to pile; (2) reclaimer/loader out of pile onto conveyor.

Emission Calculations

Average Annual Particulate (TSP) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr[1])	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (stacker)	2,425,743	0.0024	1.0	5,822	90.00%	0.07	0.29
Storage Pile (reclaimer)	2,425,743	0.0024	1.0	5,822	75.00%	0.17	0.73
TOTAL						**0.23**	**1.02**

Average Annual Particulate (PM_{10}) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr[1])	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (stacker)	2,425,743	0.0011	1.0	2,668	90.00%	0.03	0.13
Storage Pile (reclaimer)	2,425,743	0.0011	1.0	2,668	75.00%	0.08	0.33
TOTAL						**0.11**	**0.47**

Average Annual Particulate ($PM_{2.5}$) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr[1])	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (stacker)	2,425,743	0.00034	1.0	825	90.00%	0.01	0.04
Storage Pile (reclaimer)	2,425,743	0.00034	1.0	825	75.00%	0.02	0.10
TOTAL						**0.03**	**0.14**

[1] capacity converted to US tons

Equipment Traffic in Storage Areas

Calculation Basis:

Calculation methodology derived from Section 13.2.2, "*Unpaved Roads,*" dated AP-42 (11/06) (in accordance with guidance from Section 13.2.4)
Accounts for bulldozer (on pile) and front-end loader (between piles)

Emission factor (lb/VMT) = $k(s/12)^a (W/3)^b$

where:
s = silt content of aggregate (%)[1]
W = mean vehicle weight [2]
k, a, b = empirical constants for industrial roads [3]

Empirical Constants for Industrial Roads			
	PM-2.5	**PM-10**	**PM-30**
k	0.15	1.5	4.9
a	0.9	0.9	0.70
b	0.45	0.45	0.45

Note:
1. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06
2. From AP42, Section 13.2.2, Table 13.2.2-3, Dated 11/06
3. From AP42, Section 13.2.2, Table 13.2.2-2, Dated 11/06

Calculation of Average Annual Controlled Emission Rate

Average Annual Controlled Emission Rate (tons/yr) = Vehicle Miles traveled in an day (VMT/day) * EF (lb PM / VMT) * (number of days per year)* (1-control efficiency) / (2000 lb/ton)
Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Emission Calculations:

Average Annual Particulate Emissions (TSP) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	0.25	3.80	0.9	70.00%	0.01	0.05
Front-end loaders	0.25	3.80	0.9	70.00%	0.01	0.05
TOTAL					**0.02**	**0.10**

Average Annual Particulate Emissions (PM10) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	0.25	1.01	0.3	70.00%	0.00	0.01
Front-end loaders	0.25	1.01	0.3	70.00%	0.00	0.01
TOTAL					**0.01**	**0.03**

Average Annual Particulate Emissions (PM$_{2.5}$) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	0.25	0.101	0.0	70.00%	0.00	0.00
Front-end loaders	0.25	0.101	0.0	70.00%	0.00	0.00
TOTAL					**0.00**	**0.00**

Wind Erosion

Calculation Basis:

Calculation methodology derived from EPA documents:
"Fugitive Emissions from Integrated Iron and Steel Plants", Section 2.2.4, Figure 3-9 (March 1978) and
"*Iron and Steel Plant Open Source Fugitive Emission Evaluation*", Section 2.1.3 and Table 2-1, Equation (8) (August 1978)
Climatological data from New Orleans Regional Airport Weather Station (Jan-Dec 2007)

$$\text{Emission factor (lb/ton)} = 0.05(s/1.5)(D/90)(d/235)(f/15)$$

where:
s = silt content of aggregate (%)[1]
D = duration of storage (days)
d = dry days per year[2]
f = percentage of time wind speed exceeds 12 mph

Note:
1. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06
2. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06

Calculation of Average Annual Controlled Emission Rate

Average Annual Controlled Emission Rate (tons/yr) = tons of aggregate through storage cycle (tons) * EF (lb PM / VMT) * (1 - Control efficiency)
Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Emission Calculations:

Average Annual Particulate Emissions (TSP) from Wind Erosion from Storage Piles

Aggregate	Throughput[1]	Emission Factor	Uncontrolled Emissions Rate	Overall Control Efficiency	Controlled Average Emission Rate	
	(tons/yr)	(lb/ton)	(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
HBI Briquettes	404,290	0.01	5,582	95.00%	0.03	0.14

Average Annual Particulate Emissions (PM$_{10}$) from Wind Erosion from Storage Piles

Aggregate	TSP Controlled Avg. Emission Rate		% PM 10	Controlled Average Emission Rate	
	(lb/hr$_{avg}$)	(tpy)	%	(lb/hr$_{avg}$)	(tpy)
HBI Briquettes	0.03	0.14	13.00%	0.004	0.018

Average Annual Particulate Emissions (PM$_{2.5}$) from Wind Erosion from Storage Piles

Aggregate	TSP Controlled Avg. Emission Rate		% PM$_{2.5}$[2]	Controlled Average Emission Rate	
	(lb/hr$_{avg}$)	(tpy)	%	(lb/hr$_{avg}$)	(tpy)
HBI Briquettes	0.03	0.14	1.95%	0.001	0.003

[1] Assume 2-month supply onsite at all times

[2] Per an EPA document showing that PM2.5/PM10 for metallic ore and coal piles is 0.15- www.epa.gov/ttn/chief/conference/ei15/session14/cowherd.pdf

Emission Point Source Calculations

Emission Point Identifier: **13**
Description: **Transfer & Product Screening Station No. 2 (Post Pile)**

Inputs

Description	Value	Units
Annual Hours of Operation	8,760	hr/yr
Designed Production Rate	2,205,000	tons/yr
Design Volmetric Vent Rate	40,000	Nm^3/hr
Cleaned Gas Particulate Concentration	4.57	mg/Nm^3
PM$_{10}$ Mass Fraction	100.0%	%
PM$_{2.5}$ Mass Fraction	90.0%	%

Emissions Summary

Pollutant	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
TSP	0.40	1.77
PM$_{10}$	0.40	1.77
PM$_{2.5}$	0.36	1.59
Hg	4.03E-08	1.77E-07
Cd	2.02E-07	8.83E-07
Cr	2.02E-05	8.83E-05
Mg (as MgO)	2.42E-03	1.06E-02
Mn	2.14E-04	9.36E-04
Pb	2.02E-05	8.83E-05
Ni	1.21E-04	5.30E-04
Cu	4.03E-05	1.77E-04
Zn	8.06E-05	3.53E-04

Basis:

The baghouse captures particulate emissions from the screening operations. This includes two transfer operations.

Calculations:

Calculate Emissions

Pollutant	Normal Dedusting Airflow Rate	Nominal Clean Gas Concentration	Mass Fraction of Total PM	Average Emission Rate	Annual Emission Rate
	(Nm^3/hr)	(mg/Nm^3)	%	(lb/hr)	(tons/yr)
TSP	40,000	4.57	100%	0.40	1.77
PM$_{10}$	40,000	4.57	100%	0.40	1.77
PM$_{2.5}$	40,000	4.57	90%	0.36	1.59

Metal Emissions (based on ore content)

Pollutant	Concentration (ppm)	Average Emission Rate (lb/hr)	Annual Emission Rate (tpy)
Hg	0.1	4.03E-08	1.77E-07
Cd	0.5	2.02E-07	8.83E-07
Cr	50	2.02E-05	8.83E-05
Mg (as MgO)	6000	2.42E-03	1.06E-02
Mn	530	2.14E-04	9.36E-04
Pb	50	2.02E-05	8.83E-05
Ni	300	1.21E-04	5.30E-04
Cu	100	4.03E-05	1.77E-04
Zn	200	8.06E-05	3.53E-04

Emission Point Source Calculations

Case: **36**
Description: **Remet / Fines Storage**

Inputs	Value	Units
Material storage time	8,760	hrs
Storage duration	365	days
Average wind speed	12.0	mph
Number of dry days per year	288	days
Percentage of time wind speed > 12 mph	50%	% (actual)
Annual Material Throughput[1]	130,000	tons/yr
Excess Receiving Capacity (Annual)	110%	%
Conveyor System Service Factor (Hourly)	100%	%
Transfer rate-stacker	50	tons/hr
Transfer rate-reclaimer	50	tons/hr
Wind Erosion Chemical Suppression	95%	%
Control efficiency -stacker	90%	%
Control efficiency -reclaimer	75%	%
Material moisture content	3.2	%
Material silt content	4.3	%
Storage area ground surface silt content	6.0	%
Percentage PM 10 in material, estimate as screened material	20.0%	%
Pile maintenance/traffic hours	365	days
Dozer miles per day	0.5	mi
Dozer average weight	20	tons
Front end loader miles per day	1	mi
Front end loader average weight	20	tons
Control efficiency - maintenance/traffic	70%	%

http://www.ncdc.noaa.gov/oa/climate/online/ccd/avgwind.html
Corpus Christi mean number of days with precipitation .01" or more (previous 63 years) http://lwf.ncdc.noaa.gov/oa/climate/online/ccd/prcpdays.html, 7/27/12
http://www.ncdc.noaa.gov/oa/climate/online/ccd/avgwind.html
Estimated maximum reject of ore or product combined

[1] Across multiple remet and storage piles and bunkers, facility annual total

Emissions Summary

TSP

Source Description	**Average Emission Rate (lbs/hr)**	**Annual Emission Rate (tpy)**
Material Transfers In/Out of Storage	0.00	0.02
Equipment Traffic in Storage Area	0.15	0.64
Wind Erosion	0.02	0.08
TOTAL	**0.17**	**0.74**

PM$_{10}$

Source Description	**Average Emission Rate (lbs/hr)**	**Annual Emission Rate (tpy)**
Material Transfers In/Out of Storage	0.00	0.01
Equipment Traffic in Storage Area	0.04	0.17
Wind Erosion	0.00	0.02
TOTAL	**0.04**	**0.20**

PM$_{2.5}$

Source Description	**Average Emission Rate (lbs/hr)**	**Annual Emission Rate (tpy)**
Material Transfers In/Out of Storage	0.001	0.00
Equipment Traffic in Storage Area	0.004	0.02
Wind Erosion	0.001	0.00
TOTAL	**0.005**	**0.02**

[1]Per AP-42 (Table 13.2.2-2), PM-30 is assumed equivalent to toal suspended particle matter (TSP)

Metal Emissions (based on ore content)

Pollutant	**Concentration (ppm)**	**Average Emission Rate (lb/hr)**	**Annual Emission Rate (tpy)**
Hg	0.1	1.70E-08	7.44E-08
Cd	0.5	8.49E-08	3.72E-07
Cr	50	8.49E-06	3.72E-05
Mg (as MgO)	6000	1.02E-03	4.46E-03
Mn	530	9.00E-05	3.94E-04
Pb	50	8.49E-06	3.72E-05
Ni	300	5.09E-05	2.23E-04
Cu	100	1.70E-05	7.44E-05
Zn	200	3.40E-05	1.49E-04

Material Transfers In/Out of Storage

Calculation Basis:

Calculation methodology derived from Table 12.5-4 "*Pile Formation, Stacker, Pellet Ore,*" AP-42 (10/86).
Transfer Locations: (1) stacker/dump to pile; (2) reclaimer/loader out of pile onto conveyor; (3) equipment traffic within pile area.

Emission Calculations

Average Annual Particulate (TSP) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr^1)	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (drop point)	143,014	0.0024	1.0	343	90.00%	0.004	0.017
TOTAL						**0.004**	**0.017**

Average Annual Particulate (PM$_{10}$) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr^1)	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (drop point)	143,014	0.0011	1.0	157	90.00%	0.002	0.008
TOTAL						**0.002**	**0.008**

Average Annual Particulate (PM$_{2.5}$) Emissions from Material Storage Pile Transfers

Transfer Location	Annual Throughput	Emission Factor	Number of Drops	Uncontrolled Emissions	Overall Control Efficiency	Controlled Emission Rates	
	(tons/yr^1)	(lb/ton)		(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Storage Pile (drop point)	143,014	0.00034	1.0	49	90.00%	0.001	0.002
TOTAL						**0.001**	**0.002**

[1] capacity converted to US tons

Equipment Traffic in Storage Areas

Calculation Basis:

Calculation methodology derived from Section 13.2.2, "*Unpaved Roads,*" dated AP-42 (11/06) (in accordance with guidance from Section 13.2.4)
Accounts for bulldozer (on pile) and front-end loader (between piles)

Emission factor (lb/VMT) = $k(s/12)^a (W/3)^b$

Empirical Constants for Industrial Roads			
	PM-2.5	**PM-10**	**PM-30**
k	**0.15**	**1.5**	**4.9**
a	**0.9**	**0.9**	**0.70**
b	**0.45**	**0.45**	**0.45**

where:
s = silt content of aggregate (%)^1
W = mean vehicle weight ^{2}
k, a, b = empirical constants for industrial roads ^{3}

Note:
1. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06
2. From AP42, Section 13.2.2, Table 13.2.2-3, Dated 11/06
3. From AP42, Section 13.2.2, Table 13.2.2-2, Dated 11/06

Calculation of Average Annual Controlled Emission Rate

Average Annual Controlled Emission Rate (tons/yr) = Vehicle Miles traveled in an day (VMT/day) * EF (lb PM / VMT) * (number of days per year)* (1-control efficiency) / (2000 lb/ton)
Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Emission Calculations:

Average Annual Particulate Emissions (TSP) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	1	7.08	3.5	70.0%	0.044	0.19
Front-end loaders	1	7.08	7.1	70.0%	0.089	0.39
TOTAL					**0.13**	**0.58**

Average Annual Particulate Emissions (PM10) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	1	1.89	0.9	70.0%	0.012	0.05
Front-end loaders	1	1.89	1.9	70.0%	0.024	0.10
TOTAL					**0.035**	**0.16**

Average Annual Particulate Emissions ($PM_{2.5}$) from Equipment Traffic in Storage Areas

Vehicle Type	Vehicle Miles Traveled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rates	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
Dozers	1	0.189	0.1	70.0%	0.0012	0.005
Front-end loaders	1	0.189	0.2	70.0%	0.0024	0.010
TOTAL					**0.0035**	**0.016**

Wind Erosion

Calculation Basis:

Calculation methodology derived from EPA documents:
"Fugitive Emissions from Integrated Iron and Steel Plants", Section 2.2.4, Figure 3-9 (March 1978) and
"*Iron and Steel Plant Open Source Fugitive Emission Evaluation*", Section 2.1.3 and Table 2-1, Equation (8) (August 1978)
Climatological data from New Orleans Regional Airport Weather Station (Jan-Dec 2007)

Emission factor (lb/ton) = 0.05(s/1.5)(D/90)(d/235)(f/15)

where:
s = silt content of aggregate (%)
D = duration of storage (days)
d = dry days per year^2
f = percentage of time wind speed exceeds 12 mph

Note:
1. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06
2. From AP42, Section 13.2.2, Table 13.2.2-1, Dated 11/06

Calculation of Average Annual Controlled Emission Rate

Average Annual Controlled Emission Rate (tons/yr) = tons of aggregate through storage cycle (tons) * EF (lb PM / VMT) * (1 - Control efficiency)
Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Emission Calculations:

Average Annual Particulate Emissions (TSP) from Wind Erosion from Storage Piles

Aggregate	Throughput	Emission Factor	Uncontrolled Emissions Rate	Overall Control Efficiency	Controlled Average Emission Rate	
	(tons/yr)	(lb/ton)	(lb/yr)	%	(lb/hr$_{avg}$)	(tpy)
Remet / Fines	130,000	0.024	3,087.1	95.00%	0.018	0.077

Average Annual Particulate Emissions (PM_{10}) from Wind Erosion from Storage Piles

Aggregate	TSP Controlled Avg. Emission Rate		% PM 10	Controlled Average Emission Rate	
	(lb/hr$_{avg}$)	(tpy)	%	(lb/hr$_{avg}$)	(tpy)
Remet / Fines	0.018	0.077	20.00%	0.004	0.015

Average Annual Particulate Emissions ($PM_{2.5}$) from Wind Erosion from Storage Piles

Aggregate	TSP Controlled Avg. Emission Rate		% $PM_{2.5}^{2}$	Controlled Average Emission Rate	
	(lb/hr$_{avg}$)	(tpy)	%	(lb/hr$_{avg}$)	(tpy)
Remet / Fines	0.018	0.077	3.00%	0.0005	0.0023

Per an EPA document showing that PM2.5/PM10 for metallic ore and coal piles is 0.15- www.epa.gov/ttn/chief/conference/ei15/session14/cowherd.pdf

Emission Point Source Calculations

Emission Point Identifier: **30**
Description: **Process Water Degassers**

Inputs

Description	**Value**	**Units**	Comments
Annual Hours of Operation	8,760	hr/yr	
Designed Production Rate	2,205,000	tons/yr	Manfacturer
Total Vapor Exhaust Rate	55,000	kg/hr	Manfacturer
CO Mass Fraction	200	ppm	Manfacturer
CO2 Mass Fraction	2080	ppm	Manfacturer
Methane Mass Fraction	40	ppm	Manfacturer

Emissions Summary

Pollutant	**ppm**	**Average Emission Rate (lb/hr)**	**Annual Emission Rate (tpy)**
CO	200	24.3	106
CO2	2080	252	1,105
CO2e	2120	257	1,126

Emission Point Source Calculations

Case: **33**
Description: **Cooling Tower**

Inputs	**Value**	**Units**
Annual operating hours	8,760	hrs
Cooling tower circulating water rate	20,000	gal/min
Percent drift	0.0005%	%
Total Dissolved Solids Content	35,000	ppm
Percent of Total PM that is PM$_{10}$[1]	3.0%	%

[1] Based upon document "Calculating PM$_{10}$ Emissions from Cooling Towers" by Joel Reisman and Gordon Frisbie (Abstract No. 316, Session No. AM-1b)

Emissions Summary

Pollutant	**Average Emission Rate**	**Annual Emission Rate**
	(lbs/hr)	**(tpy)**
TSP	2.61	11.44
PM$_{10}$	0.08	0.34
PM$_{2.5}$	0.08	0.34

[1] Assume TSP=PM$_{10}$=PM$_{2.5}$

Calculation Basis:

Calculation methodology based upon a conservative mass balance approach.

Calculation of Average Annual Emission Rates

Emission Rate (lb/hr$_{avg}$) = Drift rate (gal/min)*TDS content (ppm)/1,000,000*60min/hr*0.13368 ft^3/gallon *62 lb/ft^{3*}1.5 Cycles
Emission Rate (tpy) = Average Controlled Emission Rate (lb/hr$_{avg}$) * annual operating hours (hrs/yr) / (2000 lb/ton)

Emission Calculations

Pollutant	**Cooling Tower Circulating Water Rate**	**% Drift**	**Drift Rate**	**Total Dissolved Solids Content**	**Average Emission Rate**	**Annual Emission Rate**
	(gal/min)	**%**	**(gal/min)**	**(ppm)**	**(lb/hr$_{avg}$)**	**(tpy)**
TSP	20,000	0.0005%	0.10	35,000	2.61	11.44
PM$_{10}$	20,000	0.0005%	0.10	35,000	0.08	0.34
PM$_{2.5}$[1]	20,000	0.0005%	0.10	35,000	0.08	0.34

[1] Assuming PM$_{10}$ = PM$_{2.5}$

Emission Point Source Calculations

Case: **39**
Description: **Paved Road Fugitive Dust**

Inputs	Value	Units
Days of Operation per year	365	days/yr
Average road surface silt loading	9.7	%
Fleet Average Weight	10.50	tons
Vehicle Miles Travelled - Heavy Duty	50	miles/day
Number of Days with > 0.01 inches of rain	77	days/yr
Control efficiency - Sweeping	90%	%

Corpus Christi mean number of days with precipitation .01" or more http://lwf.ncdc.noaa.gov/oa/climate/online/ccd/prcpdays.html, 7/27/12

Emissions Summary

Source Description	Pollutant	Average Emission Rate	Annual Emission Rate
		(lbs/hr)	(tpy)
Paved Road Fugitive Emissions	TSP[1]	0.25	1.08
	PM_{10}	0.05	0.21
	$PM_{2.5}$	0.01	0.03

[1] Per AP-42 (Table 13.2.2-2), PM-30 is assumed equivalent to total suspended particle matter (TSP)

Calculation Basis:

Calculation methodology derived from Section 13.2.1, "*Paved Roads,*" dated AP-42 (11/06)
Accounts for fall vehicular traffic on paved roads (assumes 1 mile ot paved road at facility)

$$Emission\,factor(lb/vehicle\,mile\,traveled) = k * \left(\frac{sL}{2}\right)^{0.65} * \left(\frac{W}{3}\right)^{1.5} - C$$

where: sL = road surface silt loading, g/m^2 [1]
W = average vehicle weight, tons [2]
k = particle size multiplier, lb/VMT [3]
C - emission factor for 1980's vehicle fleet exhaust, brake wear, and tire wear, lb/VMT [4]

Note:

1. From AP42, Section 13.2.1, Table 13.2.1-4, Dated 11/06
2. From AP42, Section 13.2.2, Table 13.2.2-3, Dated 11/06
3. From AP42, Section 13.2.1, Table 13.2.1-1, Dated 11/06
4. From AP42, Section 13.2.1, Table 13.2.1-2, Dated 11/07

Particle Size Range	*k*	*C*
< 30 micrometers	0.082	0.00047
< 15 micrometers	0.02	0.00047
< 10 micrometers	0.016	0.00047
< 2.5 micrometers	0.0024	0.00036

Emission factor extrapolated for natural mitigation (lb/VMT) $E_{ext} = E\,[1 - (P/365)]$

where: E = Uncontrolled emission factor
P = Number of days with > 0.01 inches of precipitation

Emission Calculations:

Calculation of Average Annual Controlled Emission Rate

Average Annual Controlled Emission Rate (tons/yr) = Vehicle Miles travelled in an day (VMT/day) * EF (lb PM / VMT) * (number of days per year)* (1-control effciency) / (2000 lb/ton)

Average Controlled Emission Rate (lb/hr) = Average Controlled Emission Rate (tons/yr)* (2000 lb/ton)/8760 hrs/yr

Average Annual Particulate Emissions (TSP) from Vehicular Traffic on Paved Roads

Emission Point	Vehicle Miles Travelled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rate	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
All Vehicular Traffic	50	1.182	59.1	90%	0.25	1.08
TOTAL					**0.25**	**1.08**

Average Annual Particulate Emissions (PM_{10}) from Vehicular Traffic on Paved Roads

Vehicle Type	Vehicle Miles Travelled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rate	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
All Vehicular Traffic	50	0.230	11.5	90%	0.05	0.21
TOTAL					**0.05**	**0.21**

Average Annual Particulate Emissions ($PM_{2.5}$) from Vehicular Traffic on Paved Roads

Vehicle Type	Vehicle Miles Travelled	Emission Factor	Uncontrolled Emission Rate	Overall Control Efficiency	Controlled Emission Rate	
	(VMT/day)	(lb/VMT)	(lb/day)	%	(lb/hr$_{avg}$)	(tpy)
All Vehicular Traffic	50	0.034	1.7	90%	0.01	0.03
TOTAL					**0.01**	**0.03**

Emission Point Source Calculations

Case: **34**
Description: **Emergency Generator**

Given:
1) 2500kWe Diesel Emergency Power Generator
2) Nominal Emissions data provided by a NSPS Certified manufacturer

Mechanical	kWm	2500
	Load %	100
NOx	g/kWh	5.822
HC	g/kWh	0.179
CO	g/kWh	0.689
PM	g/kWh	0.085

	Fuel consumption (gph)	lb CO2/gallon
CO2	174.6	22.51

3) Fuel: 15ppm (0.0015%) Sulfur Diesel
4) SO2 emission factor: AP42 Chapter 3, Section 4: 8.09E-3*S = 8.09E-3*.0015 = 1.2E-5 lb/hph = 0.0077 g/kWh
5) NSPS IIII Hours based on Readiness and Maintenance limit of 100 hours (emergency use not limited)

Nominal Emissions Calculation - Requested Limit

Pollutant	100% Load (g/kWh)	100% Hourly Per Unit (lbs/hr)	100% Annual Emission Rate (tpy)
NOx	5.822	32.09	1.60
HC	0.179	0.99	0.049
CO	0.689	3.80	0.190
PM	0.085	0.47	0.023
SO2	0.0077	0.04	0.002
CO2	-	3930.81	196.540
CO2e	-	3930.81	196.540

Hours	Per Unit
100%	100

Emission Point Source Calculations

Case: **35**
Description: **Fire Pump**

Given:

1) 175kWe Diesel Emergency Power Generator
2) Nominal Emissions data provided by a NSPS Certified manufacturer

Mechanical	kWm	175
	Load %	100
NOx	g/kWh	5.822
HC	g/kWh	0.179
CO	g/kWh	0.689
PM	g/kWh	0.085

	Fuel consumption (gph)	lb CO2/gallon
CO2	11.4	22.51

3) Fuel: 15ppm (0.0015%) Sulfur Diesel
4) SO2 emission factor: AP42 Chapter 3, Section 4: 8.09E-3*S = 8.09E-3*.0015 = 1.2E-5 lb/hph = 0.0077 g/kWh
5) NSPS IIII Hours based on Readiness and Maintenance limit of 100 hours (emergency use not limited)

Nominal Emissions Calculation - Requested Limit

Pollutant	100% Load (g/kWh)	100% Hourly Per Unit (lbs/hr)	100% Annual Emission Rate (tpy)
NOx	5.822	2.25	0.112
HC	0.179	0.07	0.003
CO	0.689	0.27	0.013
PM	0.085	0.03	0.002
SO2	0.0077	0.00	0.000
CO2	-	256.61	12.831
CO2e	-	256.61	12.831

Hours	Per Unit
100%	100

Emission Fugitives Calculations

Case:	NA
Description:	Natural Gas fugitive losses
Process Equipment IDs:	NA

Given:

1) Emission factors derived from information presented in "Handbook for Estimating Methane Emissions from Canadian Natural Gas Systems" prepared for GRI Canada on May 25, 1998" page nos. 34-35.

Equipment	Reference *	Number of Pieces	Emissions Factor (kg/h/src)	GHGe Emissions (tons/yr)
Onsite Equipment Inventory				
Flexible Hoses - Main Burner A	Connector	200	1.10E-04	4.45E+00
Compensator	Other Flow Meter	200	9.06E-06	3.68E-01
Flexible Hoses - Main Burner B	Connector	80	1.10E-04	1.78E+00
Compensator	Other Flow Meter	80	9.06E-06	1.47E-01
Flexible Hoses - NG	Connector	80	1.10E-04	1.78E+00
Compensator - NG Outlet Manifold	Other Flow Meter	4	9.06E-06	7.35E-03
Compensator - NG Outlet Manifold	Other Flow Meter	4	9.06E-06	7.35E-03
Equipment after pressure reducing and metering station:				
Pressure Regulation Valves (PRV)	PRV	5	1.67E-02	1.69E+01
Safety Valves (PSV)	PRV	2	1.67E-02	6.75E+00
Shut Off Valves (SOV)	Block Valve	7	1.11E-03	1.57E+00
Flow Control Valves (FV)	Control Valve	8	1.97E-02	3.19E+01
Vent Valve	PRV	1	1.67E-02	3.38E+00
Equipment after TOP from Houston/KM:				
Flow Control Valves (FCV)	Control Valve	2	1.97E-02	7.99E+00
Shut Off Valves (SOV)	Block Valve	2	1.11E-03	4.50E-01
Pressure Control Valves (PCV)	PRV	2	1.67E-02	6.75E+00
			Total	**84.3**

* Reference refers to the equipment liste

Attachment 3

Vendor Information

A Better Mousetrap: The History of Midrex Technologies

Part 1

By John T. Kopfle
Director - New Business Development

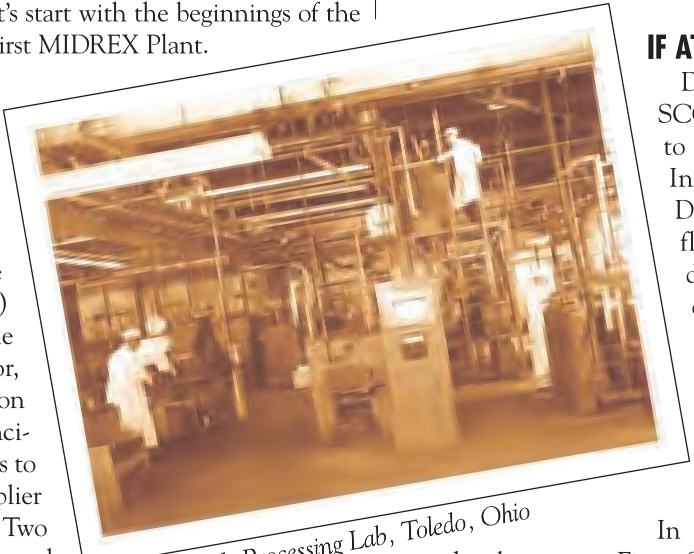
Introduction

The MIDREX® Direct Reduction Process has become the world's most successful direct reduction technology. Each year since 1987, over 60 percent of the world's direct reduced iron has been produced in MIDREX® Direct Reduction Plants. Since 1969, 53 MIDREX® Modules have been installed in 18 countries. As with most industrial technologies, the history of the MIDREX Process has been a combination of vision, entrepreneurship, hard work, mistakes, and luck.

As we celebrate the 30th anniversary of the founding of Midrex Corporation this year, *Direct from Midrex* will retrace the fascinating history of the company and the process. Each issue will cover a portion of the three decades. Let's start with the beginnings of the process and the building of the first MIDREX Plant.

IN THE BEGINNING...

The genesis of the MIDREX Process was technology and market opportunity. The story begins in the 1920s with the founding of the Surface Combustion Company (SCC) in Toledo, Ohio, USA. The company, Midrex's predecessor, was formed to apply combustion processes and heat transfer principles to industrial needs. SCC was to become the world's largest supplier of industrial furnace equipment. Two technological developments proved to be of particular significance in the evolution of the MIDREX Process. First was the design of shaft furnaces for minerals processing, particularly iron ore hardening, or induration. Second was the technical expertise SCC acquired in the combustion and reforming of hydrocarbon gases.



SCC Minerals Processing Lab, Toledo, Ohio

In 1936, SCC started making hydrogen from natural gas in the laboratory. Two years later, a gentleman named Julius Madaras of Detroit approached SCC with an idea for a process to directly reduce iron ore using hydrogen. The process required 0.65 Nm^3 of hydrogen per kilogram of DRI. The economic incentive for the process was the availability of fine hematite ore from iron sands and natural gas in Texas; oilmen and ranchers were among the potential investors. The developers believed this process could compete with blast furnace-produced steel.

After performing successful bench-scale tests, a corporation was formed to develop the technology. A 1.2 m diameter, 1.8 m tall reactor was built and installed at Hooker Chemical Company in Niagara Falls, New York, then moved to Longview, Texas. After some testwork, World War II intervened, and the effort was terminated.

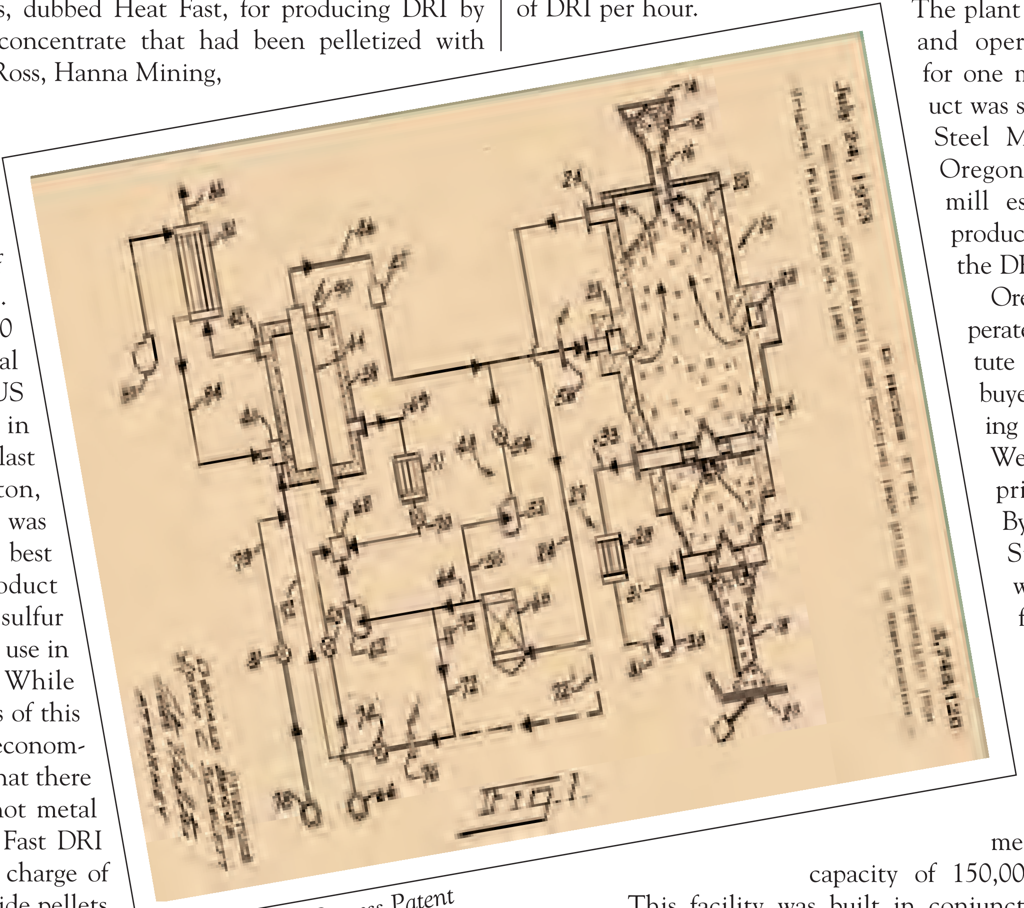
IF AT FIRST YOU DON'T SUCCEED...

During the war, some researchers at SCC began thinking about other ways to reduce iron ore. The director of the Institute of Gas Technology, one Dr. Sullivan, had experience with fluidized beds, which were being developed to crack oil for production of high octane gasoline. A laboratory-scale reactor was built, but the test runs were not successful in completely reducing the iron ore. Due to several personnel changes, no one championed the fluidized bed project and it was terminated.

In 1946, there were two momentous developments. First, SCC developed a shaft furnace for indurating iron oxide pellets. Project sponsors were Pickands-Mather, Interlake Iron, Bethlehem Steel, and Youngstown Sheet and Tube Company. Second, Raymond Patterson approached SCC with a process for reducing mill scale with granulated charcoal. The mill scale was mixed with charcoal, placed in a paper tube, then inserted into an alloy tube. Reducing gas was introduced to the tube and after several hours, the paper tube was removed and the product was crushed and magnetically separated. Then, the magnetic fraction was reduced again with gas, producing a 97 percent iron powder.

SCC formed a subsidiary, Superior Metals Corporation, to produce the powder commercially. A plant was set up in Toledo and began operations in early 1947; the product was sold for 12 cents per pound. About six months later, orders began to drop, and it was found that Hoeganes, a Swedish company, was selling a similar product for eight to ten cents per pound. Superior Metals decided to abandon the technology and the company was dissolved in 1951. Interestingly, Hoeganes still operates plants today in Sweden and the US.

During the late 1940s and 1950s, SCC sold 41 shaft furnaces for indurating iron oxide pellets. In 1959, Midland-Ross Corporation acquired SCC and continued to operate the company as a separate division. In 1963, the Surface Combustion Division (SCD) developed a process, dubbed Heat Fast, for producing DRI by reducing iron ore concentrate that had been pelletized with fine coal. Midland-Ross, Hanna Mining, and National Steel built a $2 million demonstration plant in Cooley, Minnesota, and a sizable quantity of DRI was produced. Approximately 9,000 tons of the material was tested by the US Bureau of Mines in an experimental blast furnace in Bruceton, Pennsylvania. This was believed to be the best use for the product because its high sulfur content would limit use in electric furnaces. While the technical results of this trial were good, an economic analysis showed that there was no benefit in hot metal cost by using Heat Fast DRI versus the standard charge of 100 percent iron oxide pellets. The SCD stopped work on the process.

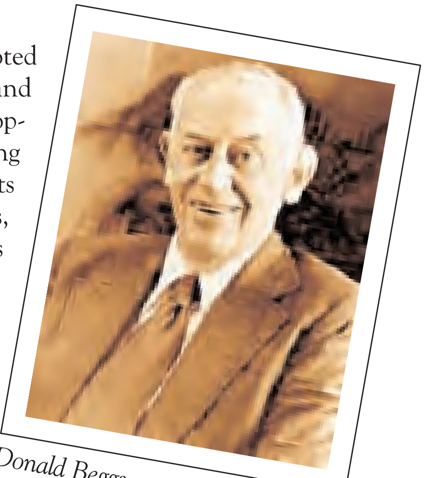


Early MIDREX Process Patent

At about the same time, the Steel Company of Canada and Lurgi Corporation of Germany produced DRI in a rotary kiln using coal as the reductant. These low-sulfur pellets proved an excellent feed material for Stelco's Edmonton, Alberta EAF plant. This technology was the forerunner of the SL/RN Process.

BEGGS PICKS A WINNER

In 1966, Midland-Ross noted the results in Canada and discussed the idea of developing a DR process for producing highly metallized iron pellets for EAF use. Donald Beggs, Manager of the SCD's Research Group, conceived the idea of employing natural gas reforming to produce high quality gas which would be used to reduce iron oxide pellets in a shaft furnace. Beggs proposed that a pilot plant be built in Toledo, incorporating a 0.46 m inside diameter shaft furnace with a capacity of180-225 kg of DRI per hour.



Donald Beggs

The plant was built in 1967 and operated successfully for one month. The product was shipped to Oregon Steel Mills in Portland, Oregon, and the steel mill established a new production record with the DRI.

Oregon Steel was desperate for a scrap substitute because Japanese buyers were purchasing scrap on the US West Coast and the price skyrocketed. By mid-year, Oregon Steel contracted with Midland-Ross for a full-scale prototype plant in Portland. It consisted of two modules, each with a 3.7 m (12 foot) diameter furnace, and a capacity of 150,000 tons per year.

This facility was built in conjunction with Oregon Steel's new meltshop, but it was owned and operated by Midland-Ross, with the product sold on a long-term contract. Midland-Ross decided to retain ownership of the facilities rather than sell equipment and engineering services. Operations began on May 17, 1969.

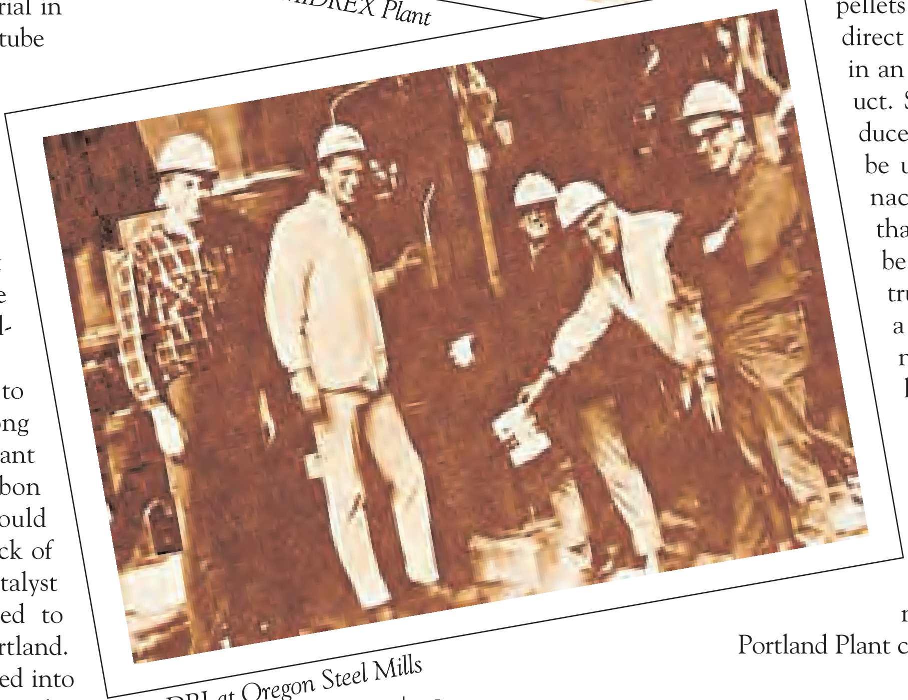
There were a number of problems that were resolved in the laboratory and during start-up of the plant, and the solutions were crucial to the ultimate success of the process. These included:

Carbon Formation

The conventional wisdom in 1968 was that reforming of natural gas was not possible without a large excess of CO_2 and H_2O, and that stoichiometric reforming would cause catalyst degradation. However, the SCD had been performing stoichiometric reforming for 20 years and was confident it would be successful on a commercial basis. The company built a 0.2 m by 7.3 m reformer at its research center. One key to the success of this endeavor was the use of a strong catalyst able to withstand the formation of Boudouard carbon and also endure the crushing action which occurs when the tube cycles between operating and idle conditions.

Oregon Steel Mills MIDREX Plant

Boudouard carbon is formed via the Boudouard reaction, shown in Table I. The concept for preventing it was to place an inert material in the bottom of the catalyst tube (where the gas enters), then active catalyst above. This would enable the gas to be heated above the dangerous temperature range before reforming began. Unfortunately, all attempts at using this approach in the pilot plant failed, and considerable carbon was formed.



First DRI at Oregon Steel Mills

The decision was made to provide an extremely strong catalyst for the Portland Plant so that if Boudouard carbon formed, the catalyst would remain intact. Due to the lack of facilities for producing the catalyst in Toledo, the SCD decided to coat the catalyst in-situ at Portland. The carrier material was loaded into the tubes and a solution of catalyst material was pumped into them. Then the solution was drained out. To provide inert material at the bottom of the tube, water was pumped into the bottom of the tube to remove the catalyst, and then drained.

Boudouard Reaction

$$2CO = CO_2 + C$$

Stoichiometric Reforming Reactions

$$CH_4 + H_2O = CO + 3H_2$$

$$CH_4 + CO_2 = 2CO + 2H_2$$

Table I Reforming Reactions

The plant started up and never experienced a Boudouard carbon problem. After much discussion and research, the SCD figured out the reason for the carbon formation problems in the Toledo laboratory and the lack of them in Portland. The concept developed at Portland is now standard practice in MIDREX Plants.

DRI Degradation

Another conventional wisdom in 1968 was that swelling and weakening of low gangue iron oxide pellets would occur during direct reduction, resulting in an unacceptable product. Since the DRI produced at Portland would be used in electric furnaces, it was crucial that low gangue feed be used. While it was true that when using a typical blast furnace reducing gas high in CO, product degradation did occur, the SCD had confirmed in laboratory tests that with a higher H_2 content, this was not a problem. The Portland Plant confirmed this result.

Clustering

During initial operations, clustering of DRI in the shaft furnace was a continuing problem. SCD surmised that methanation might be the cause by raising the bed temperature to the fusion point of iron. Methanation occurs when the reforming reactions shown in Table I reverse. This result was confirmed when methane was injected into the shaft furnace, suppressing the methanation reaction. Thus was born methane enrichment of the bustle gas, standard practice at all MIDREX Plants today.

In the next issue of DFM, we'll continue our story of the evolution of Midrex. Included will be the purchase of Midrex by Willie Korf and the ensuing growth of the company in the 1970s.

A Better Mousetrap: The History of Midrex Technologies
Part 2

By John T. Kopfle.
Director - New Business Development

Note: Part 1 of this series traced the founding of Midrex's predecessor company, the development of the MIDREX® Direct Reduction Process, and the building of the first commercial plant at Oregon Steel Mills.

A GERMAN VISIONARY APPEARS

In 1948, a nineteen-year-old German named Willy Korf became a general partner in his family's trading company following the death of his father. In the early 1950s, the company grew successfully by selling building materials, including lumber, roofing tiles, and reinforcing steel for rebuilding Germany after World War II.

In 1953, the company began selling welded wire mesh in partnership with another firm. However, difficulties in procuring the material persuaded Korf to begin producing it himself, and he began doing so in 1955 in Kehl, Germany. This enterprise soon faced another problem - obtaining wire rod, the feedstock for mesh. Korf was able to procure rod from France and Belgium. Several of the German steel companies, however, resented Korf's competition, and in 1961 he signed an agreement not to sell wire mesh in Europe, except for Switzerland and Austria, for ten years. Eventually, due to this and other problems, Korf sold his wire mesh operations.

Undaunted, the German entrepreneur began producing other kinds of steel products at home and abroad, including joists, wire, and rebar. This led to more disputes with German steelmakers. In 1968, Korf decided to produce his own steel and he started up a mini-mill in Kehl with scrap as the feedstock. The mill was later named Badische Stahlwerke.

Georgetown Steel

Young Korf and family

Willy Korf

Badische Stahlwerke

Again, the steel companies took offense to Korf's aggressiveness and tried to force him out of business. Their strategy was to manipulate scrap prices to make it difficult for independent steelmakers. Korf decided to find a way to make steel without scrap. In 1968, Midland-Ross presented a paper on the MIDREX® Technology at the AIME General Meeting in New York City. In the audience was Dr. Menthertz Gaier of Korf Industries. A short time later, a meeting was arranged in Toledo between Willy Korf and the Surface Combustion Division of Midland-Ross. Korf told his hosts that he wanted a direct reduction plant designed for him immediately for Georgetown Steel in South Carolina, where he was building a mini-mill, and for Hamburg, Germany, where he planned to develop a steel complex. The SCD people were reluctant, because the Oregon Steel direct reduction plant was just completing construction and would not be operating for several weeks. Korf persisted, and wrote a check on the spot for SCD to begin design work. Both plants were rated at 400,000 tons per year, included 4.9 m diameter furnaces, and started up in 1971.

In 1969, Midland-Ross formed the Midrex Division to exploit the market potential of the MIDREX Process. How was the name chosen? The story is that William Marston, Vice President and General Manager of the SCD, had received 32 potential names from the Midland-Ross public relations department, which were written on a page in two columns of 16 each. Marston closed his eyes, stabbed at the paper with his pencil, and selected the name "Midrex."

In 1970, Midrex contracted to build a MIDREX® Series 400 Module for Sidbec-Dosco in Quebec, Canada. In April, 1973, Midrex achieved what became known as "direct reduction's finest hour." Iron Age Magazine stated:

"The direct reduction plant at Sidbec-Dosco went onstream in April of 1973. Within a week, the Midrex Plant was up to its rated capacity of 1,000 tons a day. This was something of a landmark because the commercialization of direct reduction has seen a goodly number of set-backs and disasters."

In December, Midland-Ross agreed to sell the Midrex Division to Korf. To paraphrase a noteworthy television commercial, Korf liked the technology so much he bought the company. Midrex Corporation was formed and moved to Charlotte, NC in 1974. Two reasons for the move were Korf's desire for Midrex to be close to the Georgetown mill and to North Carolina National Bank, a source of financing for the fledgling company.

Sidbec MIDREX® Plants

A Successful Decade

1971 Georgetown Steel
Hamburg Stahlwerke
1973 Sidbec-Dosco I
1976 SIDERCA
1977 SIDOR I
Sidbec-Dosco II
1978 QASCO
Acindar
1979 SIDOR II

A SUCCESSFUL DECADE

By 1974, there were MIDREX Plants operating in the United States, Germany, and Canada. During the later part of the 1970s, numerous plants were sold, many in the developing world. Countries such as Argentina, Qatar, and Venezuela needed steel to support development, and possessed the capital (some via revenues from petroleum sales) and natural gas required for direct reduction. The facilities were often government-owned. By 1979, Midrex had started up over five million tons (Mt) of DR capacity. Annual world direct reduced iron production grew from 0.8 Mt in 1970 to 7 Mt in 1980. During the early to mid-1970s, there was considerable optimism regarding future steel and DRI production. In 1975 some analysts, including the International Iron and Steel Institute and Battelle Laboratories, were forecasting that annual world steel production would reach 1 billion tons and annual DRI production would be 130 Mt by 1985. The actual figures were 719 Mt and 11 Mt!

During the decade of the 1970s, numerous innovations were developed for the MIDREX Process, including larger shaft furnaces, in-situ reforming, heat recovery and cold briquetting.

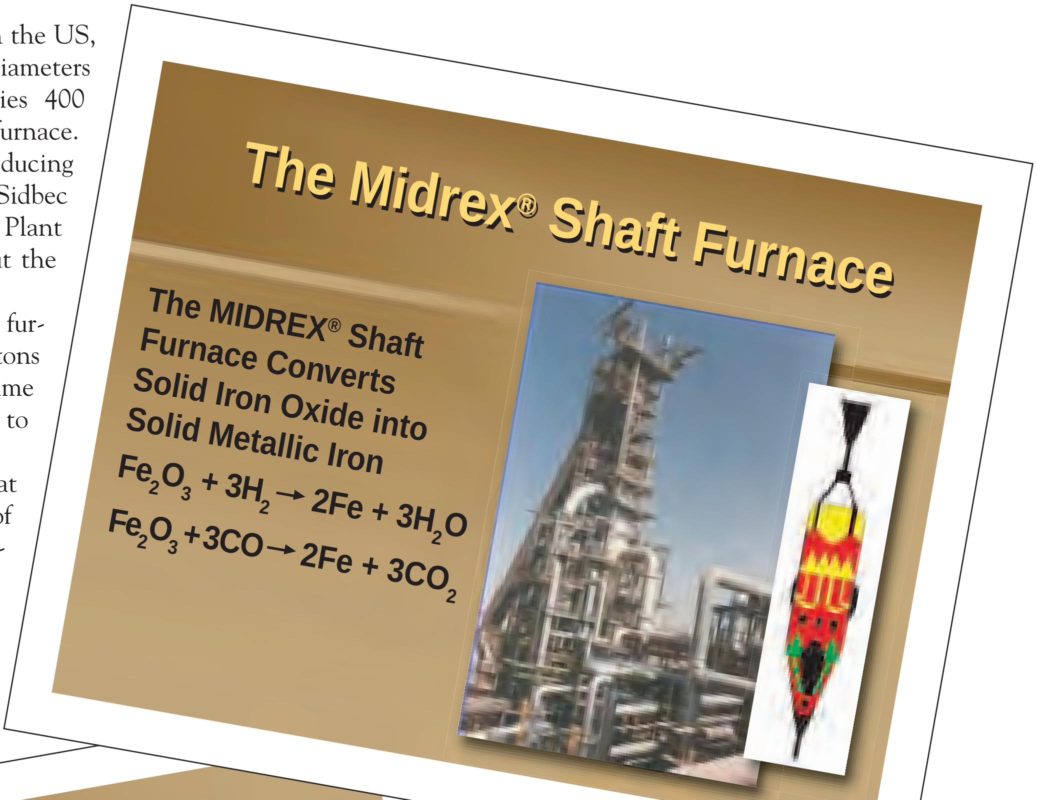
Larger shaft furnaces

Much of the success of the MIDREX Process has been due to the development and refinement of shaft furnace technology by Surface Combustion and later Midrex. Initial pilot plant tests began in the 1960s on a vessel with a diameter of 0.4 m. The first prototype commercial units at Portland had an inside diameter of

3.7 m. The next four modules installed in the US, Germany, Canada, and Argentina, had diameters of 4.9 m. After 1973, MIDREX® Series 400 Modules generally had a 5.0 m diameter furnace. The first 5.5 m furnace, capable of producing 600,000 t/y of DRI, began operations at Sidbec in 1977. The MIDREX MEGAMOD® Plant incorporates a 6.5 m furnace (more about the MEGAMOD in Part 3).

Today, Midrex typically designs shaft furnaces for a specific production rate of 10 tons per day cubic meter of reduction volume (t/d-m³). Midrex can provide facilities to produce from 300,000 t/y to over 2 Mt/y.

The MIDREX® Shaft Furnace operates at low pressure, which provides a number of benefits, including easy charging and discharging and the ability to use burden-feeders to facilitate material flow. These features have enabled MIDREX Plants to use a wide variety of iron ores, including lump ores.

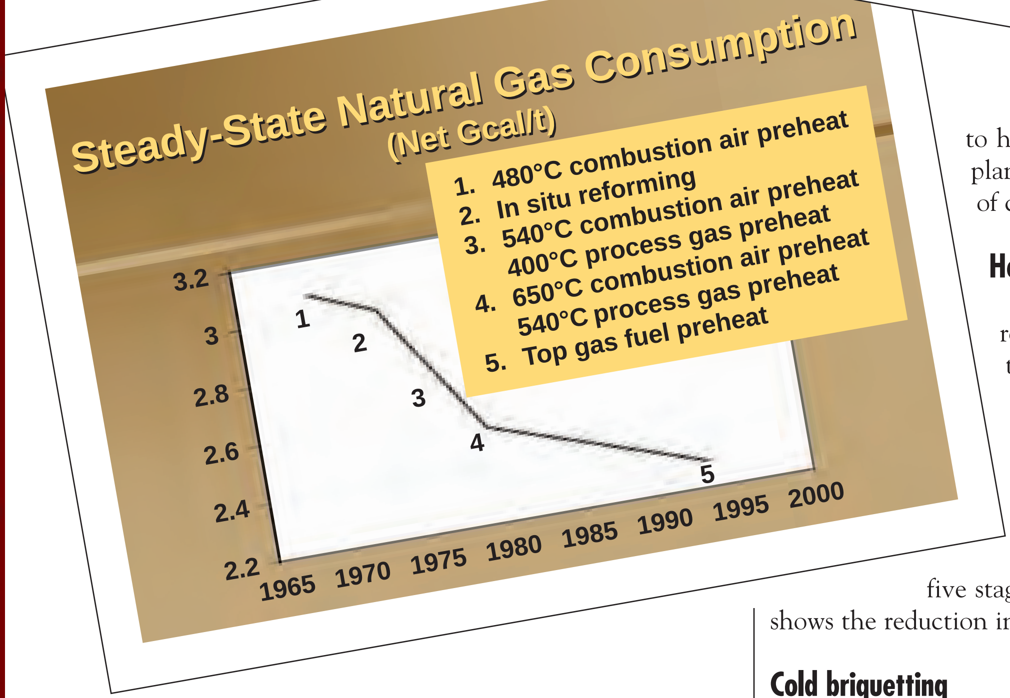


The Midrex® Shaft Furnace

The MIDREX® Shaft Furnace Converts Solid Iron Oxide into Solid Metallic Iron

$Fe_2O_3 + 3H_2 \rightarrow 2Fe + 3H_2O$

$Fe_2O_3 + 3CO \rightarrow 2Fe + 3CO_2$



Steady-State Natural Gas Consumption

(Net Gcal/t)

1. 480°C combustion air preheat
2. In situ reforming
3. 540°C combustion air preheat
400°C process gas preheat
4. 650°C combustion air preheat
540°C process gas preheat
5. Top gas fuel preheat

To date, MIDREX Plants have run on more than 25 types of pellets and more than 30 types of lump ores. Several MIDREX Plants operate with 50 percent and more lump ore on a routine basis.

In-situ reforming

Iron is an excellent catalyst for reforming natural gas. Thus, Midrex developed the idea of adding a small amount of natural gas to the furnace cooling zone, which is converted via reforming to hydrogen and carbon monoxide, increasing plant capacity. This also provides the benefit of carbon addition to the DRI.

Heat recovery

The first generation MIDREX Plants recovered a minimal amount of heat from the flue gas. The capital expenditure required for more extensive heat recovery was not justified because of the low cost of natural gas. However, soon afterward, the price of gas increased, and more heat recovery was added. At present, MIDREX Plants can be designed with five stages of heat recovery. The graph on the left shows the reduction in gas consumption over the years.

Cold briquetting

Direct reduction plants generate metallized fines, and if they are charged directly to the electric arc furnace, they are often lost in the slag and baghouse. To remedy this situation, plants began screening out fines and producing cold briquettes, which are more effectively charged to the EAF. Today, this is the most common means for recovering metallized fines.

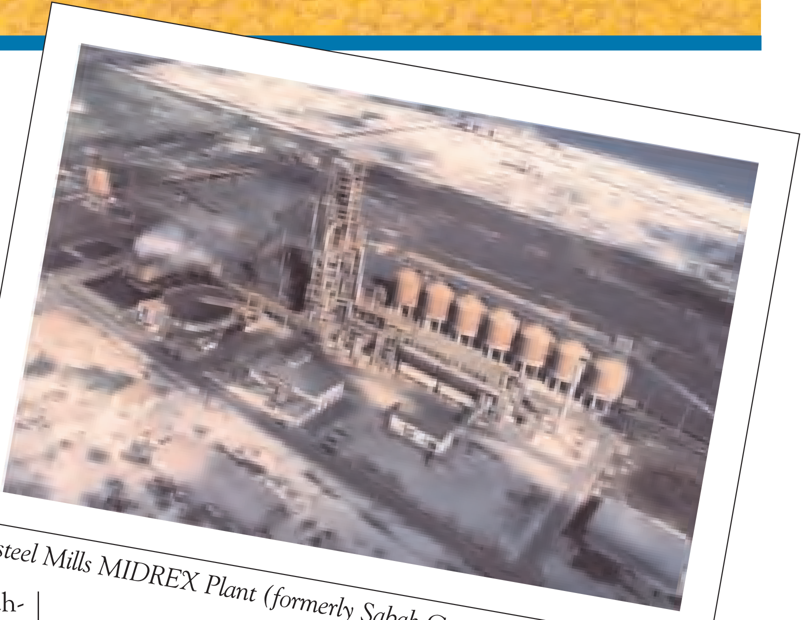
Part 3 in this series will discuss the sale of Midrex to Kobe Steel and the development of hot briquetting and the MIDREX MEGAMOD®.

A Better Mousetrap: The History of Midrex Technologies

Part 3

By John T. Kopfle.
Director - New Business Development

Note: Part 2 of this series reviewed Willy Korf's purchase of Midrex and the commercial success of the 1970s.



Amsteel Mills MIDREX Plant (formerly Sabah Gas Industries)

FROM ONE "K" TO ANOTHER

During the early 1980s, recession in the US caused a downturn in the metals industries worldwide, and world steel production stagnated at around 700 Mt/y from 1980 to 1987. Willy Korf's empire, built on a large amount of debt, began to crumble. By 1983, he was in bankruptcy and forced to sell off parts of his holdings. Among the interested buyers for Midrex was a major Japanese steel company. Kobe Steel had built the MIDREX® Plant in Qatar and was impressed with the technology and in 1984, Kobe purchased Midrex. The intent was not to utilize the technology in its operations, but to earn a profit for its Engineering & Machinery Division from selling plants.

MAKING DRI PORTABLE

In the early 1980s, the State of Sabah in Malaysia began developing an industrial complex on Labuan Island, near Borneo, to make use of extensive natural gas reserves. The complex was to include a methanol plant, power plant, and direct reduction facility. The government contracted with Midrex to build a plant to produce a new product, hot briquetted iron, for sale on the open market. Although some MIDREX Plants sold DRI, it is not the ideal merchant product because of its tendency to reoxidize and the handling problems it causes in meltshops not designed for its use. HBI, however, can be stored and handled like scrap, and melted using existing equipment.

The design of a hot briquetted iron plant presented many challenges, because it was necessary to eliminate the cooling zone, discharge the DRI hot from the shaft furnace, and briquette it. Midrex determined that a conical shaped bottom was the best approach. A flat bottom system with multiple mechanical discharge devices was ruled out because of concerns about high wear of the mechanical parts and frequent downtime due to buildup of material on the bottom of the furnace. High burden pressure at the bottom of the furnace has a tendency to compress the hot, malleable metallic DRI onto the bottom surface. In cases where multiple mechanical discharge devices would be used, there was a concern that failure of a mechanical unit would create a stagnant column of material in the shaft furnace, resulting in overheating and severe cluster formation due to prolonged exposure to the hot gases entering the shaft furnace.

The Sabah Plant started up in late 1984. It was built without a cooling gas inlet device, since cooling was not required. However, Midrex soon discovered that material flow problems were endemic with this design, and an "easy-flow" device, similar to the cooling gas inlet, was added. This worked well, and has become a standard part of MIDREX design.

Another difficulty was breaking the string of briquettes. They are produced in strings two wide, and thus must be broken in two dimensions. The first breaker design incorporated a star arrangement, which was not satisfactory. This was modified to the present hammer mill design.

With the success of the Sabah Gas Plant (now owned by Amsteel Mills), Midrex had developed a new technology.

There are now eight operating MIDREX HBI Modules The total accumulated production from these facilities is over 70 million tons.

DRI: AN IDEA WHOSE TIME HAS COME

By the late 1980s, the recession had ended, and world steel production began increasing. Continued mini-mill growth increased the "scrap intensity of steelmaking." This pressure on scrap supplies, and higher prices, increased demand for DRI.

In 1987, Midrex experienced a breakthrough with the signing of contracts for three new facilities, OPCO, VENPRECAR, and Essar Steel.

OPCO

In 1981, Midrex began discussions with CVG Ferrominera Orinoco of Venezuela about modifying an idled direct reduction facility in Venezuela. A feasibility study confirmed the viability of the project to produce HBI for merchant sale, and a contract was signed with Kobe Steel in December 1987 to execute the project. This was a unique job in several respects. First, it involved the use of existing steam reformers paired with a new MIDREX® Shaft Furnace. Second, the new shaft furnace was the first MEGAMOD® design, capable of one million tons per year. Third, Kobe Steel would lease the plant, modify it, operate it for a number of years, then turn it back to CVG. The plant, dubbed OPCO, started up in 1989.

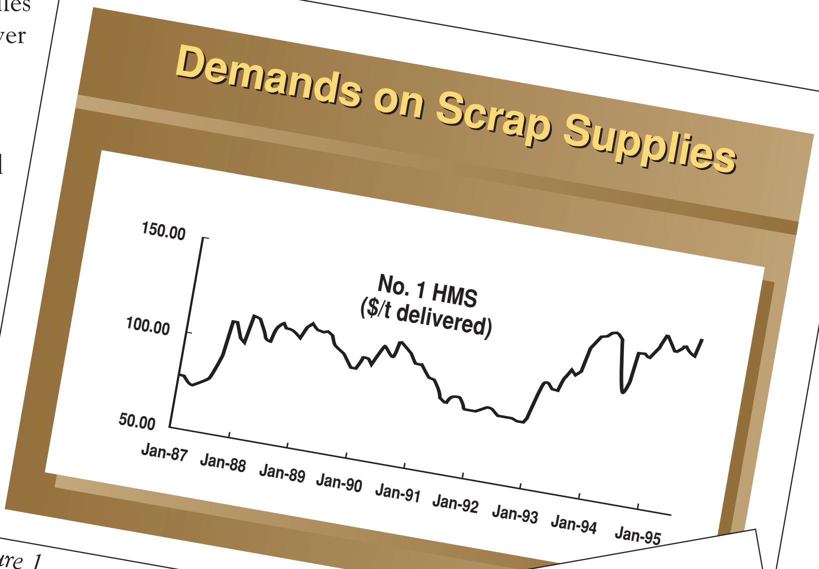


Figure 1

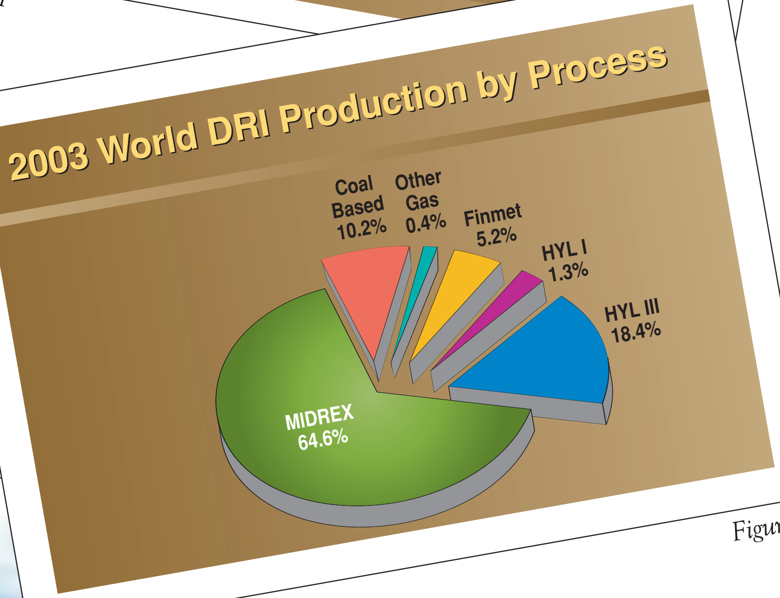


Figure 2

OPCO

VENPRECAR

Midrex signed a contract with SIVENSA, Venezuela's largest private steelmaker, to build a 600,000 t/y HBI plant. Start-up of the VENPRECAR Plant (originally known as SIDECAR) was in 1990.

Essar Steel

Midrex contracted with Essar Steel of India to relocate two idled MIDREX Plants from Germany to India and convert them to produce HBI. Start-up was in 1990.

Following upon the successful construction and start-up of these three facilities, Midrex entered a time of unprecedented demand for DR plants. This was driven by a number of factors:

- The US economic expansion of the 1990s, the longest in history
- The growth of mini-mills – in the US, EAF production grew 40% from 1985-1996
- The movement of EAF steel producers into higher quality products via innovations such as thin slab casting
- The need for clean charge materials to produce these grades of steel
- Increasing demands on scrap supplies and the resulting price increases – from January 1992 - January 1994, the price of No. 1 HMS rose from $84/ton to $135/ton, shown in *Figure 1*

As a result of this high DRI demand, there were 18 MIDREX Modules started up from 1992-2000, with a total capacity of nearly 14 million tons per year and an installed value of about $2.1 billion. Each year since 1987, MIDREX Plants have produced over 60 percent of the world's DRI, shown in Figure 2.

THE MEGAMOD

As early as 1976, Midrex had discussed the idea of a plant to produce one million tons per year. This would have major economic benefits, since the cost of a process plant does not increase linearly with capacity. Thus, larger plants have a lower capital cost per ton of output. The largest MIDREX Plant built prior to 1990 was the ANSDK facility in Egypt, with a rated capacity of 0.716 Mt/y, using a 5.5 m diameter furnace.

During the design of the OPCO Project, it was decided to incorporate a 6.5 m diameter shaft furnace to allow for the production of 0.83 Mt of HBI annually. Although the 6.5 m furnace is capable of producing over one million tons per year with a standard MIDREX Reformer, the original OPCO Plant was limited to 0.83 Mt/y because of the use of existing steam reformers. In 1996, a new "mini" MIDREX® Reformer was added, allowing the plant to produce 1 Mt/y.

Experience with the OPCO furnace was favorable and the design has proven to be very flexible with respect to iron ore use. The plant has produced over its rated capacity when required by market conditions, operating with up to 50 percent lump ore.

Eight MEGAMOD Shaft Furnaces have been built in Venezuela (2), India, Mexico, the United States, Trinidad, South Africa, and Korea.

EXPANDING THE ENVELOPE

In addition to hot briquetting and the MEGAMOD, Midrex continued to develop the technology to improve operating efficiency, reduce maintenance requirements, and improve the economic viability of the process.

Multiple reductant options

Midrex supplied shaft furnaces for two projects using COREX™ Offgas as the reductant - Saldanha Steel in South Africa and Hanbo Steel in Korea.

Lower energy consumption

State-of-the-art MIDREX Plants can incorporate up to five stages of heat recovery. MIDREX Plants consistently achieve the lowest natural gas consumption in the industry.

Flexibility in iron ore use

Midrex continued to investigate suitable iron ores, and design plants to incorporate the widest range of feed materials possible. The Saldanha furnace was designed to accept up to 70 percent lump ore.

Oxide Coating and Oxygen Injection

One of the limiting factors for production in MIDREX Plants is material sticking in the shaft furnace at high temperatures. The use of limestone or other coatings on the iron ore pellets and lump enables an increase in reduction temperature (through the use of oxygen injection to the reducing gas) without sticking.

REBIRTH OF COAL-BASED TECHNOLOGY

In the 1960s, the Midrex Division of Surface Combustion developed the Heat Fast Process for reducing iron ore using coal in a rotary hearth furnace. A two million dollar demonstration plant was built in Cooley, Minnesota, and 9,000 tons of the DRI was tested by the US Bureau of Mines in an experimental blast furnace in Bruceton, Pennsylvania. This was believed to be the best use for the product because its high sulfur content would limit use in electric furnaces. While the technical results of this trial were good, an economic analysis showed that there was no benefit in hot metal cost by using Heat Fast DRI versus the standard charge of 100 percent iron oxide pellets. Surface Combustion then stopped work on the process and concentrated on the gas-based MIDREX Process.

Several times during the next 25 years, especially when natural gas prices increased, Midrex considered resurrecting Heat Fast as a coal-based process.

Kakogawa Works in Japan

In 1989, the effort began in earnest and Kobe Steel, Ltd. (KSL) became a development partner. The process was renamed "FASTMET®." A 150 kg/h process simulator, with a 2.7 meter diameter hearth, was built at the Midrex Technical Center. Over 100 campaigns were conducted from 1992-94. Based on this success, KSL constructed a 8.5 meter diameter, 2.5 t/h demonstration plant at its Kakogawa Works in Japan. The facility started up in 1995 and was the world's first rotary hearth DR plant to make highly metallized DRI. During the late 1990s, Midrex and KSL embarked on a concerted effort to contract the first commercial FASTMET Plant.

The concluding part of this series will discuss the downturn and subsequent recovery of the world steel industry, the present strong market for MIDREX Plants, and the focus of technical and commercial developments for the future.

A Better Mousetrap: The History of Midrex Technologies Part 4

By John T. Kopfle.
Midrex Technologies, Inc.

Note: Part 3 of this series covered the purchase of Midrex by Kobe Steel, the strong direct reduction market of the 1990s, and Midrex's continuing technological improvements.

THE BUST

Steel cycles typically last five to ten years, and by the late 1990s, the strong run was coming to an end. In addition, the "Dot-com" boom, driven by the growth in Internet stocks, came to a crashing halt, and the US stock market plummeted. The combination of these factors resulted in a downturn in the metals industries beginning around 1997. The steel industry was included, and from January 1998 to January 1999, the prices of scrap and HBI dropped by about $50/t. By late 2001, selling prices for HBI were below production cost. (See Figure 1).

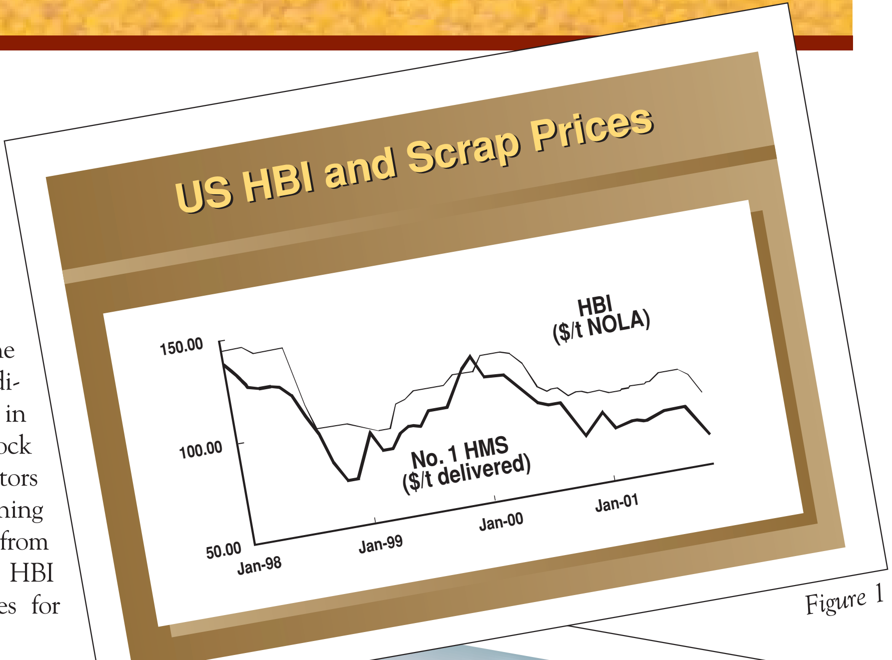


Figure 1

One constant in the direct reduction technology business is that demand for plants follows steel and DRI prices. With the downturn that began in 1998, direct reduction project development activity also dropped. Midrex, however, and its parent Kobe Steel realized that it was essential to develop new technologies to position the business for the inevitable recovery in three to five years. Thus, the significant development program that Midrex has maintained over the years was continued. Benefits of these enhancements include reduced capital and operating costs, lower energy consumption, and greater flexibility in operations. Among the areas of work were oxygen use, hot charging, process control, and coal-based technology.

Oxygen Use – MIDREX® Plant operators are continually developing methods to increase productivity. It has long been known that oxygen can provide for higher reducing gas temperatures and can increase production of H_2 and CO from methane. In the early 1990s, several MIDREX Plants, including SIDOR and Acindar, began experimenting with oxygen injection to the bustle gas. The higher temperature obtained was useful in overcoming the temperature drop in the shaft furnace caused by in-situ reforming. The net result was a productivity increase of up to 10 percent.

OEMK

At present, 19 MIDREX Plants employ oxygen injection. More details on oxygen use can be found in the article, "Tendencies in In-Situ Reforming" in the third quarter 2004 *Direct from Midrex*.

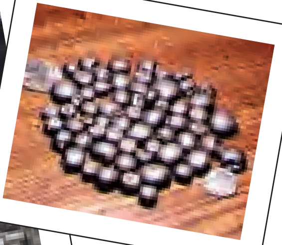
There is, however, a limit to oxygen injection because excessive bustle gas temperatures can cause clustering in the shaft furnace. Midrex has now developed the OXY+® System, which uses a partial oxidation reactor to generate additional reductants and avoid the problem of high bustle gas temperature. With OXY+, MIDREX Plants have the potential economically to increase productivity. The first OXY+ System is now being installed at OEMK in Russia.

Hot Charging – Nearly all captive MIDREX Plants (those with an adjacent steel mill) cool the DRI after it is discharged from the shaft furnace. The cooled DRI is stored and later charged into the EAF, where it is reheated and melted. An alternative is to feed hot DRI from the shaft furnace into the EAF and thereby utilize the DRI's sensible heat. This saves energy, increases productivity, and reduces electrode and refractory consumption. Some DR plants practice hot charging via use of transport containers and conveyors. This is a viable approach, but there can be problems with temperature loss, availability, and product degradation. To address these issues, Midrex developed the HOTLINK® System, in which the EAF is placed beneath the shaft furnace and hot DRI (HDRI) is charged by gravity.

HOTLINK delivers HDRI to the EAF at 700-750° C. It also has the flexibility to produce any combination of HDRI and DRI. For more details, see the fourth quarter 2003 *Direct from Midrex*. Midrex is now negotiating a contract for the first HOTLINK System.

Process Control – There has been a tremendous evolution of process control technology during the 35 years since start-up of the first MIDREX Plant. Midrex, in cooperation with Siemens, has now developed SIMPAX®, a suite of control software for MIDREX Plants that will provide enhanced optimization and equipment protection. The software runs on its own separate workstation linked to the existing Basic Automation System. This output can be used to determine the most efficient mode of operation and will be interlocked to the existing control system to provide additional equipment protection. The benefits of SIMPAX include higher product quality and enhanced process control in real time.


Kakogawa Works


ITmk3 Nuggets


Mesabi Nugget Facility

Coal-based Technology – As mentioned in part 3 of the "Better Mousetrap" series, in 1989 Midrex and Kobe Steel began development of the coal-based FASTMET® Process. A 150 kg/h process simulator was built and operated at the Midrex Technical Center and a 2.5 t/h demonstration plant was constructed at KSL's Kakogawa Works in Japan. The first FASTMET facility was sold to Nippon Steel and started up in 2000. This 190,000 t/y plant processes steel mill waste at the company's Hirohata Works, producing DRI that is fed to a BOF. KSL also built a 14,000 t/y plant at its Kakogawa Works for recycling zinc-bearing wastes. This facility started up in 2001. With these successes, FASTMET has become the leading rotary hearth reduction technology.

In 1996, KSL began experimenting with the melting of FASTMET-type pellets containing iron ore and carbon. A unique phenomenon was discovered, that if the temperature was raised to about 1,450° C after reducing the iron ore, the iron and slag separate, resulting in a nugget of nearly pure iron plus carbon. This process was named ITmk3, for "ironmaking technology mark three," with Mark I being blast furnace ironmaking and Mark II natural gas-based direct reduction. The ITmk3 reaction occurs in the solid/liquid co-existence phase of the iron-carbon diagram, which is different than traditional ironmaking processes. ITmk3 promises to be a revolutionary process, enabling production of a premium grade iron product without use of coke. It offers the following benefits: reduction and slag separation in one step, low process temperature, minimal FeO refractory attack, clean slag separation, and the ability to use low grade iron ores and wastes.

Based on these promising results, tests were conducted on the Midrex FASTMET Process Simulator. KSL then built a four-meter diameter pilot plant at the Kakogawa Works and tested the process in 1999 and 2000. These results were also successful, and an effort was begun to site a demonstration plant. Ultimately, a partnership of the State of Minnesota, KSL, Ferrometrics, Cleveland-Cliffs, and Steel Dynamics was formed. A 25,000 t/y facility, named Mesabi Nugget, was built in Silver Bay, Minnesota, USA, in the "Iron Range." Start-up was in 2003 and several campaigns were conducted through the summer of 2004 to test the process at larger scale and confirm operating parameters.

The partners are now developing the first commercial plant.

The China Juggernaut

Although there were many "gloom and doom" predictions about steel and DRI prices early in the new millennium, the steel cycle performed as always, and a recovery began in early 2002. The major factor was China, whose development created tremendous demand for all commodities, including steel and DRI.

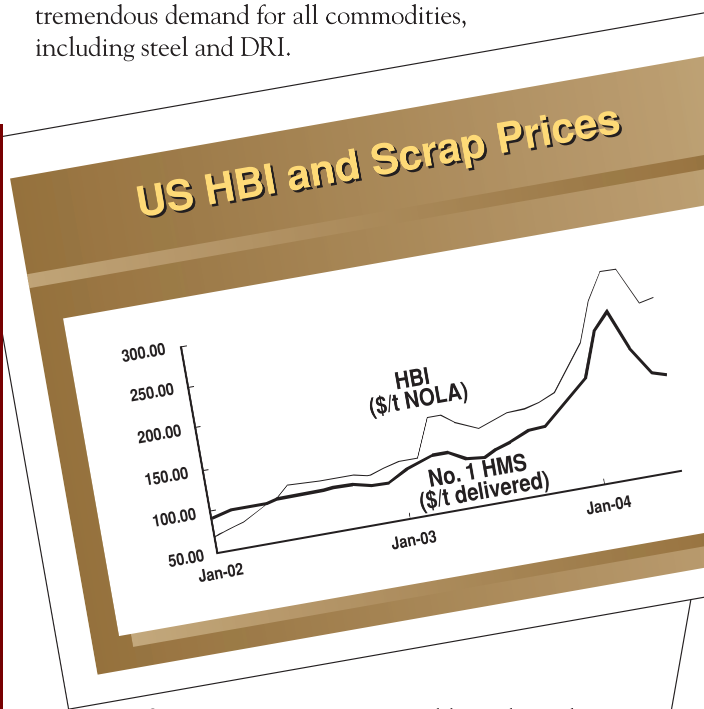


Figure 2

In 2003, world steel production reached an all-time high of 950 million tons. Due to strong Chinese demand, declines in the value of the dollar, metallics supply restraints, and high freight rates, prices of metallics skyrocketed, as shown in Figure 2. In early 2004, merchant HBI prices reached $275/t FOB Venezuela. Prices dipped in May and June of 2004, but subsequently returned to the $300/t level and remain there as of November.

As happened in the 1990s, these high metallics prices spurred demand for direct reduction facilities, and Midrex has contracts for plants in Trinidad, Saudi Arabia, and Malaysia, and is negotiating with potential clients in the Middle East and Russia. These projects contain a number of unique features, including plant relocation, HOTLINK, hot charging, and oxygen use and will further advance the MIDREX Technology and experience base.

Despite the strong DR market now, steel cycles will continue and there will be downturns in the future. To help offset those periods, Midrex is engaged in an effort to identify and pursue opportunities in areas other than iron and steel. The intent is to apply the company's skills in engineering, procurement, and construction management to promising process technologies in areas such as non-ferrous metals, energy, and environmental. The strategy is to obtain rights to proprietary technologies via licenses or partnerships. Midrex hopes to partner with several technology owners over the next few years.

A Storied Past, an Exciting Future

This concludes the remarkable story of the men and women who made Midrex the leading direct reduction technology company. We salute all of them, from the early researchers who worked on four direct reduction concepts before finding success, through Donald Beggs, the "Father" of the MIDREX Process, to Willy Korf, the visionary who grew the company, to the countless men and women at Midrex, Kobe Steel, MIDREX Process Licensees and partners who have dedicated their careers to direct reduction. The technology is a tremendous success: since 1969, MIDREX Plants have produced over 400 million tons of DRI and they routinely exceed design capacity. Each year since 1987, MIDREX Plants have produced over 60 percent of the world's DRI.

What does the future hold? Given the strong steel and metallics markets, prospects for sales of MIDREX Plants are outstanding for the next several years. Midrex realizes, however, that prosperity is never assured, and we must continue refining and enhancing the MIDREX Process to ensure that it remains the leading gas-based reduction technology.

Susumu Okushima, Kobe Steel, pictured with James D. McClaskey, Midrex Technologies, Inc., celebrating the 20th anniversary of Kobe Steel's purchase of Midrex

Development and commercialization of coal-based technology by Midrex and Kobe Steel continues, and this should be a major contributor in coming years. Diversification into non-steel areas is promising and will expand in the future.

As we stated in the first quarter 2004 *Direct from Midrex* commentary, Midrex has a storied past, and we look forward to an exciting future.

MIDREX® Processes

Masaaki ATSUSHI, Hiroshi UEMURA, Takashi SAKAGUCHI
Plant Engineering Department, Iron Unit Division, Natural Resources & Engineering Business

Since 1978, when a plant based on the MIDREX process was built in Qatar for producing direct reduced iron, Kobe Steel and MIDREX Technologies, Inc., have collaborated to make many technical improvements in the process. The largest MIDREX module, having an annual production capacity of 1.8 million tones, began operation in 2007. The MIDREX module, together with a melt shop, now has a production capacity comparable to that of a blast furnace. This paper presents an overview of the history of the technical developments in these processes, as well as the latest developments in this field.

Introduction

MIDREX direct reduction ironmaking (hereinafter referred to as the "MIDREX process") reduces iron ore using natural gas. The original process was developed by the Midland-Ross Co., which later became MIDREX Technologies, Inc. (hereinafter referred to as "MIDREX Technologies"), a wholly owned subsidary of Kobe Steel. A pilot plant was built in Toledo, Ohio in 1967. The first commercial plant, having a production capacity of 150 thousand tonnes/year, was built in Portland, Oregon, in 1969.

The process was immature in 1978, when Kobe Steel began the construction of a plant with a production capacity of 400 thousand tonnes/year in the State of Qatar. Kobe Steel significantly modified the design, exploiting the company's technologies developed through blast furnace operation, and stabilized the then new process. On the other hand, MIDREX Technologies also carried out various improvements to the plants they built in various countries. These were all integrated in the early 1980s, making the process nearly complete 1).

The maximum production capacity in 1984, when Kobe Steel became affiliated with MIDREX Technologies, was 600 thousand tonnes/year. Later improvements, made by Kobe Steel in collaboration with MIDREX Technologies, have dramatically increased the production capacity. In 2007, the scale reached 1.8 million tonnes/year, which is comparable to that of a small blast furnace.

1. Characteristics of reduced iron

The MIDREX process produces direct reduced iron (hereinafter referred to as "DRI"). The process reduces iron ore using a reforming gas made from natural gas. The DRI is used mainly as the raw material for electric arc furnaces (EAFs), as a clean iron source substitute for scrap iron.

Pores are left behind in the DRI after oxygen has been removed. These pores, if filled with water, for example, can cause the iron to reoxidize with ambient oxygen, generate heat and occasionally ignite a fire. This makes it difficult to transport the product by ship or to store it in the open air over an extended period of time. To resolve this issue, Kobe Steel developed a technology for compacting DRI into briquette iron at a temperature of around 700℃. DRI has an apparent density of 3.4 to 3.6t/m^3, while the briquette iron has an apparent density of 5.0 to 5.5t/m^3.

The reoxidation issue had restricted DRI manufacturing sites to the vicinity of steelmaking plants. The hot briquette technology has eliminated this site restriction, making it possible to build a reduced ironmaking plant where resources such as natural gas, iron ore and power are less costly. The product, hot briquette iron (hereinafter referred to as "HBI"), can be exported by sea to steelmaking plants and rolling mills in other countries. This has expanded the number of potential sites for MIDREX plants all over the world 1).

Table 1 compares the chemical and physical properties of DRI and HBI, while **Fig. 1** shows the appearance of DRI and HBI.

The global production of DRI increased dramatically from 790 thousand tonnes/year in 1970 to 68.45 million tonnes in 2008. DRI made by the

Table 1 Specification of DRI and HBI

	DRI	HBI
Fe total（%）	90〜94	←
Fe metallic（%）	83〜89	←
Metallization（%）	92〜95	←
Carbon（%）	1.0〜3.5	←
P*（%）	0.005〜0.09	←
S*（%）	0.001〜0.03	←
Gang*（%）	2.8〜6.0	←
Mn, Cu, Ni, Mo, Sn Pb and Zn（%）	trace	←
Bulk density（t/m^3）	1.6〜1.9	←
Apparent density（t/m^3）	3.4〜3.6	5.0〜5.5
Discharge temperature（℃）	40	80

* depends on components of iron ore

Fig. 1 Appearance of DRI and HBI

58 modules operating & 4 modules under construction in 19 countries.
Total capacity of MIDREX Process=48.4 million ton/y

Fig. 2 World's MIDREX plants

MIDREX process accounts for about 60% of global production.

Fig. 2 shows the worldwide locations of MIDREX plants.

2. MIDREX process

Fig. 3 is a flow chart for the MIDREX process. Either lump ore, or pellets prepared for direct reduction ironmaking, are charged as raw material from the top of a shaft furnace. The ore is reduced inside the furnace and the reduced iron is discharged from the bottom of the furnace. Reductant gas blown in from about the middle of the shaft furnace reduces the raw material above the nozzle and escapes from the top of the furnace. The cooling gas, which circulates in the lower portion of the furnace, cools the DRI. Both the charging and discharging ports are dynamically sealed by a sealing gas, allowing the continuous charging of raw material and discharging of DRI.

The reaction occurring in the shaft furnace is the well-known reduction reaction of iron, described as follows:

$Fe_2O_3 + 3CO \rightarrow 2Fe + 3CO_2$

$Fe_2O_3 + 3H_2 \rightarrow 2Fe + 3H_2O$

The exhaust gas (top gas) emitted from the top of the shaft furnace is cleaned and cooled by a wet scrubber (top gas scrubber) and recirculated for reuse. The top gas containing CO_2 and H_2O is pressurized by a compressor, mixed with natural gas, preheated and fed into a reformer furnace. The reformer furnace is provided with several hundreds of reformer tubes filled with nickel catalyst. Passing through these tubes, the mixture of top gas and natural gas is reformed to produce reductant gas consisting of carbon monoxide and hydrogen. The reaction that occurs in the reformer tubes is as follows:

$CH_4 + CO_2 \rightarrow 2CO + 2H_2$

$CH_4 + H_2O \rightarrow CO + 3H_2$

$2CH_4 + O_2 \rightarrow 2CO + 4H_2$

$CO + H_2O \rightarrow CO_2 + H_2$

$CH_4 \rightarrow C(S) + 2H_2$

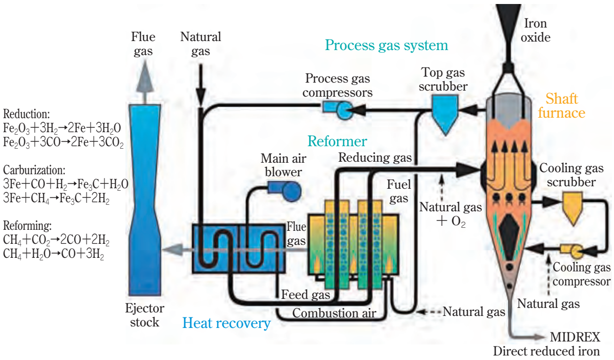


Fig. 3 MIDREX process flow sheet

3. History of the development of the MIDREX process

3.1 Operation of MEGAMOD® shaft furnace: Raw material coating (1990 -)

There was an urgent need to upsize the shaft furnace in response to the market need for an increased production capacity. To achieve this, Kobe Steel and MIDREX Technologies began development by

- conducting analyses using the three-dimensional finite element method,
- conducting two-dimensional model experiments for verification and
- improving raw material characteristics on the basis of reduction/pulverization tests.

As a result, the shaft diameter was increased to 5.5m and then to 6.5m (MEGAMOD shaft furnace). This has increased the production capacity from the previous maximum of less than 400 thousand tonnes/year, first to 800 thousand tonnes/year, and then to 1.5 million tonnes/year [1).]

A technology was devised to raise the temperature of reducing gas (bustle gas) by coating the raw material with lime hydrate which has a melting point higher than that of DRI. This has raised the reducing gas temperature to about 900℃ and improved shaft furnace productivity by more than 10%.

3.2 Oxygen injection into reducing gas (2000 -)

Injecting high purity oxygen into the hot reducing gas has further raised the reducing gas temperature to about 1,000℃ (**Fig. 4**). Although a portion of hydrogen and carbon monoxide is consumed by combustion with oxygen, raising the temperature of the reducing gas has improved shaft furnace productivity by 10 to 20%[2), 3)].

3.3 Improvement of oxygen injection technology (2005 -)

The oxygen injection, described above, has evolved into an improved technology, called "OXY＋®", which was made possible by the introduction of a partial combustion technique. As shown in **Fig. 5**, the OXY＋ employs a combustor in addition to the reformer. The combustor partially burns natural gas and oxygen to produce hydrogen and carbon monoxide, which are added to the reducing gas generated by the reformer[2), 3)].

Fig. 6 shows the transition of shaft furnace productivity.

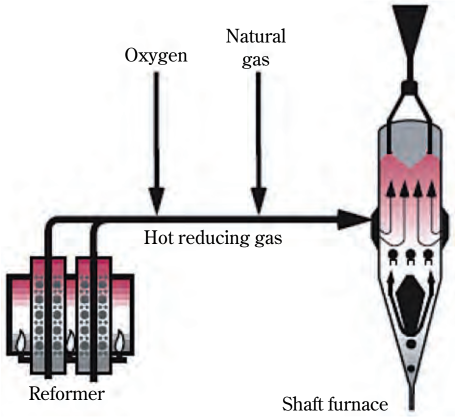


Fig. 4 Oxygen injection flow

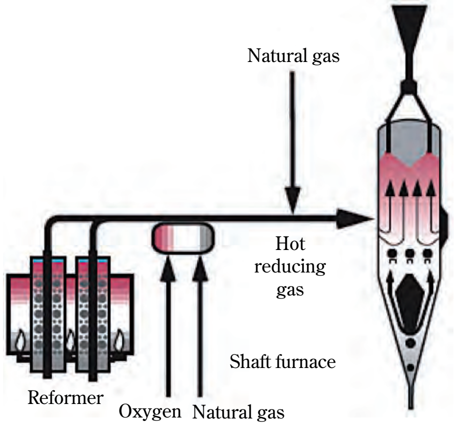


Fig. 5 OXY＋ flow

3.4 Development of a shaft furnace, SUPER MEGAMOD®, and enhancement of engineering (2007-)

The experience of operating the shaft furnace with a diameter of 6.5m has led to the construction of a larger shaft furnace at Saudi Iron & Steel Company in Hadeed, Saudi Arabia in 2007. This shaft furnace has a diameter of 7.15m and an increased production capacity of 1.8 million tonnes/year (**Fig. 7**).

Another shaft furnace, SUPER MEGAMOD, currently under development, is to have a further increased production capacity in the range of 2 million tonnes/year. The increased size of the shaft furnace enlarges the entire facility, which requires even more sophisticated design and construction

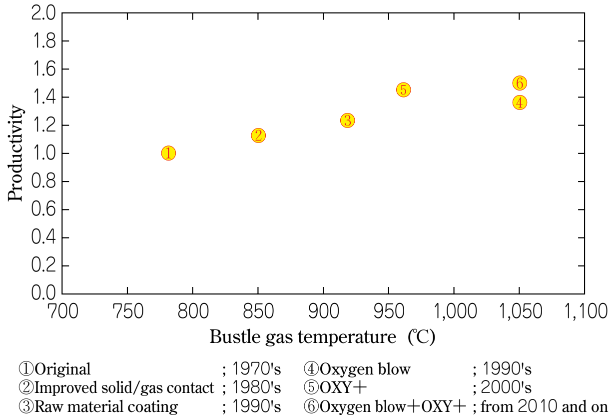


Fig. 6 Changes in productivity of MIDREX shaft furnace

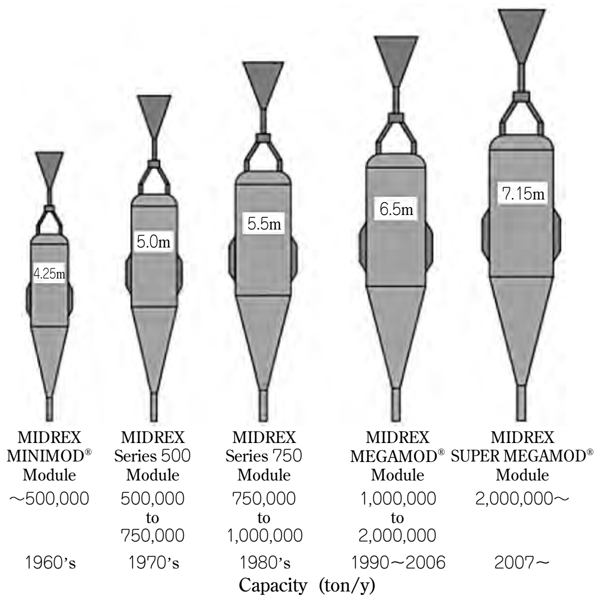


Fig. 7 Changes in shaft furnace diameter and annual production

management. Because of this, since 2004, a three-dimensional CAD has been adopted for the entire designing process. The three-dimensional CAD allows the retrieval of structural calculation data, as well as the direct output of isometric drawings of piping and material spreadsheets. The three-dimensional CAD is also utilized at construction sites for planning construction schedules. **Fig. 8** is a three-dimensional CAD drawing showing an entire DR Plant for Qatar Steel.

3.5 Delivery record

Table 2 shows the delivery record of MIDREX plants. The following is an outline of the major plants.

3.5.1 LION plant

This plant with a rated capacity of 1.5 million tonnes/year was constructed for the Lion Group, Malaysia, and began operations in 2007 (**Fig. 9**)4). It produces two types of products, hot DRI (HDRI) and HBI. The HDRI is supplied as hot metal to a neighboring EAF facility by a hot transport vessel. The HBI is mainly exported and is occasionally used by the neighboring EAF facility.

3.5.2 HADEED Module-E plant

This is the world's largest MIDREX plant. Constructed at the Saudi Iron & Steel Company in Hadeed, Saudi Arabia, it began operations in 2007 (**Fig.10**)4). The plant was the first to adopt a shaft

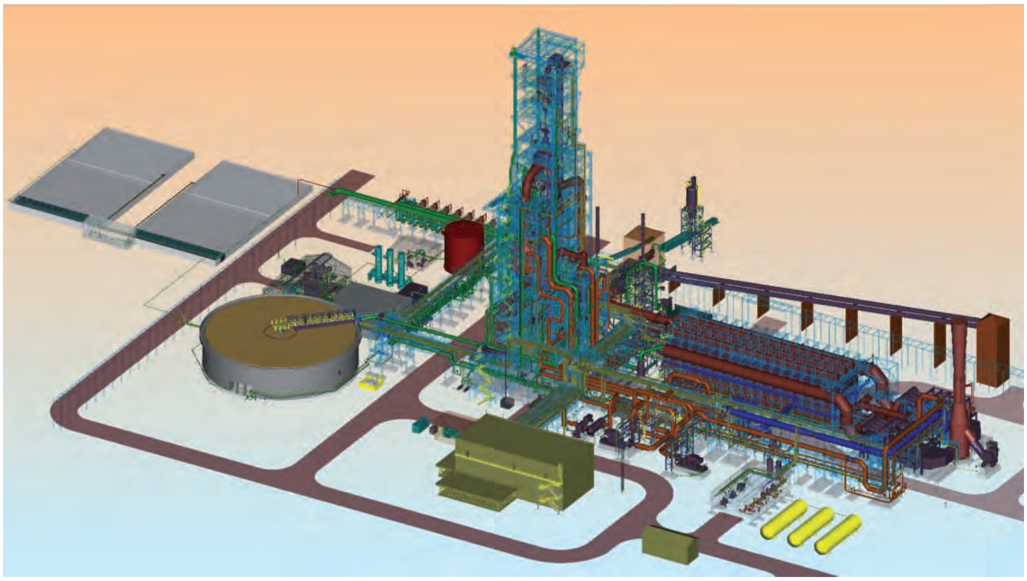


Fig. 8 CAD drawing of QASCO Module-II plant

Table 2 Recent delivery record of MIDREX plants

	Plant	Location	Capacity (million tont/y)	Start up
*	EZDK III	Egypt	0.8	2000
	Essar Steel Module-IV	India	1.0	2004
	Nu-Iron	Trinidad	1.6	2006
	Essar Steel Module-V	India	1.5	2007
	HADEED Module-E	Saudi Arabia	1.76	2007
*	QASCO Module-II	Qatar	1.5	2007
	LGOK Module-II	Russia	1.4	2007
	Al-Tuwairqi Damman	Saudi Arabia	1.0	2007
	LION	Malaysia	1.54	2008
	Essar Steel Module-VI	India	1.8	2009
*	SHADEED	Oman	1.5	2010
	ESISCO	Egypt	1.76	2010
	Al-Tuwairqi Pakistan	Pakistan	1.28	2010

*: Kobe Steel constructed

Fig. 9 LION plant

furnace having a diameter of 7.15m and has a rated capacity of 1.8 million tonnes/year. This capacity is comparable to that of a small blast furnace. The plant produces both HDRI and DRI. The HDRI that it produces is supplied as hot metal directly to a neighboring EAF facility, being transferred by a hot transport conveyor. The DRI is stored temporarily in a silo and supplied to the neighboring EAF facility as necessary.

3.5.3 QASCO Module-Ⅱ

This plant with a rated capacity of 1.5 million tonnes/year was constructed by Kobe Steel at Qatar Steel Company in the State of Qatar and began to operate in 2007 (**Fig.11**). It is to be noted that in 1975 Kobe Steel delivered a plant, Module-I, with a rated capacity of 400 thousand tonnes/year, to Qatar Steel Company. The design and operational improvements that Kobe Steel made on the MIDREX process, which was still immature at the time, have stabilized operations. This plant has won high acclaim and led to an order for Module-Ⅱ.

The Module-Ⅱ plant produces both DRI and HBI. The DRI is supplied to a neighboring steelmaking plant, while the HBI is exported. The DRI is melted at the steelmaking plant and is supplied to a rolling mill that produces billets, rebars and wire rod coils to be exported.

Fig.10 HADEED Module-E plant

Fig.11 QASCO Module-II plant

Fig.12 SHADEED plant

3.5.4 SHADEED plant

In 2008, Kobe Steel completed construction of a plant with a capacity of 1.5 million tonnes/year at SHADEED Iron & Steel Co. in Oman (**Fig.12**). The plant, which is to produce DRI and HBI, is currently in the preparation stage. It is the first plant to adopt a method called HOTLINK® for supplying HDRI to an adjacent EAF by gravity.

4. Recent technological trend

4.1 Hot discharge of DRI

Conventionally, DRI was cooled before being discharged from the shaft furnace. Technical modifications are being implemented to discharge hot DRI (HDRI) in order to improve the specific energy consumption and productivity of the plant, including the downstream steelmaking process. A combination of two discharge methods, cold and hot, was proposed and implemented to allow flexibility in production planning, which improves productivity[3), 5)].

Fig.13 depicts the overall flow in an integrated steel mill equipped with a MIDREX plant. The following three methods (**Fig.14**) allow the transfer of HDRI from the shaft furnace to the downstream steelmaking plant:

a) transfer and supply by a hot transport vessel (Fig.14-①)
b) transfer and supply by a hot transport conveyor (Fig.14-②) and
c) supply by gravity (HOTLINK) (Fig.14-③ and **Fig.15**).

The HDRI discharge methods have been adopted by various plants as summarized in **Table 3**.

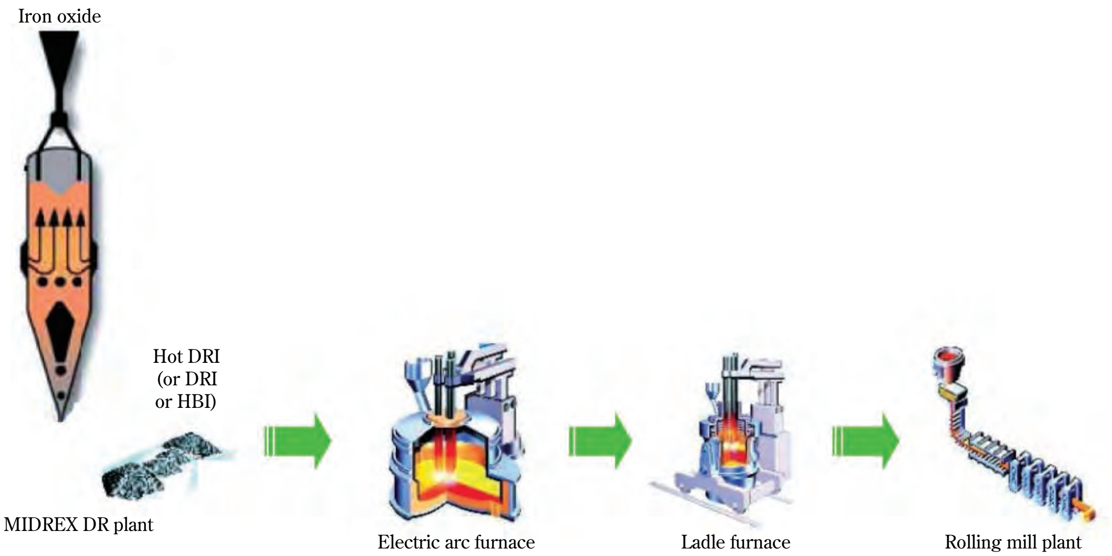


Fig.13 Overall flow sheet for integrated steel mill equipped with MIDREX plant

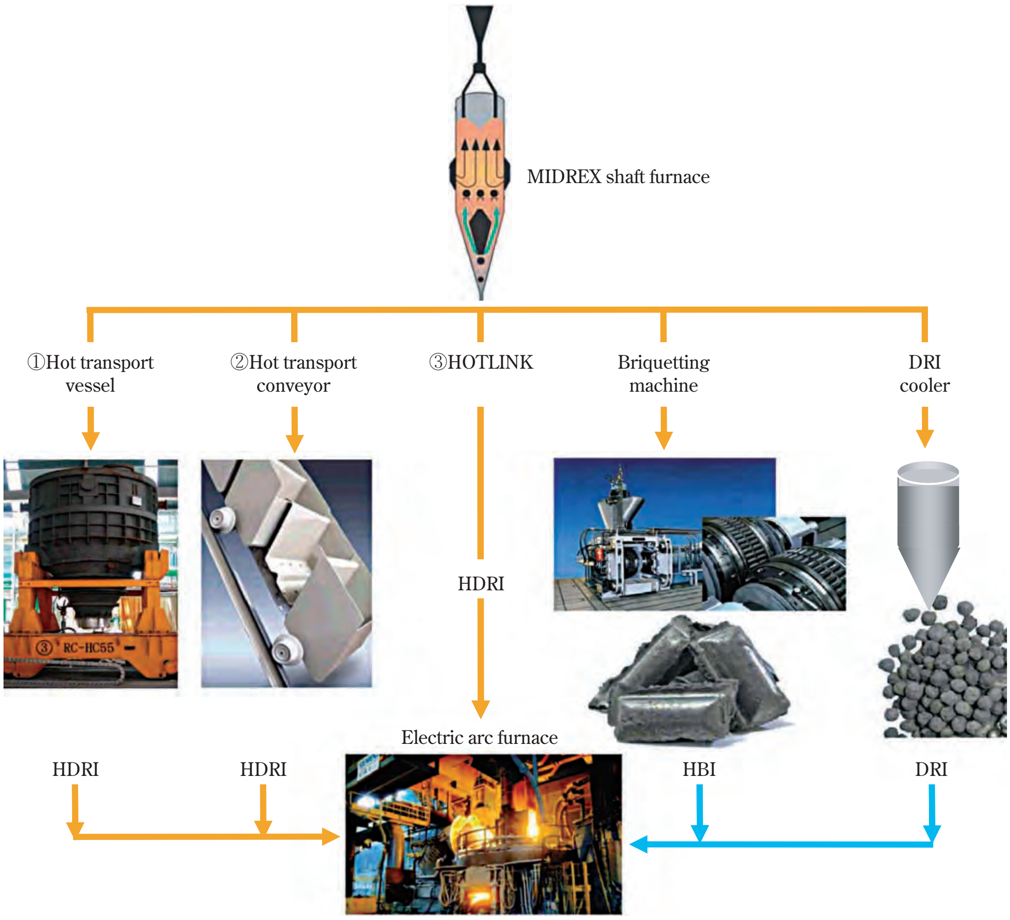


Fig.14 Variation of discharging products

4.2 Improving specific energy consumption and productivity using HDRI

Supplying HDRI at an elevated temperature directly to a steelmaking plant significantly improves the specific energy consumption and productivity of the plant. As shown in **Fig.16**, raising the HDRI supply temperature saves power consumed by the electric arc furnace (EAF). In addition, this power saving reduces the consumption of the EAF's electrode, which decreases the operational cost (**Fig.17**).

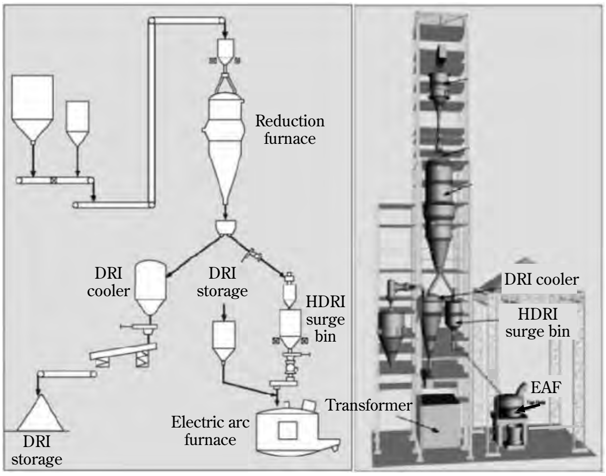


Fig.15 Material flow of HOTLINK and equipment arrange

Table 3 MIDREX plants discharging HDRI

PLANT	LOCATION	START-UP	TYPE SYSTEM
Essar steel Module-I, II, III, IV	India	1999～2004	Hot transport vessel
HADEED Module-E	Saudi Arabia	2007	Hot transport conveyor
LION	Malaysia	2008	Hot transport vessel
ESISCO	Egypt	2010	HOTLINK
SHADEED	Oman	2010	HOTLINK

Furthermore, charging HDRI to an EAF shortens the cycle time of the EAF, which increases the production volume by 10 to 15%.

4.3 CO_2 emission reduction

Various improvements have been made to the MIDREX process to reduce the specific energy consumption of the process, including downstream steelmaking, and to improve the productivity of the shaft furnace. These energy-saving measures not only decrease the operational cost, but also decrease the environmental burden with reduced emissions of CO_2 and other types of exhaust.

The MIDREX process, which is based on natural gas, emits intrinsically less CO_2 than other processes using coal. Because of this, the MIDREX process can also contribute to emission reduction in coal based ironmaking processes. For example, charging HBI produced by a MIDREX plant into a blast furnace reduces CO_2 emissions as a whole.

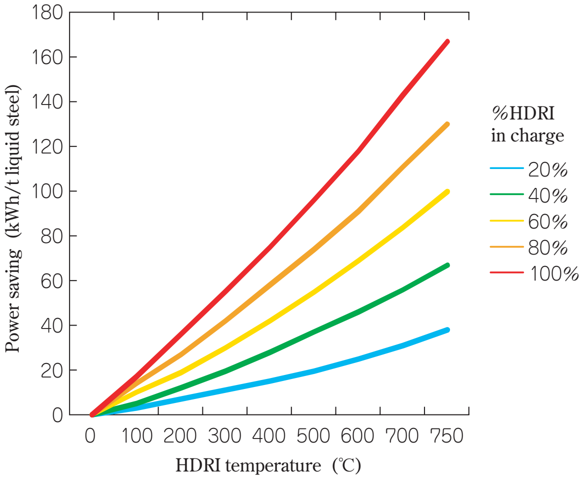


Fig.16 Correlation between HDRI temperature and power savings at EAF

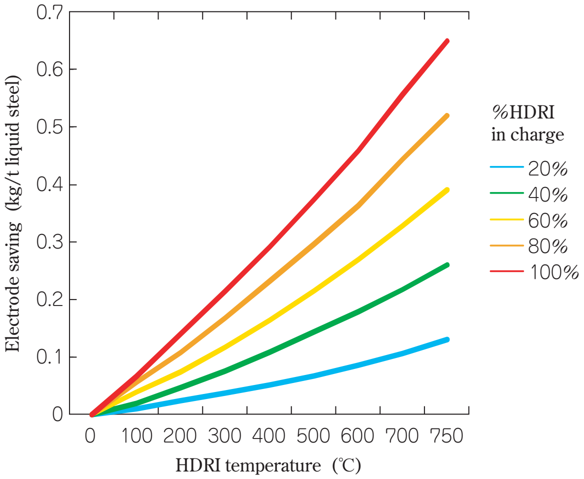


Fig.17 Correlation between HDRI temperature and electrode savings at EAF

4.4 Combination with coal-based fuel

The MIDREX process can utilize not only the reducing gas modified from natural gas, but also coke oven gas and other reducing gases derived from PET coke or from bottom oil generated in oil refineries. Thus the construction of MIDREX plants, formerly restricted to sites in natural gas producing countries, no longer suffers from such limitations. For example, the MIDREX process can be incorporated into a blast furnace based ironmaking facility that has a coking process. The HBI produced by using the coke oven gas can be charged into the blast furnace to decrease the reduction load of the blast furnace. This will decrease the ratio of the reductant used as a heat source (reductant ratio) and reduce CO_2 emissions.

Fig.18 depicts the process flow of the MIDREX process combined with a gasification plant.

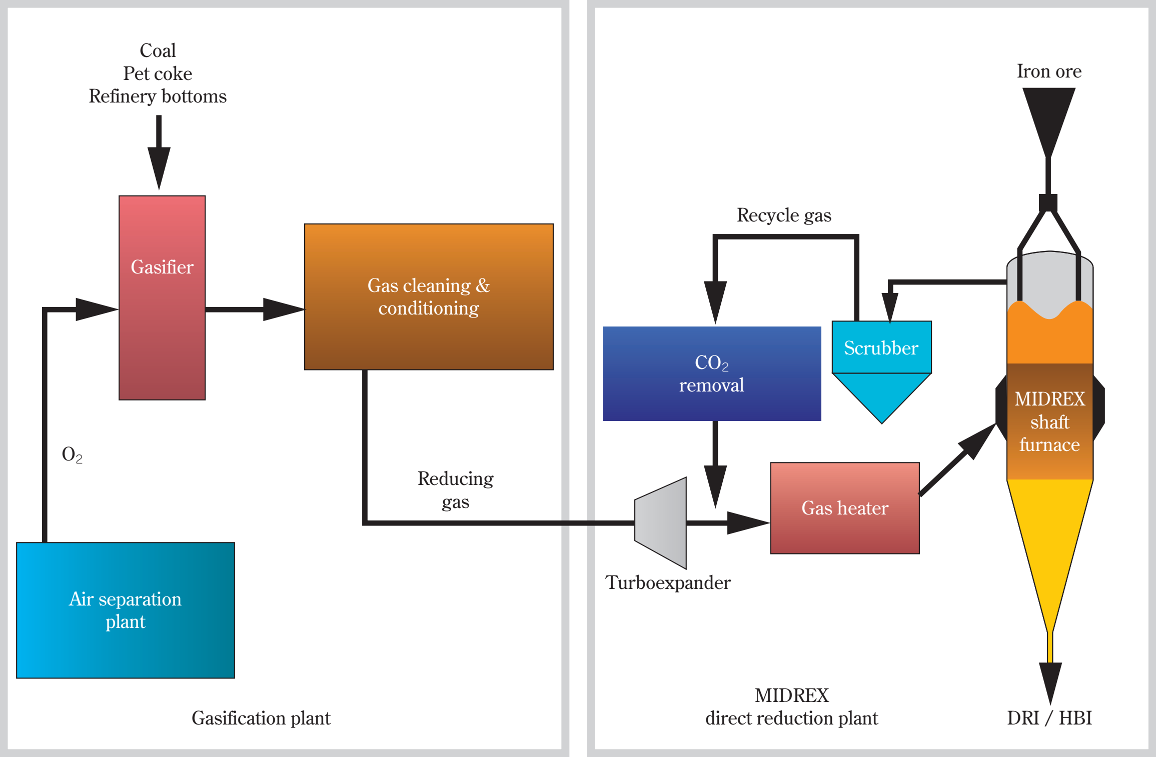


Fig.18 Process flow of MIDREX process combined with coal gasification plant

Conclusions

The origin and development of the MIDREX process have been introduced along with the new technologies that have been developed or are presently being developed by Kobe Steel.

Since the inauguration of the first commercial plant in 1969, seventy-two MIDREX plants have been built in twenty-one countries so far. The MIDREX process occupies a market share of about 60% among DRI making plants. This is a result of the improved reliability of the MIDREX process, as well as improved process efficiency, which is widely recognized and highly evaluated.

Kobe Steel will continue striving to decrease the environmental burden, increase the versatility of raw materials and further improve efficiency so as to contribute to the world's iron and steel production.

References

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ENERGIRON HYL

THE INNOVATIVE DIRECT REDUCTION TECHNOLOGY

Energiron
The most competitive way to obtain the highest quality raw material for steelmaking

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Our commitment

Danieli & C. is excited to have partnered with Techint and HYL, through an exclusive agreement, to bring the industry the Energiron brand. Energiron, the innovative direct reduction technology from HYL, is the alliance which joins our mutual capabilities and experiences to provide the leading technology for DR plants at the most competitive cost.
Danieli brings its extensive experience and tradition of excellence in engineering and plant making to form what we are sure is becoming a key player in the iron and steel industry today. We are committed to making that a reality.

Gianpietro Benedetti
Chairman and CEO of Danieli

DANIELI

By uniting the now half-century heritage of the HYL technology with the longstanding expertise of both Techint and Danieli, we can now offer the direct reduction and steelmaking industry their strongest options ever. Techint is not only one of the world's major suppliers of iron and steelmaking plant technology and equipment through Tenova, but is also one of the largest DRI-based steelmakers in the world through its Tenaris and Ternium companies.
The Energiron alliance provides the most advanced, cost effective and environmentally friendly DR technology on the market, which produces the highest quality iron ore feed material for modern, high quality steelmaking.
We have always led the field in direct reduction innovation, and we will continue with Energiron to do so in the future. That is our commitment to you.

Gianluigi Nova
CEO Tenova

tenova

Energiron at a glance

Tenova, a company of the Techint group, and HYL have joined with Danieli to form the Energiron alliance which allows the three companies to merge their know-how and technology for the design and construction of Gas Based Direct Reduction Plants under the new Energiron trademark.

Energiron is as strong as the name suggests and the unique product of this technology is more than just direct reduced iron - it s Energy combined with Iron - for the lowest cost and highest quality steelmaking applications.

Within the Energiron alliance, in which each party contributes to the continuous technology development, Danieli plays the role of exclusive main contractor for the engineering and installation of the plants world wide, whilst HYL brings its wide experience in process design and plant operation.

World steel demand is continuously increasing, both in total values and in the share of specialty steels. The Energiron alliance responds to the need for a premium raw material for competitive and quality clean steel production.

Steel production cost is a combination of raw material, energy and labor cost. There are areas where DRI production is naturally more convenient due to the availability of energy and iron ores at low prices. Energiron plant configurations can be designed to be competitive even in those areas where Natural Gas is not economically available, by using other sources of reducing gas, such as Coke Oven Gas (COG), Coal Gasification Gas (SYNGAS) and others.

Energiron DRI is a highly metallized product with controllable carbon content in the range of 0.8 to 5% which generates chemical energy in the EAF melting process.

In DRI-based integrated minimills the use of the HYTEMP▫ system allows the delivery of hot DRI directly to the EAF, thus further reducing electric energy consumption and tap to tap time.

Thanks to Energiron unique passivation characteristic the production of DRI can also be devoted to overseas sales, safely and easily even without briquetting.

Danieli and Tenova technical and commercial network

Danieli Headquarters in Buttrio, Italy

Tenova Headquarters in Milan, Italy

Energiron at a glance


DANIELI FAR EAST
DANIELI
DANIELI Metallurgical Equipment(Beijing)Co.,Ltd.
达涅利冶金设备（北京）有限公司

Pioneering achievements

1957
Startup of the first commercially successful gas-based direct reduction plant, Hylsa 1M, using the HYL Process.
Production of flat products via the EAF, based on the use of DRI.

1958
Batch charging of DRI to the EAF at 600¡C.

1965
Use of more than 30% DRI in an EAF charge, eventually increasing in stages up to 100% by 1972.

1968
Continuous feeding of DRI to the EAF.
Computerized EAF process control system put into use.

1969
Use of Foamy Slag practices.

1970
Design of pellets for direct reduction.
First full scale testing and use of DRI as a BF-BOF feed.

1972
Production of extra-deep drawing steels in EAF using DRI.

1980
Start up of the HYL Process continuous shaft furnace in Monterrey.

1984
HYL carries out first pilot operation of a direct reduction plant without external gas reformer.

1988
Use of coating of pellet/lump ores for direct reduction.

1993
HYTEMP pneumatic transportation system and hot DRI feeding to the EAF.
First Zero kWh direct reduction plant begins operation at Vikram Ispat-Grasim HYL III plant, India.

1994
HYL begins producing high carbon DRI with 3.0 to 5% carbon content.

1995
Production of ultra thin (<1mm) hot rolled coils based on 100% DRI, with HYL plant and Hylsa CSP minimill.

1997
World s first dual-discharge (DRI and HBI) plant design put into operation, Vikram Ispat-Grasim HYL plant, India.

1998
Startup of first commercial scale HYL ZR Process plant, Hylsa 4M, Monterrey.
Hot and cold DRI charging to the world s largest twin-cathode DC EAF.

1999
HYL III Plant with water producing option starts up at Hadeed, KSA.

2000
First plant to successfully use 100% lump ore charge on a routine basis (Usiba HYL plant, Brazil). Vikram Ispat-Grasim becomes second in 2006.

2001
Successful economical design of Micro-Module (200,000 tpy) plants based on HYL ZR reformerless technology.

2003
Successful design of HYL ZR process plants based on coal gasification and COG.

2005
First coal gasification-based Energiron DRI plant for Jindal Steel, India. This 2 Mtpy module is the largest plant of this kind ever to be built in the world.

2006
Energiron: Techint, Tenova HYL and Danieli team up to promote and develop the new technology for direct reduction plants.

First 1.4-Mtpy integrated DRI-based minimill complex for GHC Group, including a 1.6 Mtpy Energiron DR plant with HYTEMP System.

First HYL Micro-Module plant built, Al Nasser Industries, Abu Dhabi, UAE.


MATESI

Technology highlights

Process alternatives
The uniquely flexible ZR process scheme lets you choose the best energy source - whether it s with natural gas/steam reformer, the direct use of natural gas, or whether your conditions are best satisfied using Syngas from a coal gasifier or even Coke Oven Gas.

Product options
This is what you re after in the first place. Why settle for less than the best product for your needs? We offer the choice of highly metallized DRI with, carbon ranging from 0.8 to 5% and either cold or hot, in the form of HBI or HYTEMP iron for direct feeding to the EAF shop, and you can get there from the widest range of iron ore pellets and lump ores in the industry, including high sulfur ores.

Plant modules
Energiron plants are available in different modular sizes to meet the needs of the typical steel mill — either with the Micro-Module (200,000 tpy), or any sized plant up to 2.0 million tpy and larger.
Due to peculiar product passivation the quantity of cold DRi produced in excess can be easily sold on the market as commodity.
The Energiron plant may be self-sufficient in terms of electric energy consumption and water requirement.
It is also possible to sell CO_2 as a by-product, too.

Low environmental impact
Energiron plants are ecologically friendly and meet the most stringent environmental regulations. Process characteristics enable the lowest possible emissions of CO_2 and NO_X, and CO_2 can even be captured and sold as a lucrative by-product.

Low maintenance and high availability
The possible elimination of the reformer and the use of standard equipment of smaller size throughout the plant significantly reduces the cost of maintenance.
Further, Energiron plants are controlled by a highly automated system providing level 2 automation, requiring minimal human intervention and providing high plant availability and reliability.

Project flexibility
To bring it all together, we offer the flexibility of providing anything from the basic technology package to local equipment supply and erection up to full turnkey projects.
Our support runs from process technology through operations and maintenance training and technical assistance, not only in the DR plant but in the steel mill as well. Further, Tenova and Danieli can provide complete steel mill projects covering materials handling, direct reduction, EAF meltshop, casting and rolling and product finishing lines.

Main features of energiron technology

Product

Product flexibility:
Cold DRI, Hot DRI (HYTEMP¤ and HBI) in every modality.
The product quality can be adjusted to best fit the client needs.

Product quality:
High metallization, high carbon DRI.
Energiron is the only DR technology currently capable of producing high carbon DRI with more than 90% of the carbon as iron carbide.
Unique passivation characteristic due to high iron carbide content.

Low environmental impact

Low dust carry-over determined by the low gas velocity into the shaft furnace.

High efficiency of the selected dedusting system.

Low suspended solid content of water.

Exhaust gas re-utilization as burner gas (no gas burned in flare system) greatly reduces the environmental impact.

Incorporates CO_2 recovery system.

Low NO_x emissions due to high efficiency of the thermal equipment.

Process in brief

The process converts lump ore, iron oxide pellets or pellet / lump ore mixtures into highly metallized, stable iron product.

The most common types of iron ores have the composition of hematite (Fe_2O_3) and contain about 30% of oxygen. In the Energiron process this chemically bonded oxygen is removed by means of a reducing gas mix at high temperature.

The reducing gas is a mix of carbon monoxide (CO), hydrogen (H_2) and methane (CH_4). These gases react with the oxygen contained in iron ore giving carbon dioxide (CO_2) and steam (H_2O(g)) as products.

The reducing gas mix extracts from iron oxide the chemically bonded oxygen. The overall reduction reactions are:
(1) $Fe_2O_3 + 3\ H_2 \longrightarrow 2\ Fe + 3\ H_2O$
(2) $Fe_2O_3 + 3\ CO \longrightarrow 2\ Fe + 3\ CO_2$
moreover there is an additional reduction given by the following reaction:
(3) 3Fe¡ + $CH_4 \longrightarrow Fe_3C + 2H_2$

Reaction (3) is part of the in situ reforming reactions. The high carbon DRI produced with the Energiron plant has a very high metallization and a controlled carbon content (between 0.8 and 5%), according to EAF requirements, in the form of Fe_3C.

The process

Energiron is state-of-the-art technology for the production of premium quality DRI

General process scheme

The Energiron Direct Reduction Process is designed to convert iron ore (pellet/lump) into metallic iron by the use of reducing gases in a solid-gas moving bed shaft furnace. Oxygen is removed from the iron ore by chemical reactions based on hydrogen (H_2) and carbon monoxide (CO), for the production of highly metallized DRI.

The technology offers the flexibility to produce three different product forms, depending on the specific requirements of each user.

Reducing gases can be generated:

-directly, by in-situ reforming of natural gas inside the shaft furnace,

-in an external natural gas/steam reformer,

-from gasification of fossil fuels, biomass, etc., as Syngas

-from Coke Oven Gas (COG) sources.

In all the above cases, the process configuration corresponds to the same basic ZR scheme, adjusting the relative sizes of equipment for the particular application.

Both for the in-situ and steam reforming schemes, natural gas analysis (heavy hydro carbons content) is not a limiting factor.

Direct use of natural gas

The ZR Process is a major step in decreasing the size and improving the efficiency of direct reduction plants. Reducing gases are generated in-situ in the reduction shaft, by feeding natural gas as make-up to the reducing gas circuit. Since the reducing gases are generated in the reduction section, optimum reduction efficiency is attained, and thus an external reducing gas reformer is not required. Therefore the overall energy efficiency of the ZR process is optimized by the in-situ reforming inside the shaft furnace, since the product takes most of the energy supplied to the process, with minimum energy losses to the environment. As compared to other processes for which the overall efficiency is bellow 80%, for this scheme the efficiency is around 86%.

The impact on plant size of eliminating the external gas reformer is significant. For example, a plant of 1-Mtpy capacity requires only 60% of the area needed by other process plants for the same capacity. For additional capacity, the area required is proportionally smaller in comparison. This also facilitates locating the DR plant adjacent to the meltshop in existing operations. This plant configuration has been operated successfully since 1998 with the HYL DR 4M plant, and also was incorporated in 2001 to the 3M5 plant, both at the Ternium Hylsa steel facility in Monterrey, Mexico.

A remarkable advantage of this process scheme is the wider flexibility for DRI carburization, which allows attaining carbon levels up to 5%. This is due to the improved carburizing potential of the gases inside the shaft, which allow for the production primarily of iron carbide.

Reformed gas, Syngas, COG

Of course, Energiron plants can also use the conventional steam-natural gas reforming equipment, which has long characterized the process. Other reducing agents such as hydrogen, gases from coal, pet coke and similar fossil fuels gasifications, and coke oven gas, among others, are also potential sources of reducing gas, depending on the particular situation and availability.

This flexibility is made possible precisely because the Energiron ZR Process is independent of the reducing gas source, with no requirement to recirculate gases back to a reformer to complete the process chemistry loop.

Several projects are currently under development which will use coke oven gas as the reducing gas source, and projects using gas from coal gasification technology are also underway.

Reduction

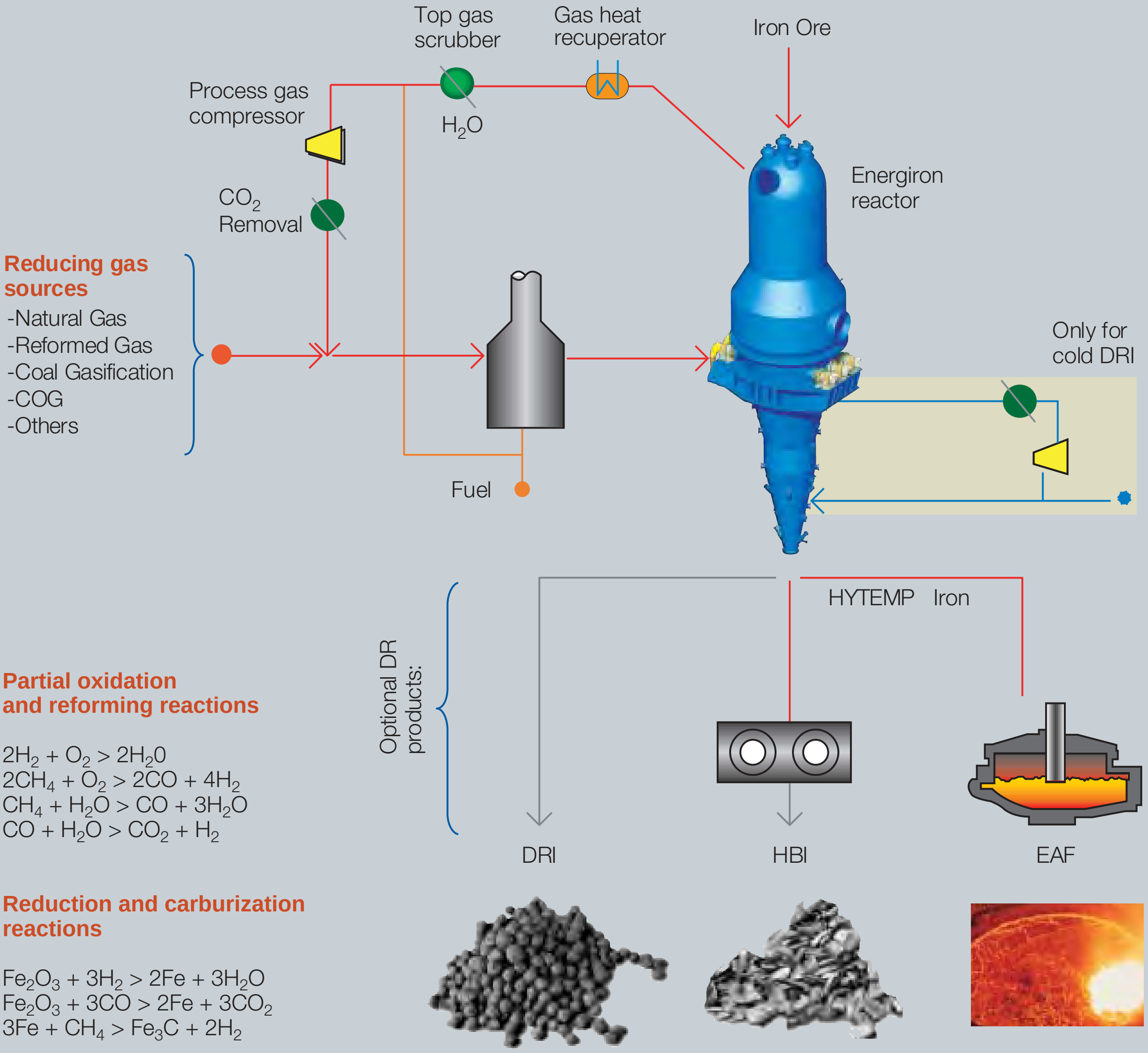
Inside the shaft furnace vessel, hot reducing gas is fed to the reduction zone and flows upward counter-current to the iron ore moving bed.

The gas distribution is uniform and there is a high degree of direct contact between gas and solids.

The exhaust reducing gas (top gas) leaves the reactor at about 400¡ C and passes through the top gas heat recuperator, where its energy is recovered to produce steam, or alternatively to preheat the reducing gas stream, and then through the quenching/ scrubbing system.

In these units, water produced during

Energiron direct reduction process



Partial oxidation and reforming reactions

$2H_2 + O_2 > 2H_20$
$2CH_4 + O_2 > 2CO + 4H_2$
$CH_4 + H_2O > CO + 3H_2O$
$CO + H_2O > CO_2 + H_2$

Reduction and carburization reactions

$Fe_2O_3 + 3H_2 > 2Fe + 3H_2O$
$Fe_2O_3 + 3CO > 2Fe + 3CO_2$
$3Fe + CH_4 > Fe_3C + 2H_2$

The process

the reduction process is condensed and removed from the gas stream and most of the dust carried with the gas is also separated. Scrubbed gas is then passed through the process gas recycle compressor, where its pressure is increased. Compressed gas, after being sent to the carbon dioxide removal unit, is mixed with the natural gas make-up, thus closing the reducing gas circuit.

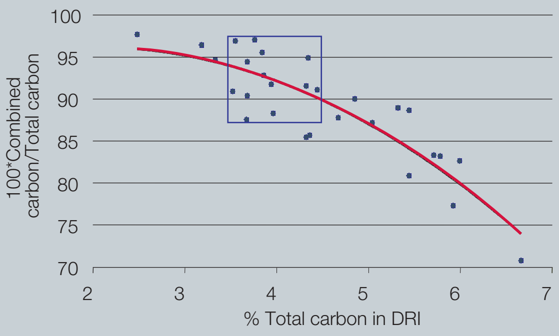
The furnace operates at a pressure of around 6 bar absolute allowing a high reactor productivity of about 9 t/h per m^2 and minimizing dust losses through top gas carry-over. This is reflected in low iron ore consumption, which keeps the operating cost low.

Removal of oxygen from the iron ore is accomplished by the action of the hot reducing gases and then the product is carburized.
A rotary valve, located at the bottom of the vessel, regulates the continuous gravity flow of the charge downward through the reduction furnace. DRI is discharged by automated mechanisms, consisting of pressurized bins and pressure locks.
For cold DRI, a cooling gas is fed to the lower conical part of the furnace at about 40¡C, flowing upward countercurrent to the DRI moving bed. The gas distribution is uniform and there is a high degree of direct contact between the gas and solid, without physical restrictions to the flow of solids or gases inside the unit.
The cooling gas exits from the upper conical part, at about 460¡C, and is then quenched/scrubbed by means of cooling water. Scrubbed cooling gas passes through the cooling gas recycle compressor to be recycled to the furnace, after being made-up with natural gas. Natural gas is injected as make-up to the cooling gas circuit for optimum efficiency and control of the cooling and carburization processes.
For hot product discharge and use, the cooling circuit is eliminated and hot DRI is continuously discharged at >700¡C. For the HYTEMP pneumatic transport system, the product is transported by means of a carrier gas to the surge bins located at the meltshop, for a controlled feeding to the electric arc furnace. For production of HBI, Hot DRI is continuously discharged at >700¡C to the hot briquetting machines arranged below.
The HBI is cooled in vibrating cooling conveyors using cooling water and then discharged to the HBI transport conveyor.

One of the inherent characteristics of the Energiron process scheme and of high importance environmentally is the selective elimination of both by-products generated from the reduction process; water (H_2O) and carbon dioxide (CO_2), which are eliminated through top gas scrubbing and CO_2 removal systems, respectively.

High carbon DRI - iron carbide
A unique benefit of the ZR process is the DRI that it produces. This product has typically a metallization of 95% and a carbon content of around 4% in the form of combined carbon. This type of product yields significant benefits in the electric furnace that no other process to date has been able to achieve.
Carbon in the DRI, mostly as iron carbide (Fe_3 C) is derived mainly from methane (CH_4) and to a lesser extent from CO. The level of carbon is adjusted by controlling the reducing gas composition and/or oxygen injection.
DRI produced with the ZR scheme is characterized by higher stability than DRI typically obtained in other DR process schemes. The reason for this is the high cementite or Fe_3C content, which inhibits the re-oxidation of metallic iron in contact with air. For a carbon content of 4%, approximately 95% is present as Fe_3C. In general, every 1% of combined carbon corresponds to 13.5% of Fe_3C. Therefore, a DRI with 4% Carbon contains more than 50% of Fe_3C.
The high percentage of Fe_3C in the DRI makes the product very stable.

High carbide iron (HCI)



Optimum range of combined carbon (cementite or iron carbide - Fe_3C) in DRI from Energiron process

DRI stability

Extensive tests were conducted to determine whether the combined carbon in DRI was a factor in improving product stability over that of conventional DRI, whether produced by HYL plants or other process technologies. In general, High-Carbon DRI is more stable than conventional DRI. This has been proven in specific tests that were performed for DRI being produced at the Ternium Hylsa 3M5 plant, before and after its conversion to the ZR Process scheme. Since stability itself is not measured but rather reactivity of the product in three distinct environments, tests were conducted in the presence of air, air plus water and air plus salt water to determine which type of DRI is more stable. These test results are shown here graphically, indicating that reactivity to air, water and salt water is significantly less over time for DRI containing high percentages of carbon as iron carbide.

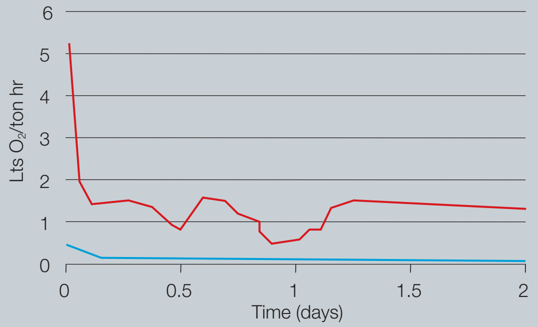
Currently, all or most direct reduction technologies carry out reduction reactions at very high temperatures—from 920 to 950 ¡C. The high reduction temperature is one of the factors determining the DRI stability. Nevertheless a more significant factor is the carbon content of DRI. DRI which has high levels of carbon content in the form of iron carbide or cementite forms a quality of DRI pellet which, while still requiring proper handling procedures, is much safer than ever before.

The HYTEMP® system

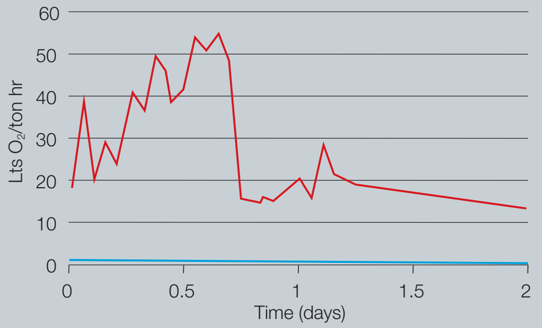
An additional technology which, on its own provides significant benefits for steelmakers, is the HYTEMP pneumatic transport system for sending hot DRI from the reduction reactor to the EAF shop. When combined with the Energiron ZR Process, the benefits increase substantially by bringing hot,

DRI reactivity in contact with air, water, and salty water

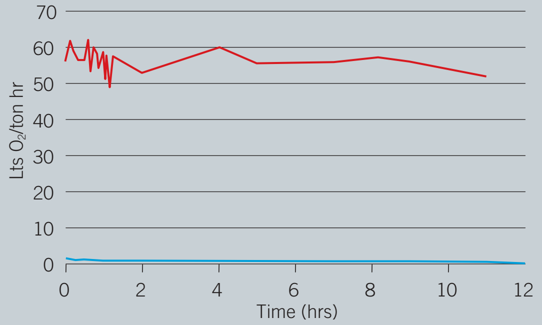
Reactivity of DRI in contact with air



Reactivity of DRI in contact with air+water



Reactivity of DRI in contact with salty water+air



— HYL ZR — Standard DRI

HYL DRI characteristics: DRI -ZR Mtz=94% C=4% (>90% as Fe_3C)

Standard DRI Mtz=94% C=2%

high carbide iron to directly feed the melting furnaces.

The HYTEMP System involves an Energiron hot discharge direct reduction reactor connected to an adjacent electric furnace mill by means of a pneumatic transport system. HYTEMP iron is DRI produced at high temperature (700 ¡C) with metallization up to 95% and controlled carbon usually around 4%, and which is pneumatically transported from the reactor discharge to the meltshop for direct feeding to the EAF. In this manner, the energy value of the hot DRI is capitalized in the EAF.

This process scheme offers the most adequate arrangement for integrated steelmaking facilities due to the important benefits capitalized in the EAF. Hot DRI is sent to the meltshop, where it is temporarily stored in insulated inert storage bins, for feeding to the furnace by continuous injection mechanisms, which deposit the material directly in the metallic bath surface.

Currently the Ternium Hylsa Monterrey 4M plant produces hot-discharge DRI, using the HYTEMP System for hot DRI transport to the meltshop (cold DRI is also produced via an external cooler if required).

A simplified scheme of the HYTEMP system is shown graphically.

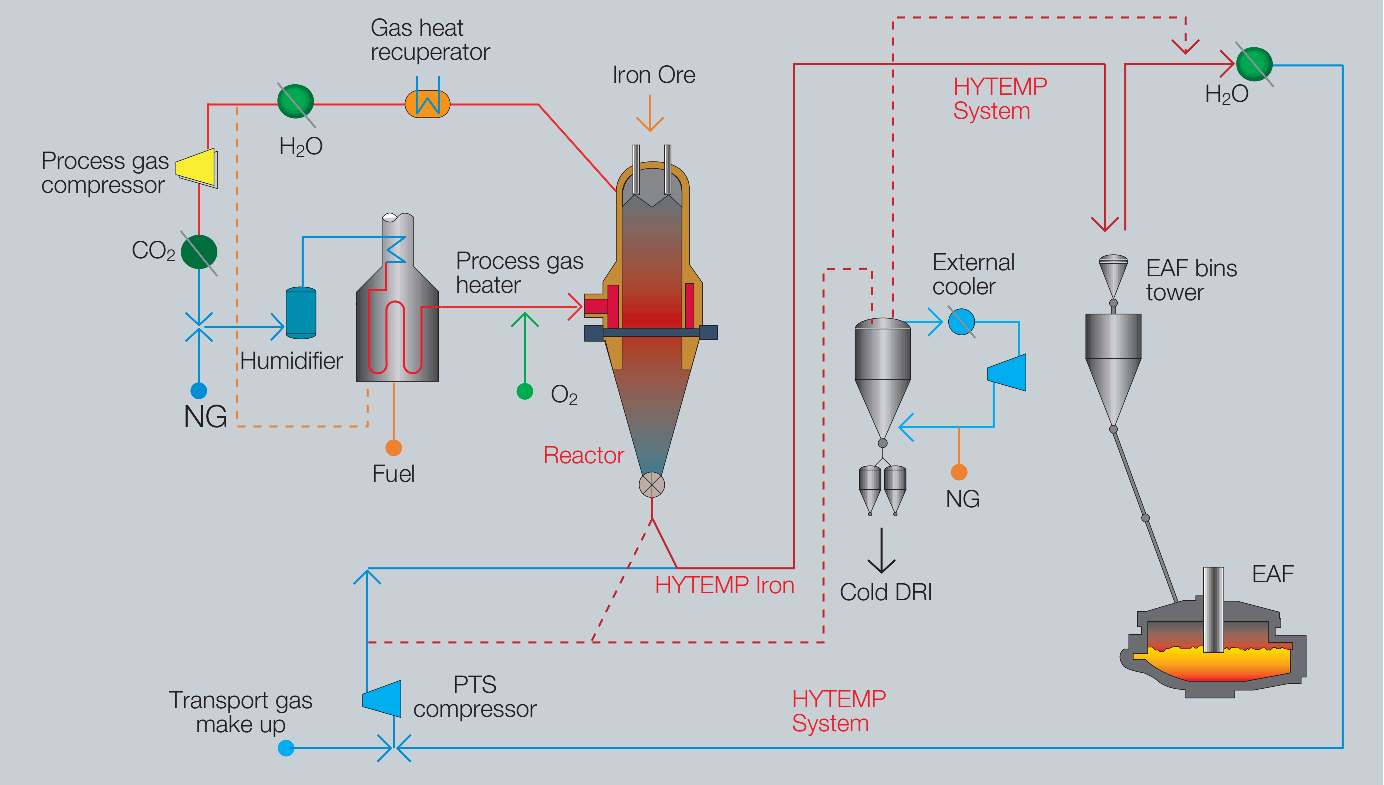
Hot DRI is discharged by means of the reactor rotary valve, through the diverter valve, which delivers material either to the HYTEMP system or to an alternative external cooler (for cold DRI production). The pneumatic transport is normally carried out at the same rate as the reactor production rate.

The product is transported to the furnace by means of a carrier gas, via the interface bin where the carrier gas is separated from the hot DRI and where depressurization also takes place.

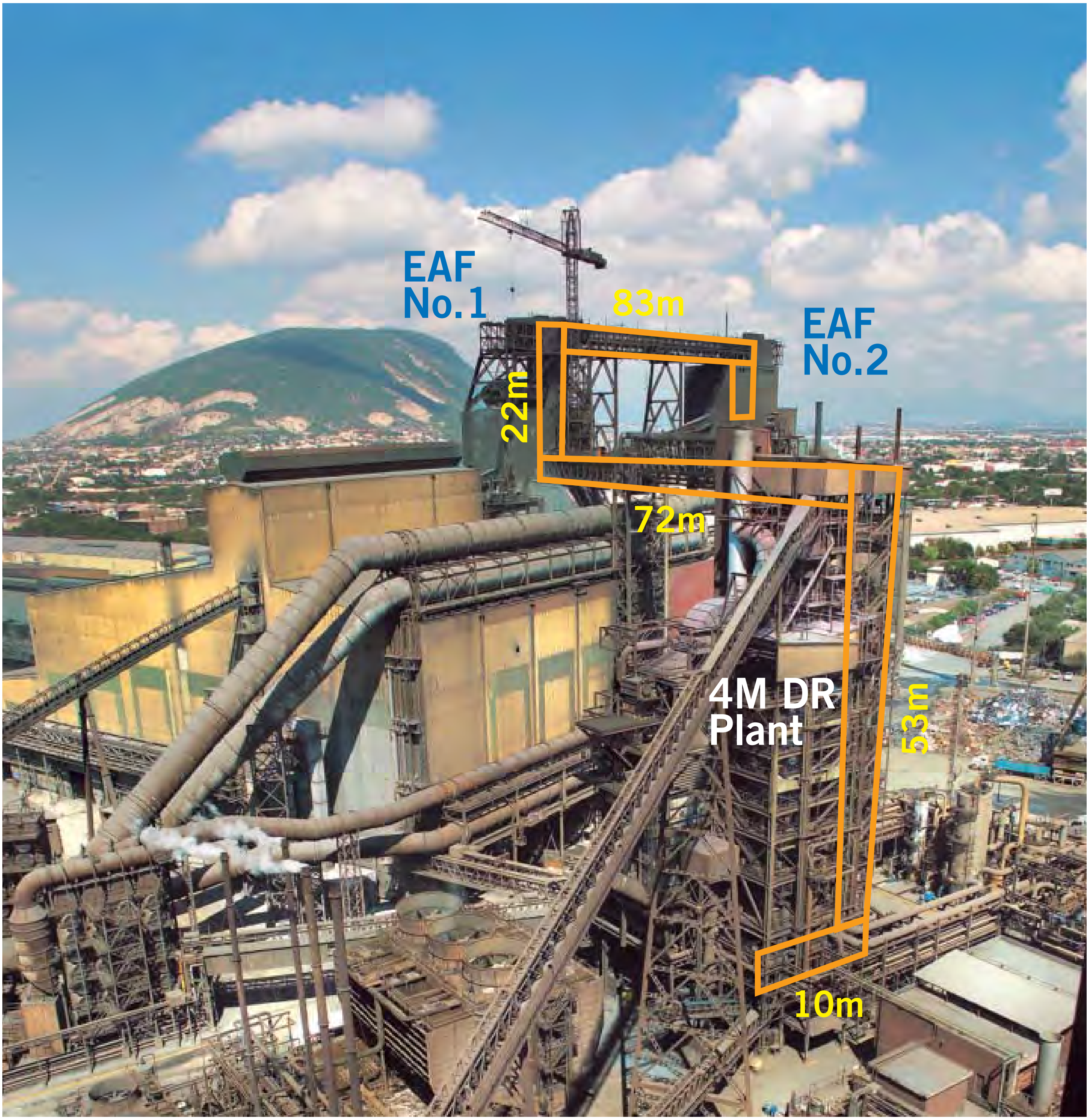
Hot DRI is discharged to the EAF surge bin, which feeds the electric arc furnace. Hot DRI is fed at about 650 ¡C to the furnace.

The carrier gas is separated from the hot DRI in the pressurized interface bins, and it is passed through a

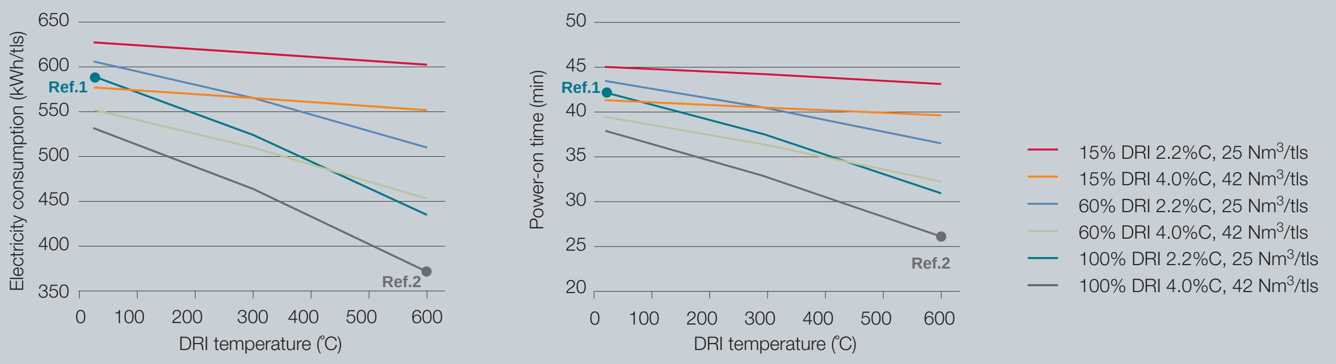
HYTEMP® system



The process



Effect of DRI temperature and carbon on electricity consumption and power-on time.



quenching/scrubbing system for cooling and cleaning. Then, the carrier gas is recycled to the pneumatic transport circuit by means of the recycle compressor, thus closing the pneumatic transport loop.

According to the particular requirements of the steelmaking facilities, an off-line discharge for DRI cooling can be used during EAF downtime, or an external cooler can be incorporated for cold DRI production, or a briquetting machine can be installed in case that part of the production is being sent for export.

As the DR plant designed for HYTEMP iron production is linked to steelmaking facilities, the reactor tower, pneumatic transport system and EAF feeding bins can be arranged in the most adequate layout to minimize distances and to match the continuous DRI output with the batch consumption of the meltshop.

Combined advantages of hot, high carbon DRI

The use of hot DRI is a proven concept in the Ternium Hylsa melt shop. In the 4M DR plant in Monterrey, the hot DRI is pneumatically transported to two EAFs. The system is shown in the photo. In this plant, hot DRI is transported through HYTEMP and fed to the DC-type EAF of the CSP▫ mill of Ternium Hylsa’s meltshop. Over 7 Mt of DRI have been transported since initial startup in 1998 making HYTEMP the most reliable and proven technology for hot DRI transport and charging to the meltshop. The benefits of high-carbon hot DRI in meltshop operations have been widely demonstrated in Hylsa s EAF. Before startup of operations of the 4M DR plant, only cold DRI to the EAF was produced in the 2M5 and 3M5 DR plants. Typical DRI quality from these plants was of 93-94% metallization and about 2.2% carbon. After the 4M plant startup, hot DRI quality has been 94% metallization and 4% carbon. The #2 EAF is a Danieli DC-type furnace with a capacity of 135 t liquid steel (ls)/heat and average active power is 110 MW. The above diagrams show the EAF operation with different percentages of DRI in the metallic charge, keeping the metallization constant at 94% for various levels of carbon. Oxygen injection is 25 Nm^3/tls for 2.2% carbon DRI and 42 Nm^3/tls for 4% carbon DRI. Production of high-carbon, high temperature DRI is reflected in important savings in the meltshop.

The sensitivity analysis on the EAF performance for two levels of carbon (2.2% and 4%) in DRI at different feeding temperatures to the EAF is shown in the above diagrams. For these cases, DRI metallization is constant at 94%. Hot DRI feed provides additional sensible heat to the EAF, reducing power consumption and tap-to-tap time, which are reflected in productivity increase. Direct feed of hot DRI in Hylsa s meltshop is carried out through the HYTEMP system. Comparative analysis, based on results of hot charging to EAF #2, related to electricity consumption and to power-on time —consequently, on productivity- are shown in the second graph. Data for different percentages of DRI with 94% metallization and various carbon levels have been included. The differences between cold DRI with 2.2% C (Ref. 1) and hot DRI at 500 ¡C with 4% C (Ref. 2) for 100% charge to the EAF are: a decrease of about 180 kWh/tls and a reduction of about 12 min in power-on time, which may represent over 25% potential productivity increase.

Energiron products

Our technology offers three product options from the same reduction process - DRI, HYTEMP Iron and HBI. Each of them can be produced to different levels of metallization and carbon. Being passivated and stable, Energiron DRI is safe to transport and store.

DRI

DRI is reduced iron pellet and/or lump, which is cooled and discharged at low temperatures (ambient). The process characteristics allow for independent control of the DRI metallization and carbon levels. Metallization can be adjusted at will, typically from 92 — 95%.

¥ Normal DRI carbon levels can be selected in the range of 0.8 — 3.0%.

¥ The ZR Process raises the bar even further by allowing the production of highly metallized high carbon content DRI. Carbon content is usually from 4 to 5%, most of which in the form of iron carbide.

¥ The high iron carbide content provides the unique passivation characteristic of this product.

HYTEMP iron

Hot DRI is discharged continuously from the reactor at >700¡C to a pneumatic transport system.

The product is transported by means of a carrier gas (reducing gas) to surge bins located above the meltshop for controlled feeding to the electric arc furnace.

HYTEMP Iron, like cold DRI can be produced to high levels of metallization and carbon for greater meltshop economy and efficiency.

The HYTEMP System is the world s first proven technology for hot discharge, transport and feeding of quality DRI to the electric furnace shop. Completely proven and reliable, with no downtime since its first implementation in 1998.

HBI

Reduced pellet or lump is discharged at temperatures >750¡C into hot briquetting presses located below the reactor discharge. Briquettes are compressed, cut and cooled to make HBI, typically for merchant sale over long distances. As with DRI, HBI is produced in controlled metallization ranges from 92 — 95%.

Carbon for HBI is typically 1.5%, higher than other HBI products that often have less than 1% carbon.

Raw and final product flexibility
Flexible input... flexible quality output.
The composition, final quality and cost of any DRI are dependent on the quality of the iron ores used for reduction. Energiron plants offers the flexibility to operate with cheaper oxide pellets, lump ores or mixtures of both.
There is no practical limitation regarding the chemical composition of the iron ore. In particular since the reforming gas is not recycled to the reformer, the process is very flexible for using high sulfur iron ores.
Moreover, due to the lower gas velocities through the reactor, the use of hydrogen-rich gases and the mechanical system sealing devices, the use of friable lump ores can be maximized.
Advantages of Energiron product quality include:

¥ Cold DRI, HBI and Hot DRI (HYTEMP▫ System) in every combination.
¥ Ability to control both metallization and carbon levels, unique in the industry. The product quality can be adjusted to best fit the client needs.
¥ High metallization, high-carbon DRI. Energiron is the only DR technology currently capable of producing High Carbon DRI with more than 90% of the Carbon as Iron Carbide.
The unique product provides virgin iron plus energy for the electric furnace.
¥ High Carbon DRI is stable and naturally passivated, and can be transported safely for commercial applications.

Energiron added-value products for competitive steelmaking

The virtual absence of residual elements makes DRI the ideal complement to EAF scrap charges, as well as the recommended charge material for producing most high grades of steel.
Additionally, DRI and HBI serve as quality metallic sources for blast furnace and BOF shops.
DRI can be continuously charged for even greater advantages. The use of this quality feedstock in the EAF provides numerous advantages in steelmaking, including:
¥ Greater uniformity and predictability
¥ Better slag foaming
¥ Increased efficiency and productivity, especially when using High Carbide DRI
¥ Lower total steel cost

The benefits of hot charging DRI to the electric furnace include, in addition to the above, a substantial productivity increase and a reduction in the electrical energy consumption, by taking advantage of the sensible heat from the DRI in the electric arc furnace.

Effect of high carbon

Carbon in Fe_3C form

Chemical Energy Contribution The conversion of Fe_3C into iron and carbon is an exothermic reaction which improves the thermal efficiency in the EAF.

Efficient use of carbon in EAF The combined carbon is totally used, minimizing external carbon (graphite) additions.

Higher stability during handling Iron carbide is more stable and is safer to be stored and transported.

Easy foamy slag generation As high carbon DRI enters in contact with free or combined oxygen.

Carbon and DRI are fed at the same time The same system controls the feeding rate of metallic charge and carbon additions.

Easier operation with intensive oxygen use As DRI feeding rate is varied, carbon is varied as well and, as a consequence, oxygen use is varied.

Typical maximum residuals levels for different steel grades

	S	Cu	Ni	Mo	Cr	Sn	Total
Rebar	0.05	0.40	0.35	0.08	0.15	0.08	1.06
Structural	0.03	0.40	0.15	0.08	0.15	0.03	0.81
Plating	0.03	0.20	0.10	0.02	0.10	0.02	0.44
Forging	0.03	0.15	0.12	0.02	0.12	0.02	0.43
Low quality flat	0.03	0.15	0.08	0.02	0.08	0.02	0.35
Drawing	0.03	0.10	0.10	0.03	0.10	0.02	0.35
Cold heading	0.03	0.10	0.08	0.01	0.08	0.02	0.29
Fine wire	0.03	0.10	0.08	0.02	0.08	0.02	0.29
Deep drawing	0.03	0.06	0.10	0.02	0.07	0.02	0.27

Typical residuals levels in scrap and in DRI/HBI

	S	Cu	Ni	Mo	Cr	Sn	Total
No.1 Bundles	0.07	0.07	0.03	0.008	0.04	0.008	0.23
Shredded	0.07	0.22	0.11	0.02	0.18	0.03	0.63
No.1 Heavy Melt	0.07	0.25	0.09	0.03	0.10	0.025	0.57
No.2 Bundles	0.07	0.50	0.10	0.03	0.18	0.1	0.98
No.2 Heavy Melt	0.07	0.55	0.20	0.04	0.18	0.04	1.08
DRI/HBI	0.005	0.002	0.009	<0.001	0.003	trace	0.02

Energiron plants

Energiron plants are designed for maximum savings in both investment and operating costs. Multiple reduction reactors can share a common reducing gas source. Auxiliary systems and materials handling equipment can be common to several modules. With the ZR process configuration, further economies are obtained based on the smaller overall plant size and equipment requirement.

Micro-module

This basic plant design concept was developed specifically for markets where small volumes of DRI are needed, and where coal-based technologies have until now been the only option.
The highly functional, no-frills version of the traditional plant provides the highest quality DRI available to companies requiring annual capacities in the range of 200,000 tpy.

Mini-module

This is the optimum sized DR plant for the typical steel mill producing quality products. It s a 500,000 tpy DR plant with the reliability and quality for which HYL is renowned.
A totally re-engineered DR module, designed with the latest technology such as reformerless (ZR) reduction and production of High Carbide Iron.
The Mini-Module takes advantage of modular design for a construction schedule that means shorter project implementation and a faster startup.

Custom modules

Energiron plants can be designed to incorporate a variety of technologies, depending on the particular needs of a client or project. Plant capacities can range anywhere up to 2.0 million metric tons per year in a single reactor, to multiple reactor plants of several million tons per year capacity.

Energiron technologies

Energiron plants can include such technologies as:

¥ CO_2 absorption systems
For recovery and sale of carbon dioxide to commercial users as a by-product;
¥ Electric power generation
In plants using a natural gas-steam reforming unit, the high pressure steam produced can be used in a turbo generator or in a set of turbines to supply power to both the DR facility as well as to the surrounding area.
¥ Minimum water consumption
Taking into consideration that the reduction reactions are carried out primarily with hydrogen, water is generated and it can be recovered for further utilization in regions where its value warrants.

Plants can be designed for cold DRI discharge, for hot discharge with HBI production or in combination with the HYL HYTEMP System to form an integrated production facility from the DR plant to the EAF mill. Dual product plants can also be built to produce both hot DRI (for either briquetting or HYTEMP transport) and cold DRI, using the cooling gas circuit or via an offline cooling vessel.
Energiron plants use simple, reliable equipment in an independent operating environment. The result is efficient operation and high plant availability.
Reforming, when carried out within the reduction reactor, eliminates the need for an external gas reformer. In plant configurations that include a reformer, a proven, standard design reformer is used that is available from a number of suppliers. The typical life of the reformer catalyst and tubes is very long, which means lower maintenance costs than processes using special design reformers.
Additionally, for the same size capacity plant, an Energiron plant reformer is significantly smaller than those of other technologies.

The reduction reactor is a highly efficient design, permitting a mass flow of solids and uniform distribution of reducing gases. The reduced product is thus more uniform throughout, in both physical and chemical characteristics.

The environment

Energiron plant emissions are in accordance with the most stringent environmental regulations anywhere. This is achieved in large part due to the process itself; while other processes require heat recovery equipment which tends to increase the NO_x levels, Energiron technology is efficient by design due to its process configuration. Thus, while achieving high overall thermal efficiency in the plant, there is no significant need for preheating the combustion air to high temperatures in the reformer (when used) or in the heater, thus eliminating the possibility of high NO_x generation. For other DR processes, this is not possible and the choice between decreasing thermal efficiency or installing expensive de-NO_x units remains.
Furthermore Energiron plants also offer the option of selective recovery of CO_2 which can be cleaned and sold as a cash by-product rather than venting it to the atmosphere.

Plant sizes standard modules

Capacity (Mtpy)	Reactor size approx. nominal ID (m)
200,000	2.5
500,000	4.0
800,000	5.0
1,200,000	5.5
1,600,000	5.7
2,000,000	6.0

7,800-8,000 hr/yr net working time

Typical consumption figures

Item	Unit		Remark
Plant capacity	tpy	200,000-2,000,000	
Metallization		‡ 93%	adjustable to suit meltshop needs
Carbon (controlled)		0.8%-5%	depending on extent of in-situ reforming
Inputs		**Specific consumption**	
Iron ore	t/t	1.38-1.40	depending on DRI carbon
Natural gas	Gcal/t	2.24-2.60	depending on DRI carbon & temperature, extent of in-situ reforming and power co-generation
Electricity	kWh/t	0-85	depending on power co-generation
Water	m^3/t	0-1.3	depending on the water recovery system
Labor	m-h/t	0.11-0.17	depending on plant size
Maintenance	US$/t	3.0-3.3	for cold-hot DRI

Global project execution

The Energiron alliance offers the unique flexibility of providing everything from a basic technology package to full turnkey projects. Our support runs from process technology through operations and maintenance training and technical assistance, not only in the DR plant but in the steel shop as well.
In addition to technology licensing and conceptual engineering, we provide maintenance and operations training, project and erection supervision, support during startup and commissioning, and a wide range of technical services after startup.
Our own qualified operations and maintenance personnel can even take charge of the operation of a DR facility on a contract basis, if requested.
We not only offer our expertise in direct reduction but in the steel mill as well. Our technical staff has years of experience providing training and technical assistance for meltshop and steel mill personnel, with unequaled success in raising productivity levels in all areas, whether or not the steel mill is a DRI-based operation. From mining, pelletizing and materials handling to steel mill operations and maintenance, computerized process controls, and even including administrative and commercial functions — we provide technologies and services for all areas of ironmaking and steelmaking.

Support for projects includes

Assistance in arrangement of project financing

Engineering, procurement and construction

Plant startup and commissioning

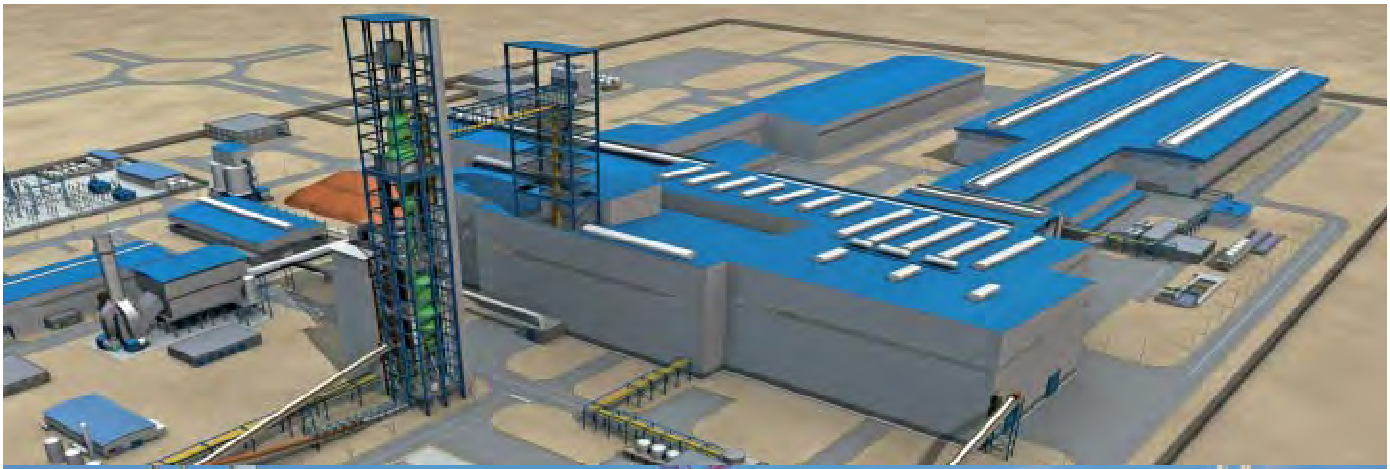
Training and know-how transfer

Rapid project implementation and learning curve







ENERGIRON
THE INNOVATIVE DIRECT REDUCTION TECHNOLOGY

www.energiron.com
info@energiron.com

Danieli headquarters
Via Nazionale, 41
33042 Buttrio (UD) Italy
Tel (39) 0432.1958111
Fax (39) 0432.1958289
www.danieli.com
info@danieli.com

Tenova headquarters
Via Monte Rosa, 93
20100 Milano, Italy
Tel (39) 02.43841
Fax (39) 02.4693026
www.tenovagroup.com
info@mx.tenovagroup.com

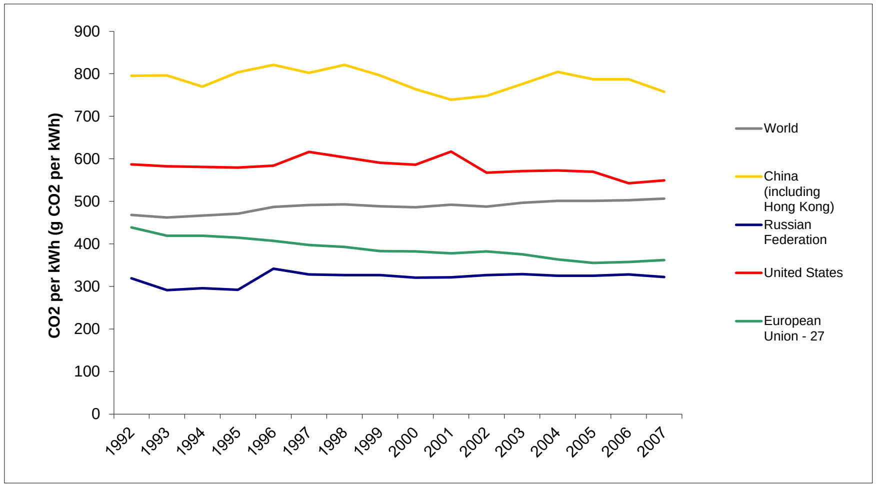
HYL Technologies
Av. Eugenio Clariond Garza, 155
San Nicolás de los Garza, N.L.
66452 México
Tel (52) 81.8865.2801
Fax (52) 81.8865.2810
www.hyltechnologies.com
info@hyltechnologies.com

Attachment 4

Energy Calculation Support Data

Emissions per kWh of Electricity and Heat Output

		TIME	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007
COUNTRY	PRODUCT	FLOW																
World	Total	CO2 per kWh (g CO2 per kWh)	468.2819	461.8466	466.9087	471.1349	486.6211	491.5516	493.3	488.2372	486.0594	491.8782	487.6168	496.5901	501.3615	501.2705	502.6507	506.7492
China (including Hong Kong)	Total	CO2 per kWh (g CO2 per kWh)	795.1731	796.0307	770.0306	803.8834	820.7678	802.5431	821.2774	796.1171	763.7152	739.2958	747.8536	776.0018	804.4473	786.9218	787.2251	757.9758
Russian Federation	Total	CO2 per kWh (g CO2 per kWh)	319.3949	291.11	296.0201	291.7552	341.8829	328.3193	326.3582	326.9592	320.761	321.4793	326.6583	329.3001	324.8552	324.9678	328.5654	322.508
United States	Total	CO2 per kWh (g CO2 per kWh)	587.138	582.215	581.17	579.23	584.088	616.448	603.647	590.509	585.917	616.843	567.369	570.863	572.479	569.737	542.455	549.357
European Union - 27	Total	CO2 per kWh (g CO2 per kWh)	438.86	419.3492	418.9673	414.7438	407.4751	397.1537	392.8263	383.4258	382.2738	377.9707	382.0315	375.4884	363.6605	354.9777	357.8205	362.2026



CO2-emission generation of electricity in Europe

	g/kWh
Griechenland	950
Polen	879
Malta	870
Zypern	792
Tschechien	652
Dänemark	624
Bulgarien	610
Rumänien	583
Niederlande	538
Irland	505
Slowenien	504
Estland	498
Deutschland	494
Italien	475
Großbritannien	469
Portugal	420
Ungarn	399
EU-27	390
Spanien	359
Lettland	318
Luxemburg	280
Belgien	267
Finnland	219
Litauen	183
Österreich	168
Frankreich	46
Schweden	15
Schweiz	11
Norwegen	4

reference source: VEÖ, Eurelectric

Operating results with hot DRI charge at Emirates Steel Industries

The use of 90% hot DRI, 10% cold DRI as the charge materials for a UHP 150t EAF with carbon-oxygen injection is demonstrating excellent energy efficiencies and plant productivity. When 100% hot-charged DRI becomes available shortly, productivity will further increase to 220tls/hr, tap-to-tap time will reduce to 41mins and electricity consumption will decrease to 380kWh/tls.

Authors: Paolo Razza and Damiano Patrizio
Emirates Steel Industries and Danieli Centro Met

In the past few decades the importance of DRI as a substitute for scrap has increased significantly. A recent example of the use of 100% DRI charge is the 150t UHP EAF FastArc™ at the Emirates Steel Industries (ESI) site in Mussafah Industrial Area, Abu Dhabi. The complex consists of a 1.6 Mt/yr direct reduction (DR) plant, 1.4Mt/yr steelmaking and casting plant, 0.62Mt/yr high-speed bar mill and 0.48Mt/yr high-speed wire rod mill.

EAF FEATURES

The split shell AC furnace operates on a 90% hot (600°C) DRI, 10% cold DRI charge, which is continuously fed via the 5th hole from the HYTEMP tower (see *Figure 1*). The EAF maintains a hot heel of 50t and has a rated tap weight of 150t liquid steel. Nominal productivity target is 196tls/hr with a tap-to-tap time of 46mins. The main geometrical data of the EAF are shown in *Table 1*.

One of the key factors in the EAF design is the location of the DRI feeding point with respect to the off-gas 4th hole. It was placed far enough from the off-gas elbow to minimise the loss of DRI fines in the fume treatment plant (FTP). For this reason the DRI entrance is on the slag door side of the furnace. The cross-section of the 4th hole was designed to reduce the off-gas speed below 30m/s.

Electricals The EAF transformer has a rated apparent power of 130 + 20% MVA and allows selection of 18 different tap positions for the best combination of arc voltage, arc current and power factor during process stages. The furnace secondary circuit is designed for a maximum current of 80kA. The main electrical data are summarised in *Table 2*.

The maximum active power applied is in the range 112-114MW, with a secondary current of 74-75kA and a specific power of 0.75MW/tls. This value can be considered one of the highest when compared with other furnaces melting DRI in amounts greater than 70%.

FastArc™ injection technology The injection system has been conceived with three kinds of injectors, each with

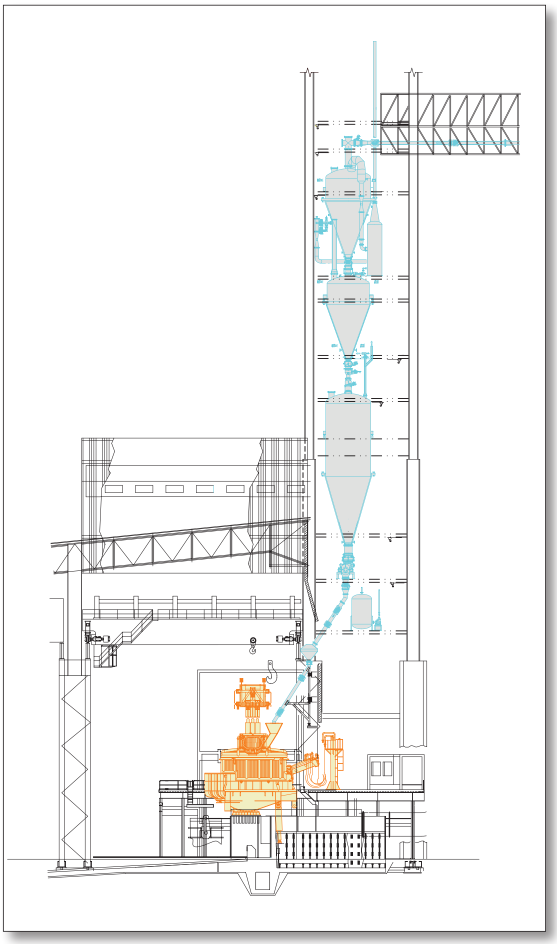


Fig 1 **EAF and HYTEMP DRI tower**

Item (diameter)	
Lower shell	7.0m
Electrode	710mm
Pitch circle	1,400mm

Table1 **Main furnace dimensions**

▷

Item	Value
EAF transformer rated power, MVA	130
Overload, %	20
Frequency, Hz	50
Primary voltage, kV	33
Secondary voltage range, V	1,250–650
Secondary voltage at full power, V	1,250–1,120
SVC rated power, MVAr	170
Series reactor rated reactance, ohm	1.3

Table 2 **Main electrical data**

Module	Oxygen flow rate, Nm^3/h	Material feed rate, kg/ min
Oxygen jet 1-2-3-4-5	2,200	–
Carbon jet 1-2-3	150	15–30
Carbon injector	–	15–30

Table 3 **EAF injection system**

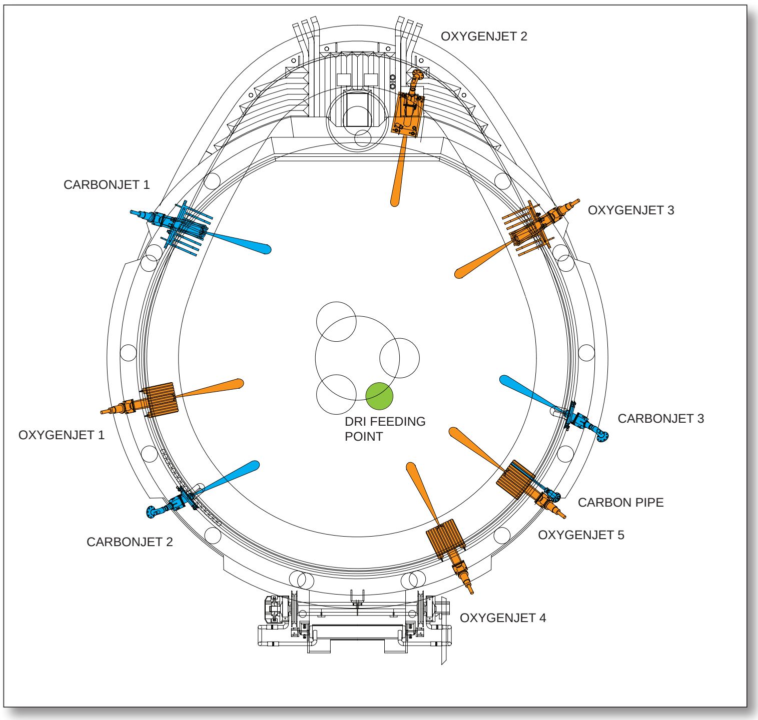


Fig 2 **Oxygen and carbon injector layout**

its own function, to achieve the best possible performance, as shown in *Table 3*.

Figure 2 shows installed injectors. The layout promotes maximum stirring effect in the region of the DRI addition point and to provide a uniform chemical input around the furnace circumference, improving thermal balance and avoiding cold spots. Carbon injectors are placed in the proximity of the oxygen injectors to promote slag foaming and to protect refractory hot spots from arc radiation.

PROCESS DESCRIPTION

DRI characteristics The main characteristics of the DRI are summarised in *Table 4*, the material analysis taken from the sampling bin during a 24-hour period.

Melting procedure *Figure 3* shows the detailed melting profile used for a 10% cold DRI, 90% hot DRI charge. The DRI feed rate is increased progressively through the heat starting from an initial value of 2,000kg/min and ramping up to the maximum value of 5,800kg/min after 13mins of power-on. Thanks to the excellent degree of metallisation, the balanced amount of carbon and the high reactivity of the high temperature pellets, very good foaming slag is achieved from the beginning of melting just after the DRI comes into contact with the oxidised hot heel.

NB: during the first 6mins the tap hole is filled with sand, the slag door breast is cleaned, and any other fettling or gunning is performed.

After 4mins from power-on, the maximum transformer tap is selected, reaching a typical maximum power in the range 100–108MW. This is even more relevant if we consider the refractory wear index (RWI), applied during the main phase of DRI feeding. The parameter reaches 230–235kWV/cm^2 for 75% of the power-on time without negative impact on the life of the refractory or the water-cooled panels. The DRI specific feed rate is 55kg/min/MW.

OPERATIONAL RESULTS

Thanks to increased utilisation of chemical energy by means of the DANARC injection system with Modules technology and the high power transformer, the furnace has achieved very fast DRI melting rates, equating to a productivity of 200tls/hr, well above the guaranteed value, and with power-on times of 35mins.

Consumptions Specific consumptions from 26 heats indicated in *Table 5* were recorded during commissioning. These results were achieved thanks to the excellent quality of the DRI in terms of metallisation and its 2.1% C. If we compare the results recorded with the utilisation of 90% of hot DRI at 600°C (column A), with 100% cold DRI charge (column B), there is a reduction of the

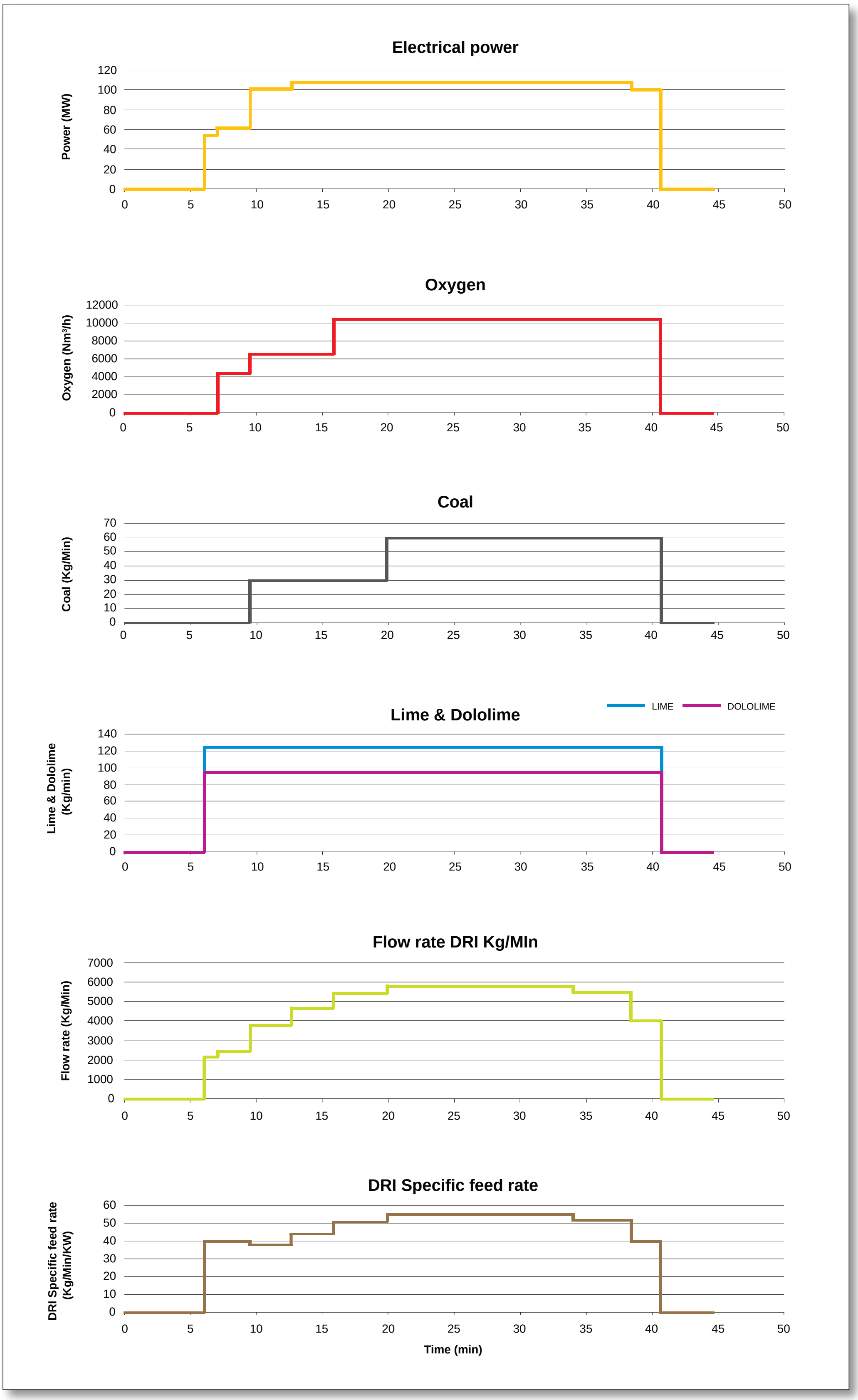


Fig 3 **Melting profiles**

▷

Time	A Fe metalisation %	B Fe total %	C Metalisation %	C %	S %
08.00	87.8	91.8	95.6	2.07	0.011
10.00	86.3	91.6	94.2	2.15	0.009
12.00	86.9	91.6	94.9	1.94	0.008
14.00	88.6	93.1	95.2	2.12	0.009
16.00	86.9	91.6	94.9	2.20	0.007
18.00	87.2	91.7	95.2	2.16	0.008
20.00	86.6	91.2	95.0	1.81	0.009
22.00	86.8	91.4	94.9	2.21	0.010
01.00	87.3	92.2	94.7	2.13	0.008
03.00	87.3	92.8	94.1	2.17	0.010
05.00	88.0	92.9	94.7	2.06	0.010
Average	87.2	92.0	94.9	2.09	0.009

Table 4 **DRI analysis (note: A=B*C)**

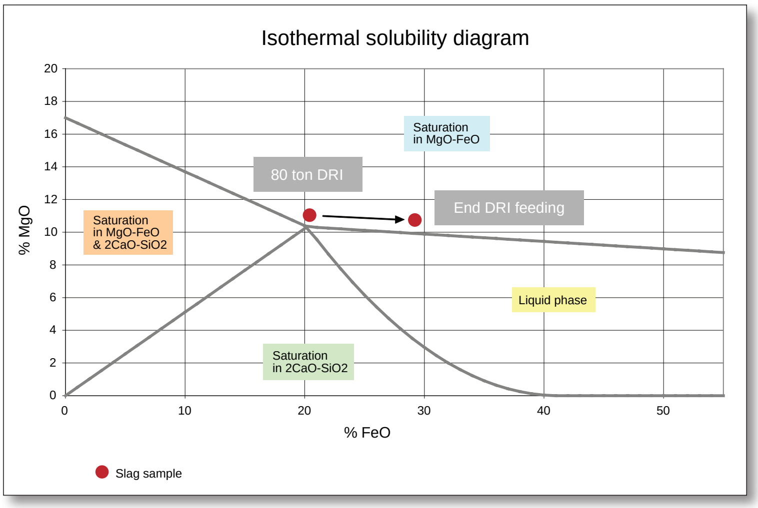


Fig 4 **Isothermal solubility diagram**

electrical energy equal to 141kWh/tls and a power-on reduction of 13.9mins.

The energy benefit comes mainly from:

- The enthalpy of the raw material corresponding to 105kWh/tls
- Additional 1.5Nm^3/tls oxygen used with hot DRI, corresponding to 4kWh/tls
- Improved metallisation of the hot DRI with respect to the cold DRI. The delta is equal to 1.7% corresponding to 23kWh/tls
- Reduced thermal losses due to decreased power-on time, corresponding to 9kWh/tls

Considering the DRI feed rate was kept at 5.8t/min (55kg/min/MW) during the main phase of the process, the oxygen injection system achieved correct decarburisation of the bath, corresponding to a specific rate of 320kg/m^2/hr.

Slag control Slag samples are taken at the end of the heat just before tapping. Typical slag composition is reported in *Table 7*.

Slag formers are continuously fed with the DRI to maintain slag basicity by compensating for the acid gangue of the pellets. To promote the initial foaming and enable the electrical power to ramp up quickly when the level of the bath is low, higher specific additions of slag formers are made in the first third of the heat, reaching values of 58kg/t DRI. During the main DRI addition this value decreases to 38kg/t DRI, increasing again to 55kg/t DRI in the final stage of the heat when bath and slag temperature are higher.

When compared with a scrap-based process particular attention must be paid to the MgO content of the slag as the refractory walls are potentially exposed to arc radiation immediately the power is switched on. An MgO-saturated slag will not only be less aggressive towards the refractory but also increase foaming. The target is a basicity IB_2 (CaO/SiO_2) is 2.1, or an IB_3 ($CaO/[SiO_2+Al_2O_3]$) of 1.7. The MgO saturation value is 10% according to *Figure 4*. The slag basicity is controlled throughout the process as shown in *Figure 4* where we can see how the slag composition changes between the 80t of DRI fed and the end of DRI feeding. Although Fe oxidation increases towards the end of the process, the MgO saturation value is maintained, allowing favourable slag foaming conditions.

In terms of refractory consumption this slag practice helped achieve the refractory guarantee figures. Regarding walls, bottom and roof delta, consumption was 2.48kg/tls for an overall refractory consumption of 4.8-4.9kg/tls. These data refer to a non-optimised hot DRI practice, so further consumption savings can be expected in the coming months.

PROCESS YIELD

One of the main targets of the project was to achieve a process yield of 86%, defined as the ratio between liquid steel and charged DRI. A value of 87.7% was attained. Particular attention was paid in the design of the furnace and of the process to avoid material losses in the FTP and through the slag door. On the basis of the data collected during the 10 heat performance test, the yield of good billet from charged DRI was 87.5% and for a 28-heat day production run, yield was 87.2%.

A deeper analysis of yield is as follows. Data on slag and FTP dust weights are collected on a regular basis. The average results for January 2010 were 145kg/tls of slag and18kg/tls of dust. Starting from the slag weight data it was possible to determine the iron lost in the slag according to the gangue content of the DRI, slag forming addition and refractory consumption. The results equate to 31kg of Fe/tls.

On this basis it was possible to determine the total losses of the charge as 140kg/tls and categorised as shown in *Figure 5*. Losses are detailed below:

- FTP dust; 18kg/tls confirmed from data collected on site, and indicating the right choice in terms of EAF roof design and DRI feeding point
- Decarburisation of the charge is 23kg/tls. The value comes directly from the difference in carbon content in the charge and the tapping carbon in the steel
- DRI gangue is 52kg/tls. According to DRI analysis the gangue content is about 4.5%
- Oxygen in the charge is 16kg/tls, calculated according to DRI metallisation of 94.85%
- Iron lost in slag is 31kg/tls. Slag losses through the slag door are minimised by process control.

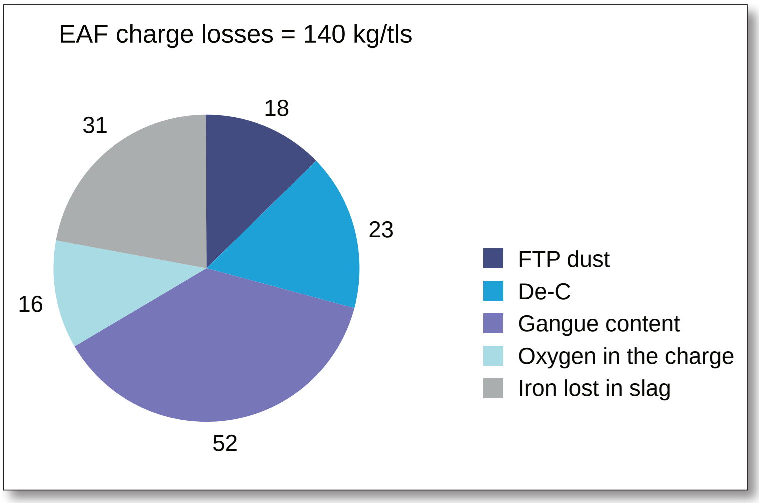


Fig 5 EAF charge losses

	A **90% hot DRI, 10% cold DRI**	**B** **100% cold DRI**
Tap-to-tap, min	45.5	58.5
Power-on time, min	35	48.9
Electrical energy, kWh/t	392	533
Oxygen, Nm^3/tls	34.8	33.3
Coal injected, kg/tls	9.7	17.3
Coal 5th hole, kg/tls	3.2	10.0
Average power, MW	102	98
Tapping temperature,°C	1,640	1,640
Steel liquid tapped, t	152	150

Table 5 Consumption figures

CaO	**MgO**	**SiO_2**	**Al_2O_2**	**MnO**	**FeO**
37.7	10.8	17.9	4.2	0.9	28.3

Table 6 Typical slag composition, %

This total loss of 140kg/tls equates to an EAF charge yield of 87.7%. It can also be expressed as a yield of 95.3%, the ratio between liquid steel and total iron in DRI which is 92%, ie, excluding the gangue content which is a direct consequence of iron ore quality. The results achieved so far represent one of the highest yields reached for 100% DRI-based processes.

FUTURE DEVELOPMENTS

A further improvement will be to increase DRI carbon content up to 2.5% with a proportional increase in oxygen total flow rate, and to operate with 100% hot DRI. DRI production will increase from 220 to 250t/hr. The plant handling system will soon be upgraded to cope with this. The expected results, based on a charge of 100% hot DRI at 600°C, will further reduce electrical energy consumption to 380kWh/tls and a power-on time of 34mins. This will increase productivity to 220tls/hr, with a tap-to-tap time of 41mins.

CONCLUSIONS

ESI's EAF process exceeded contractual figures, showing excellent thermal efficiency by exploiting the quality and enthalpy of the hot DRI charged. The plant is a good benchmark thanks to the excellent results in process time, electrical consumption and thermal efficiency. **MS**

Paolo Razza is Meltshop Manager at ESI, Mussafah Industrial Area, Abu Dhabi. Damiano Patrizio is Senior Engineer, Process Technology, at Danieli Centro Met, Butrio, Italy.

CONTACT: a.fragiacomo@danieli.ita