TECHNOLOGY OPTIONS FOR CLEAN COAL POWER GENERATION WITH CO_2 CAPTURE

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ABSTRACT

Coal contributes to about forty percent of the electricity generation in the world today and is expected to maintain its dominant share in the foreseeable future. The state-of-the-art coal-fired power plant today with 600 ºC steam temperature is about 20% more efficient than the average operating power plants, and can reduce emissions such as SO_2, NOx, and mercury to ultra-low levels. Hitachi is developing a full portfolio of new clean coal technologies aimed at further efficiency improvement, 90% CO_2 reduction, and near-zero emissions of other pollutants. The technologies under development include 700 ºC next generation ultrasupercritical boilers and steam turbines, post-combustion CO_2 absorption, oxyfuel combustion, and IGCC with CO_2 capture. This paper provides an overview of the development status of the technologies and discusses their expected performance and economic impacts, with special focus on post-combustion absorption and oxyfuel combustion which are promising near-term solutions of CO_2 capture for new and existing power plants.

Keywords: CO_2 Capture Options, Post Combustion Capture, Oxyfuel Combustion

1. INTRODUCTION

In the United States about one half of the electricity is from coal. Worldwide coal contributes to over 40% of the electricity generation today and its share is expected to increase steadily over the coming decades. The continued dominance of coal in global energy structure and the growing concern of climate change necessitate accelerated development and deployment of new technologies for clean and efficient coal utilization. Coal-fired power plants with CO_2 capture and sequestration (CCS) are widely expected to be an important part of a sensible future technology portfolio to achieve overall global CO_2 reductions required for stabilizing atmospheric CO_2 concentration and global warming.

New coal-fired plants built in the coming years will have very high efficiency and extremely low emissions by using the latest ultrasupercritical technology and pollution control systems. They will also need to be "capture ready" which means that they can be retrofitted with CCS technologies as these technologies become commercially available and still offer competitive cost of electricity compared to other means of power generation. Advanced amine-based post-combustion CO_2 capture and oxyfuel combustion are two of the promising technologies to enable CCS for new and existing coal-fired power plants.

2. POST-COMBUSTION CO_2 CAPTURE

Amine-based CO_2 separation has been used since the 1930s for applications such as natural gas purification. It is a leading technology expected to be available commercially within the next decade to enable CCS for coal-fired power stations. However, traditional CO_2 capture process utilizing conventional amine solvents is very energy intensive and is also susceptible to solvent degradation by oxygen, SOx and NO_2 in coal-fired flue gas, resulting in large operating cost. According to recent U.S. DOE/NETL studies, MEA-based CCS will increase the cost of electricity (COE) of a new pulverized coal plant by 80-85% and reduce the net plant efficiency by about 30%.

As a global technology and equipment provider for complete thermal power plants, Hitachi addresses the above challenges of amine-based CCS for coal power with the following approach: 1) development of a flexible CO_2 capture process and the latest advanced amine-based solvent with long service life and low regeneration energy requirement; 2) design integration of steam cycle and CO_2 absorber - desorber process; 3) total plant re-optimization involving the boiler, turbine, air quality control system, and CCS system.

DEVELOPMENT OF POST-COMBUSTION CO_2 CAPTURE TECHNOLOGY

CO_2 Scrubbing Process Overview

The Hitachi CO_2 scrubbing system is designed to achieve 90% capture with significant cost savings and efficiency improvement over current amine scrubbing technologies. The design approach considers the redundancy and reliability requirement according to power industry standards and the flexibility to allow plant owners to utilize common commercial amine solutions and advanced amine-based reagents.

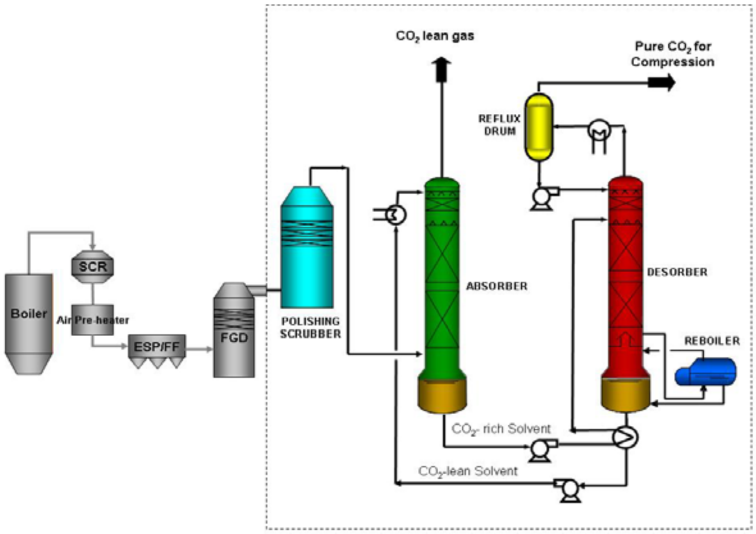


Figure 1. Process diagram of CO_2 capture system

The CO_2 capture system is based on proven process engineering principles. As shown in Figure 1, the main system components are a polishing scrubber (optional), an absorber, a desorber, a reboiler, and a reclaimer (not shown). Flue gas from the power plant or industrial facility is first sent to the polishing scrubber to reduce SO_2 and SO_3 to below about 10 ppm (combined), as well as to cool the flue gas to 40-60°C (100-140^{o}F) range for maximum CO_2 capture in the absorber. Caustic soda (NaOH) solution is used to remove SOx and therefore minimize formation of heat-stable salts (HSS) in the downstream absorber-desorber loop. The clean and cool flue gas leaving the polishing scrubber enters the packed bed absorber where it reacts with the amine-based solvent. Counter-current flow through two or more stages of structured packing maximizes contacting surface area and mass transfer. Solvent solution is injected into the top and collected from the bottom of the packing layers. CO_2-depleted flue gas leaving the top of the absorber is vented to the stack. The CO_2-rich solution leaving the bottom of the absorber is sent to the desorber via a cross heat exchanger where it gets heated. In the packed-bed desorber, pure CO_2 gas is stripped away from the CO_2-rich solution by contacting it with steam in a counter current direction. A part of the CO_2-lean solution from the bottom of the desorber circulates through a reboiler where auxiliary steam is utilized to partially vaporize the amine solution which, upon returning to the desorber provides the heat needed for amine regeneration to release CO_2. Regenerated solvent is re-sent to the absorber after it gets cooled in the cross heat exchanger. Water washing nozzles and mist eliminators are located at the top of both the absorber and desorber to prevent entrained liquid droplets from leaving the system. A reclaimer may be used to control the HSS level in the system. The reclaimer extracts a stream of amine solution, vaporizes it with the heat from auxiliary steam, and returns the vapor to the desorber, while rejecting the HSS from the bottom.

Process and Solvent Development

Hitachi started post-combustion CO_2 capture R&D specifically for coal-fired applications in the early 1990s, when the first bench-scale and pilot test programs were initiated. Since then, the company has been continually improving process design and the technology for full-scale power plant applications.

Bench-scale studies (Figure 2) with simulated flue gas have been performed regularly to screen and identify promising absorbents and additives for maximum CO_2 removal efficiencies while keeping solvent degradation and energy consumption low.

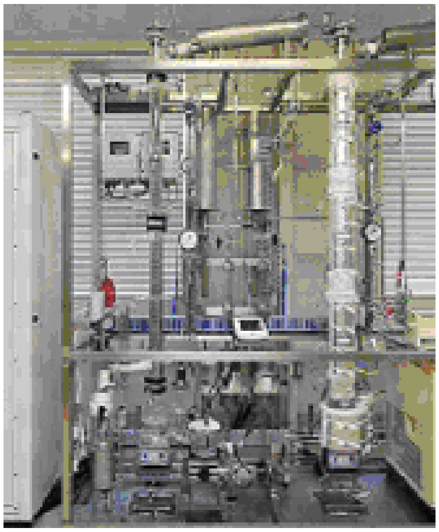


Figure 2. Bench-scale test rig

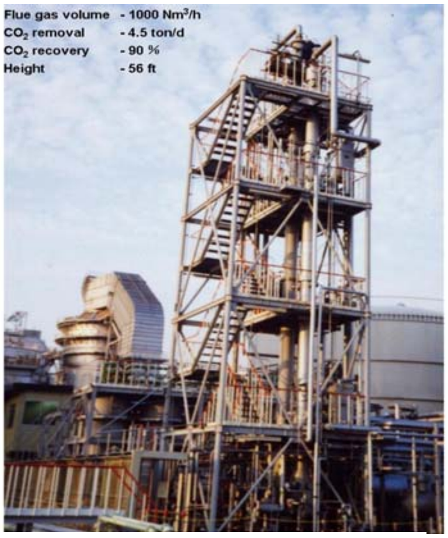


Figure 3. Pilot plant at Yokosuka

Figure 3 shows Hitachi's first CO_2 capture pilot plant built at Yokosuka Thermal Power Plant Unit 2 in cooperation with Tokyo Electric Power Corporation (TEPCO) in Japan. The slipstream test facility treated 1000 m^3N/h (620 scfm) of flue gas for CO_2 removal.

During a two-year demonstration period, five solvent solutions were tested, including a commercial MEA as benchmark and three proprietary solvent formulations. The test for H3, Hitachi's proprietary solvent formulation and the best performing solution of the five, lasted 2000 hours under various operating conditions and generated a large database of solvent and system behavior.

Figure 4 shows that in over 2000 hours of testing under various loads and inlet CO_2 concentrations, the H3 solvent consistently achieved greater than 80% CO_2 removal with the average well above 90%. H3 has specific regeneration energy of 2800 kJ / kg CO_2 which is the best among tested solvents and much lower than commercial MEA. It also has high absorption capacity, thus requiring lower liquid-to-gas ratio for 90% capture than that for MEA and resulting in significant operating cost savings. Hitachi continues the refinement of the proprietary solvent blends in its laboratories. The new solvent formulations are also being tested and compared with other commercial or near commercial solvents by independent institutions in Japan and USA.

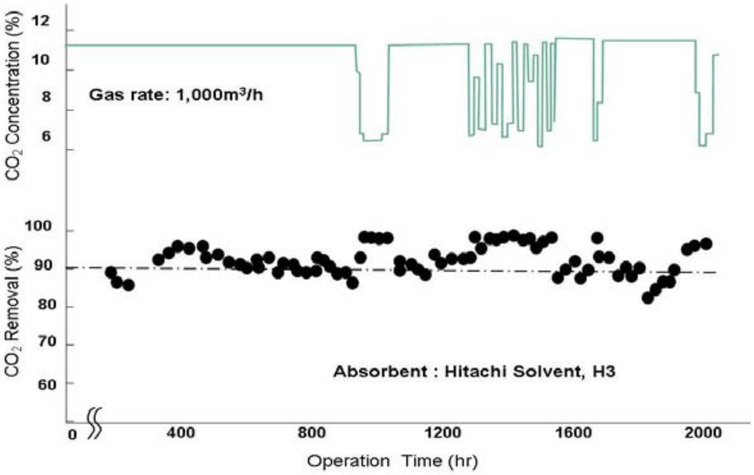


Figure 4. Long-term pilot testing of H3 solvent under various inlet CO_2 concentrations

Large Mobile Pilot Plant

Based on earlier experience in Japan, Hitachi Power Europe GmbH, in cooperation with utility partners Electrabel / GDF Suez and E.ON., is building a large mobile pilot plant for the separation of carbon dioxide from coal-fired power plant flue gases. The plant is used to generate data for the development of design concepts for both new power plant integrated with CCS or retrofit of a carbon dioxide separation plant in existing power stations.

Figure 5 shows the general arrangement of the mobile pilot plant. The plant components are built into transportable segments equivalent to overseas containers. These preinstalled segments, shown in Figure 6, are erected at the site of the host power station and connected to a slipstream of flue gas after the plant's FGD unit. The pilot plant is designed to process a flue gas volume flow of approximately 5000 Nm³/h corresponding to 5 MW_{th}. The auxiliaries and steam needed for the operation are supplied by the host power plant.

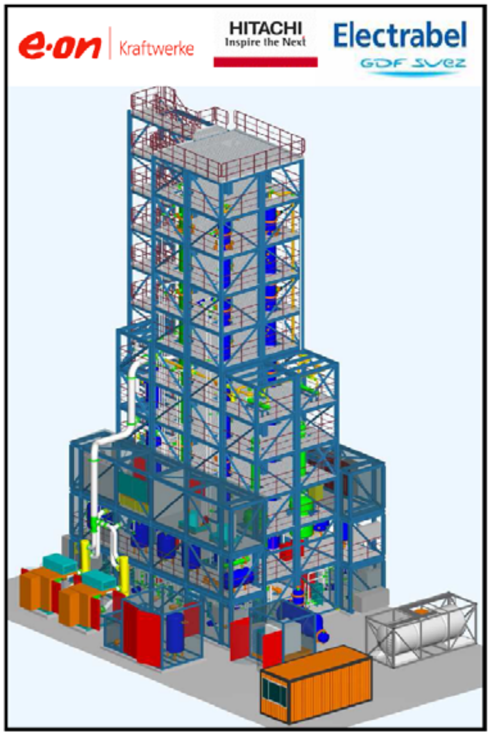


Figure 5. Pilot plant arrangement

Figure 6. Mobile pilot plant segments

Figure 7 is the process schematic of the pilot plant. Flue gases after the wet FGD of a typical existing power plant in Europe still have 200 mg/Nm³ SO_2 (~ 0.15 lb/MMBtu) and about 10 mg/Nm³ SO_3 (~0.08 lb/MMBtu). For further cleaning, a caustic soda pre-scrubber is installed upstream of the two carbon dioxide absorbers. Behind the pre-scrubber, the flue gas flow is uniformly distributed to two trains of absorber- desorber system with amine solution as the scrubbing reagent for CO_2. The two train design allows three different operation configurations :

- Parallel operation 2 x 2500 Nm³/h,
- Single line operation 1 x 2500 Nm³/h,
- Serial operation of two absorbers followed by one desorber.

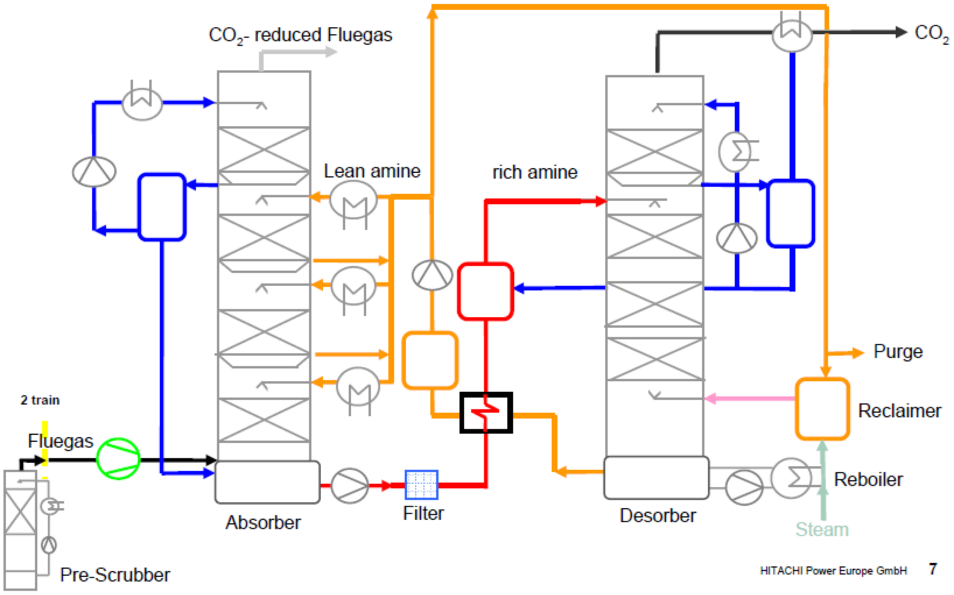


Figure 7. Flow sheet of the pilot plant

The ability to arrange the pilot plant in three configurations offers maximum flexibility. The pilot plant thus will be able to run under a very wide range of operating conditions such as process gas flow, residence time, CO_2 removal rate, allowable solvent carryover, etc, and produce reliable data for the scale-up to full size plants.

The primary objective of the pilot program is to investigate the lowest energy demand and to determine design criteria for the optimization of capital investment and operating costs. Besides this the design data to minimize the environmental impact of the CO_2 scrubbing system is of great interest. The pilot plant is fully equipped with an on-line monitoring system to continuously measure trace emissions of solvent as well as some of the degradation products. In addition, periodic campaigns are scheduled for the complete measurement of degradation product emissions.

The main parameters for testing will be:

- Investigation of the effect of solvent inter-cooling,
- Operation with different solvents,
- Influence of enhanced HSS formation by decreasing pre-scrubber efficiency,
- Investigation of increased flue gas temperature by decreasing pre-scrubber efficiency,
- Operation with different washing stage efficiencies to investigate the influence on the emissions.

Scaling up to Full Size Plant

The pilot plant has an absorber diameter of about 1 m. Currently, absorbers with diameters up to 12 m have been applied in the chemical industry. For an 800 MW plant a single absorber would need a diameter of roughly 19 m. There is a need for intermediate steps from pilot plant sizes to 800 MW_{el} full size plants. Figure 8 shows possible steps in size from a pilot plant to full size equipment.

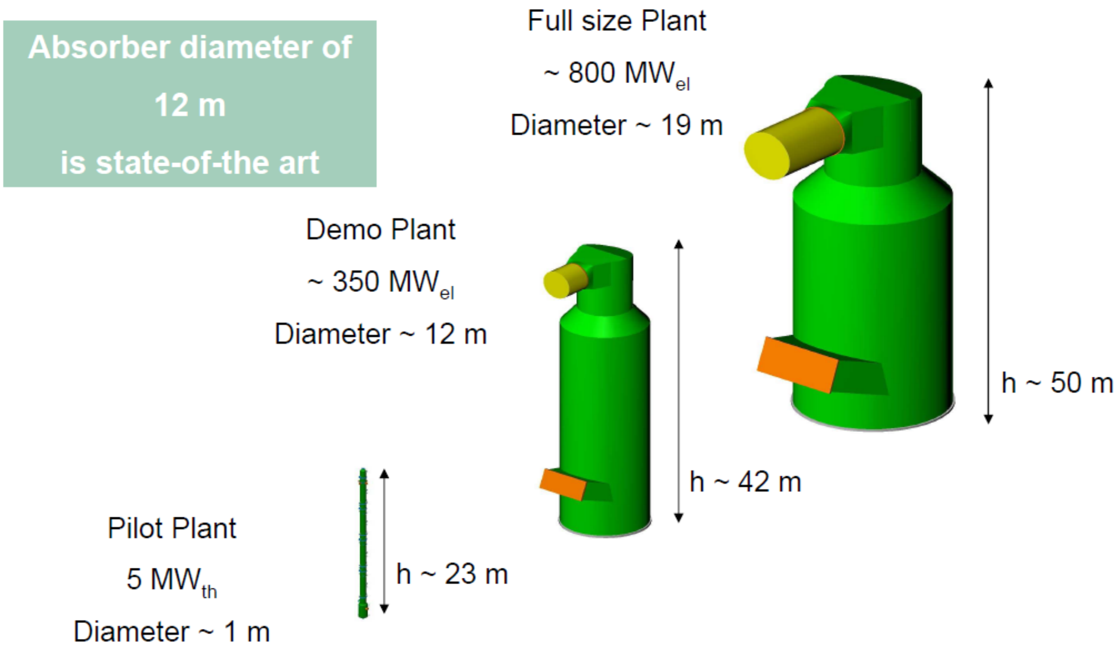


Figure 8. Steps from pilot plant to full size plant

INTEGRATING CO_2 CAPTURE INTO A POWER PLANT

The Reference Plant

A state-of-the-art, supercritical pulverized coal-fired power station currently under planning stage is selected as the reference case for the plant integration study. As shown in Figure 9, it has a single reheat steam cycle with a main steam temperature of 596°C (1105°F) and reheat temperature of 608°C (1126°F). The plant with direct sea water cooling will have a net efficiency of about 46% on a lower heating value (LHV) basis. This plant is representative of the 600°C class supercritical PC power plants that have been widely deployed in Europe and Asia.

Currently, the worldwide average of power plant net efficiency is about 30% LHV, corresponding to a carbon intensity of approximately 1100 g CO_2/kWh. The 600°C class plants, considered as the best available technology (BAT) in Europe, emit about 720 g CO_2/kWh. The next generation advanced ultrasupercritical power plant under development by Hitachi and others, will feature a 700°C (1292°F) steam cycle and reach a net efficiency in the 50% LHV range, further reducing carbon emissions to about 660 g CO_2 / kWh.

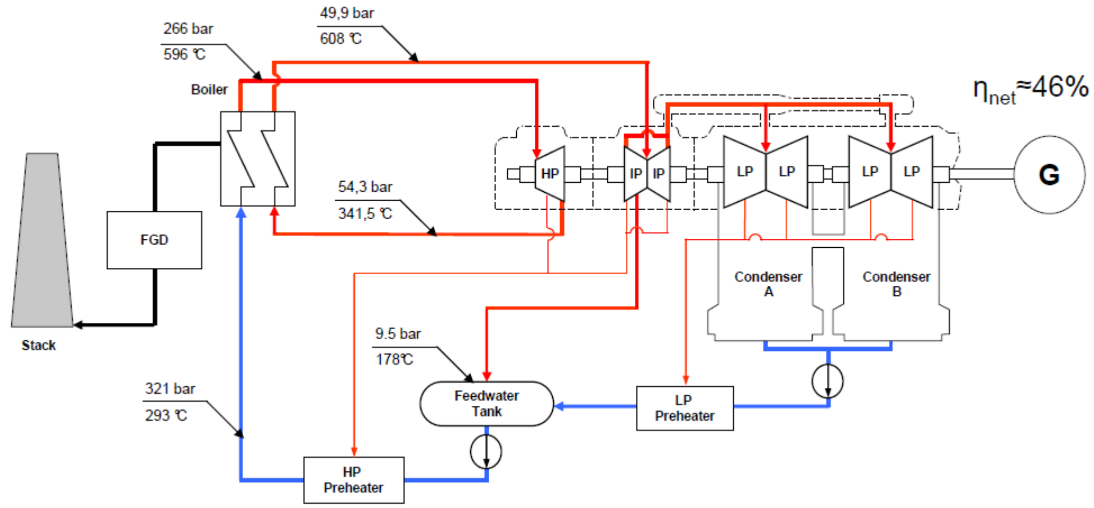


Figure 9. Heat balance diagram of an 800 MW state-of-the-art power plant

Requirements of CCS integration

Implementation of CO_2 scrubbing into a power plant poses enormous challenges to the design of the power plant itself and the post combustion capturing system respectively. Several interfaces have to be considered and optimized which necessitate modifications to the plant components. Figure 10 shows these main interfaces.

Aside from using the best solvent with the lowest regeneration energy requirement, the overall net plant efficiency of a power plant with CCS can be maximized by optimizing the integration of available heat sources and heat sinks across the entire plant system including the CCS scope. In addition to process and electrical interfaces, the required space for the carbon capture plant can be a problem especially for retrofits into existing power plants. The minimization of the required space is a major task for both process design and arrangement planning.

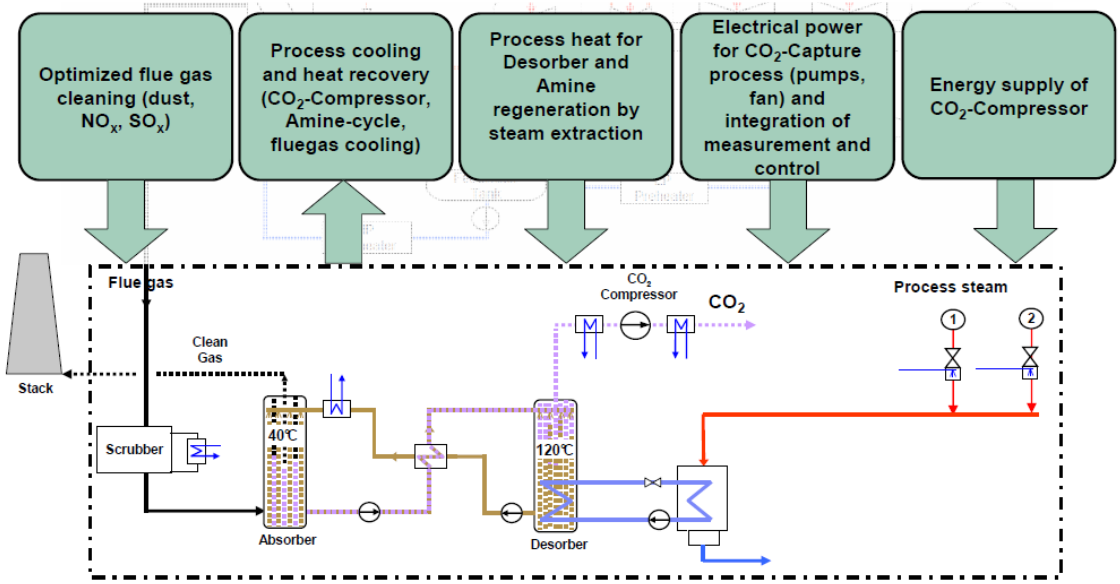


Figure 10. Interfaces between a post combustion capturing system and a power plant

Optimization of Flue Gas Cleaning System

Amine based solvent solutions are known to react readily with SO_2 and SO_3, resulting in the formation of heat stable salts and amine degradation. Generally, in the flue gas entering the CO_2 absorber, the combined SO_2 and SO_3 concentration need to be 10 ppm or less to avoid excessive solvent loss in the CO_2 capturing process. To reduce SOx emissions to this level, several approaches discussed below can be considered.

Additional Polishing Scrubber vs Single-Stage Wet FGD

For plants with older FGD units and high outlet flue gas SOx concentration, a separate polishing scrubber, also called "pre-scrubber" for the CO_2 absorber, should be considered. For existing FGD with very low SOx emissions, further reduction of SOx to the level required by CO_2 scrubbing may be achieved by upgrading FGD internals, adding spray level(s) to the existing unit, and/or applying organic acid to the FGD slurry as pH buffer to enhance SO_2 removal performance.

Although the vast majority of existing dry or wet FGDs in the world today operate with outlet SO_2 well above 10 ppm, Hitachi's open spray tower wet FGD technology with its CFD-guided design of high spray flux and variable spray density to prevent local flue gas sneakage, is capable of achieving single digit ppm SO_2. In fact six (6) Hitachi wet FGD units, including one unit recently commissioned in the United States, are in commercial operation with SO_2 removal efficiencies well above 99% and FGD outlet SO_2 well below 10 ppm. Five of these units have outlet SO_2 concentrations in the low single digits. These units were designed for ultra-low SO_2, not for CCS, because of stringent emissions regulations at these plant locations which are often due to the very high population density in the nearby areas. These "single digit" FGD units are treating flue gas from fuels with very low to very high sulfur content, including two units in Japan treating flue gas from high sulfur petroleum coke.

The "single digit" wet FGDs are designed with large numbers of spray levels and high liquid-to-gas ratios. Therefore, they have higher capital and operating costs than ordinary FGDs. However, for new plants or existing plants building a new FGD, a single stage wet FGD can be more advantageous than a combination of a primary FGD and a polishing FGD.

Clean Energy Recuperator (CER)

SO_3 can be a significant contributor to total SOx emission, especially for plants burning higher sulfur coals. As flue gas enters the FGD absorber, SO_3 condenses into fine mists which are only partially captured in the wet FGD. One way to achieve high SO_3 removal rate is by adding a wet ESP downstream of the FGD absorber. Alternately, hydrated lime, or sodium based sorbent in either powder or solution form can be injected upstream of the dust collector to capture SO_3.

CER is a technology developed by Hitachi that can effectively remove SO_3 and can contribute significantly to efficiency improvement of a power plant through recovery of flue gas waste heat. CER is derived from Hitachi's patented high dust Gas-Gas-Heater (GGH) technology which has been successfully applied to five large supercritical coal-fired power plants in Japan. In the United States, CER has also been demonstrated at a slipstream test facility at Ameren's Duck Creek Power Station burning US bituminous coals. Shown in Figure 11 as a part of the air quality control system with a wet FGD, the CER heat recovery module is located between the air preheater and the dust collecting device (ESP or Fabric Filter). The CER is a finned tube heat exchanger with water as the cooling medium. It cools the flue gas from more than $150^{o}C$ ($300^{o}F$) to about $88^{o}C$ ($190^{o}F$) and recovers a large amount of low grade energy. When integrated for turbine condensate heating, this recovered energy can reduce the power plant heat rate by about 1%. This low grade heat can also be used as a heat source for the CO_2 desorber to avoid a part of auxiliary steam (energy) consumption.

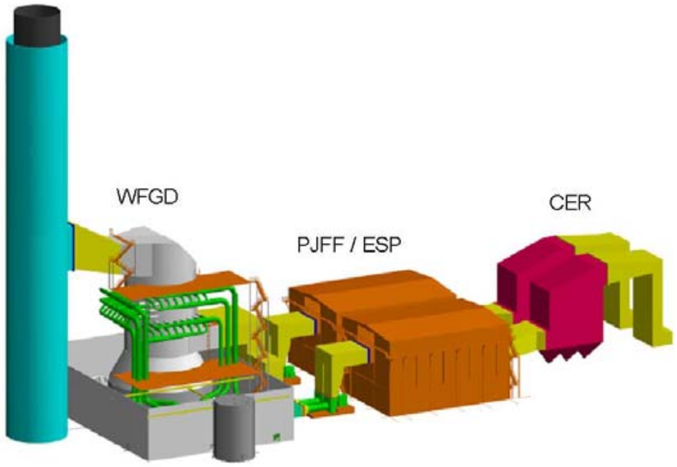


Figure 11. Hitachi Clean Energy Recuperator

Due to its operation in high ash environment and the deep cooling of flue gas, the CER removes almost all SO_3 in the flue gas. By cooling down the flue gas indirectly, water consumption for the wet FGD can be reduced by about 50%.

The selection of AQCS configuration will be on a case-by-case basis, depending on factors such as fuel sulfur content, desired level of SOx entering the CO_2 absorber, and costs for various reagents. With a rational combination of the technology options discussed above, the performance of the air quality control system will certainly meet the requirements for CCS technology without adverse effects on other parts of a coal fired power plant.

Process Integration of CO_2 Capture System

Conventional Approach

The conventional approach does not provide maximum integration of the CCS system with the steam cycle. Instead the CCS system and the steam cycle are optimized independently. The energy needed for the regeneration of the CO_2 rich solvent is the key issue for the performance and efficiency losses. Most of this energy will be provided by steam extraction from the power plant. A conventional approach for the steam extraction from the water steam cycle is shown in Figure 12. The steam for desorption is extracted at the hot or cold reheat steam line (1) or the crossover pipe (2) between the IP and LP turbine. As a result, very high losses in efficiency cannot be avoided.

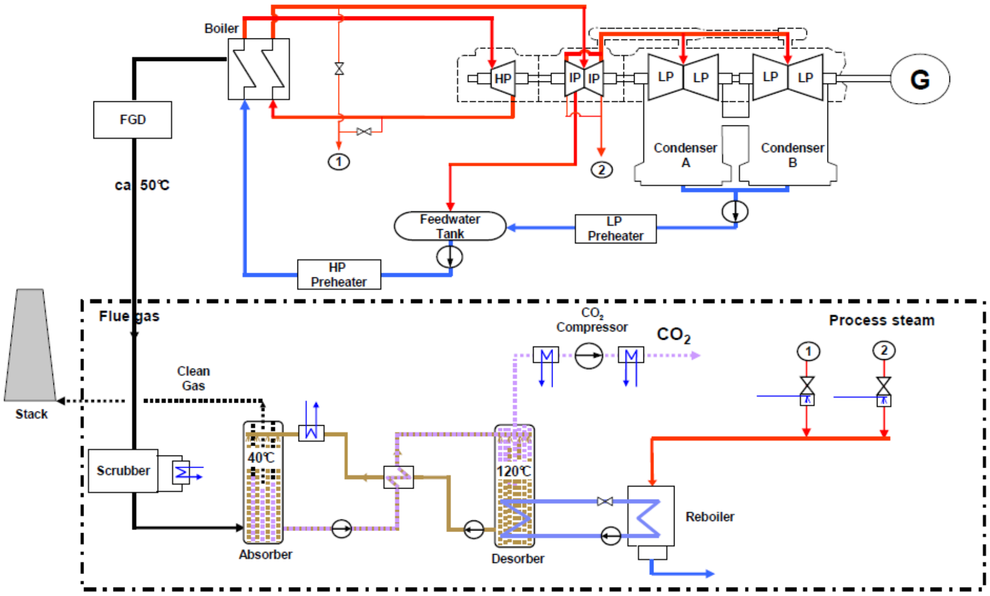


Figure 12. Water-steam cycle of a power plant with CCS process integration

This conventional approach for CO_2 capture process implementation will have a power plant net efficiency loss of about 13.1% points if MEA with a specific regeneration heat 3600 kJ/kgCO_2 is used. If the advanced solvent H3 (2800 kJ/kgCO_2) is used, the loss will be 9.1% points. An additional loss of efficiency of 2.8% points due to CO_2 compression has to be added, assuming that the CO_2 is compressed to 200 bar at 30 °C (2900 psi at 80 °F). The basis for the efficiency comparison is the reference plant described earlier. When firing a hard coal, the plant has a net efficiency of 46.9 %LHV, without CCS.

Advanced Approach

Steam cycle modifications: Due to the enormous efficiency losses of the conventional approach, a better way to integrate the post-combustion capture plant into the power plant must be found. To increase the efficiency, an optimization of the plant overall process is needed, i.e., all heat sinks of the CCS system needs to be introduced at an optimum location of the steam cycle so that no energy is wasted. This can be accomplished by integration of the condensate and the cooling water from the CCS process into the water steam cycle as well as the steam extraction from the steam turbine for the reboiler heating. Figure 13 shows an optimized water steam cycle of a power plant.

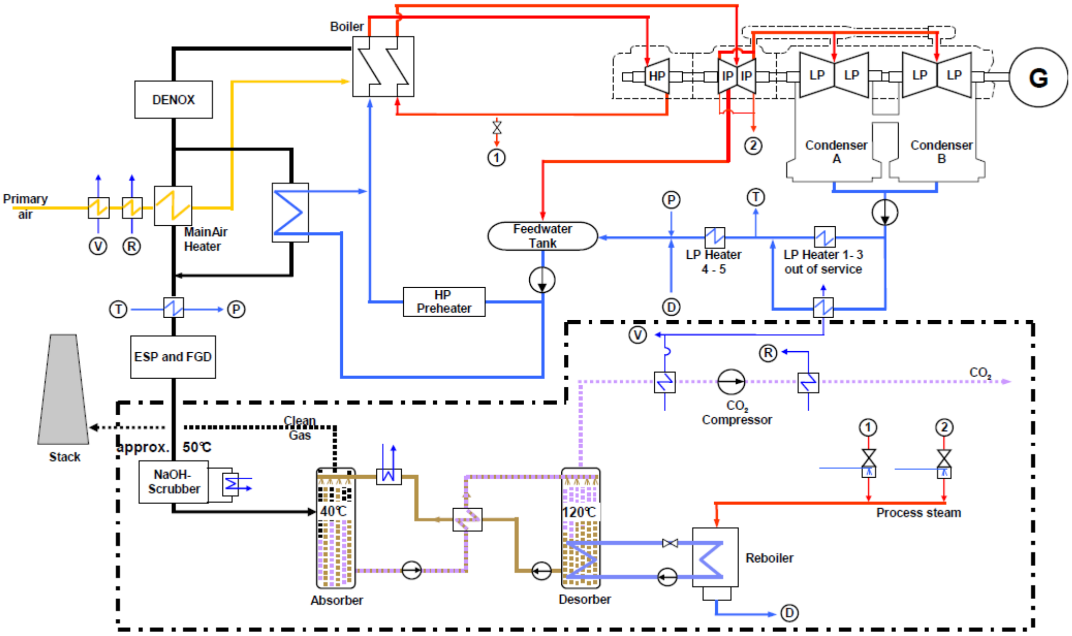


Figure 13. Optimized water-steam cycle of a power plant with CCS process integration

The modifications of the water steam cycle are as follows:

- The condensate from the reboiler heating steam is reintegrated into the main condensate line downstream of the condensate preheater No. 5.

- The CCS process has a large demand for cooling water capacity. A part of the heat transferred by the CCS system to the cooling water can be recovered to warm up the condensate upstream of the feedwater tank/deaerator. As a result, the LP heaters No. 1-3 can be bypassed and unloaded, which results in increased steam cycle efficiency.

- Another possibility is a modification of the air preheating and flue gas system which allows the use of the waste heat of the CCS process to warm up the feedwater as well as main condensate. A part of the waste heat from the CO_2 cooling at the desorber outlet (V) and the waste heat of the CO_2 compressor (R) are used for air preheating before entering the main air heater. Since these waste heats are used for air preheating, a part of the flue gas heat can be shifted to the feedwater line by using a heat exchanger in parallel to the main air heater. The remaining heat amount of the flue gas downstream the main air heater can be used for main condensate preheating (P).

- The thermodynamically preferred location for the steam extraction is the crossover pipe (2) between the IP- and LP- steam turbine. This extraction can be partially switched to the cold reheat steam line (1) to increase the pressure of the extraction steam, if necessary.

Steam turbine modifications: The large amount of heat for the regeneration makes modifications to the standard steam turbine design necessary. For 90% CO_2 separation about 25-30% of the live steam flow or approximately 60% of the exhaust steam flow has to be extracted. Depending on the possibilities for steam extraction at the steam turbine the following design considerations are required:

Regardless whether the steam is extracted from the crossover line between IP- and LP-steam turbines, the reheat steam line or both, the blades of the HP- and IP-turbine must be designed for the increased pressure/enthalpy drop across all stages. The casted outer casing of the IP- turbine must be designed according to the increased mass flow of the steam extraction for the CCS process. The LP turbine must be able to accommodate large flow variations due to the process steam extractions (in some cases, the steam turbine will also have to be able to continue operation with no process steam extractions, when the CCS is not in operation.) The length of the last stage blades (LSB) of the LP- turbines must be optimized according to the new exhaust steam flow requirement, (which is less with CCS). Operation with CCS will require shorter LSBs to avoid excessive exhaust losses due to ventilation and low load operation. Optionally a crossover valve between the IP- and LP-turbines can be used to reach the required steam pressure for the supply of the heat quantity for the CCS process. The crossover valve maintains a constant steam pressure on IP- turbine outlet and the extraction stub, which would minimize the modifications required for HP and IP turbine design. However, the crossover valve itself creates its own design challenges that need to be considered and moreover, throttling losses of the crossover valve will decrease the cycle efficiency.

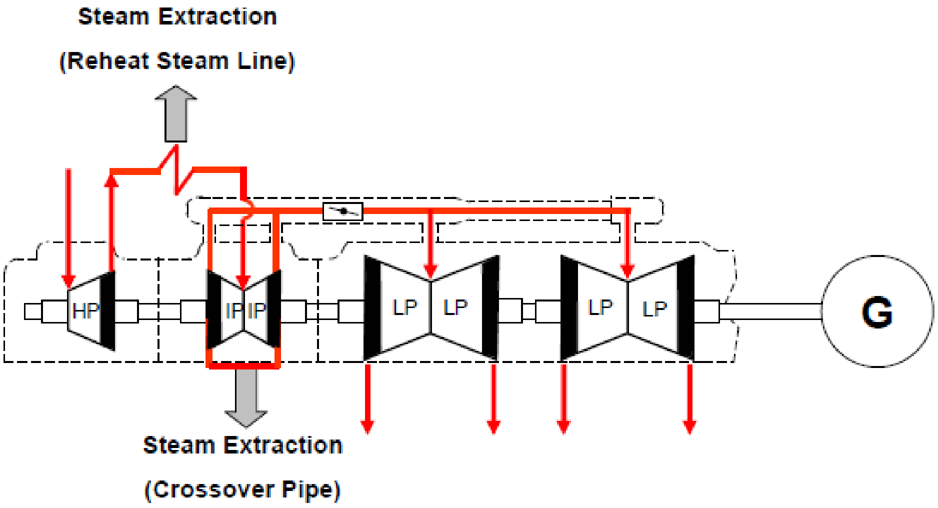


Figure 14. Modifications of the steam turbine

The above mentioned design requirements can be considered in the planning phase for new power plants. In case of existing plants, the required modifications at the steam turbine for the steam extraction might be executed with a turbine retrofit.

Impact on net plant efficiency: As a result of the modifications for the water-steam cycle and the steam turbine, the loss of net efficiency is reduced to only 7.8 % points for H3 and 7.5 % points for next generation solvent (NGS) with regeneration energy of 2500 kJ/kg CO_2 (including CO_2 compression to 200 bar). The efficiency loss for the conventional approach discussed earlier is also plotted in Figure 15 for comparison.

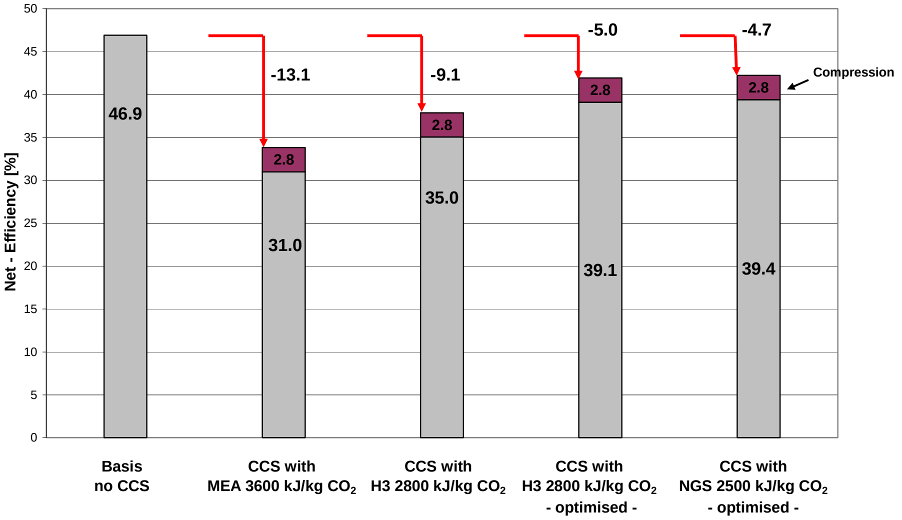


Figure 15. Results of optimization

Arrangement of the Power Plant with CCS

The area needed for the CCS plant depends partly on the maximum size of the absorbers and desorbers. The state-of-the-art absorber diameter of 12 m means that the flue gas of a 350 MW$_{el}$ demo size plant can be treated. Therefore the minimum number of trains required for an 800 MW$_{el}$ full size plant is three. With further development of large single absorber diameters the required space can be reduced by using fewer absorbers and/or desorbers. In Figure 16 possible arrangements with two to three absorbers and one to three desorbers are shown. With these arrangements the installation area can be reduced by up to 25% from the base arrangement with 3 absorber and 3 desorbers. Figure 17 shows the arrangement of the CO$_2$ capture island with the optimized design.

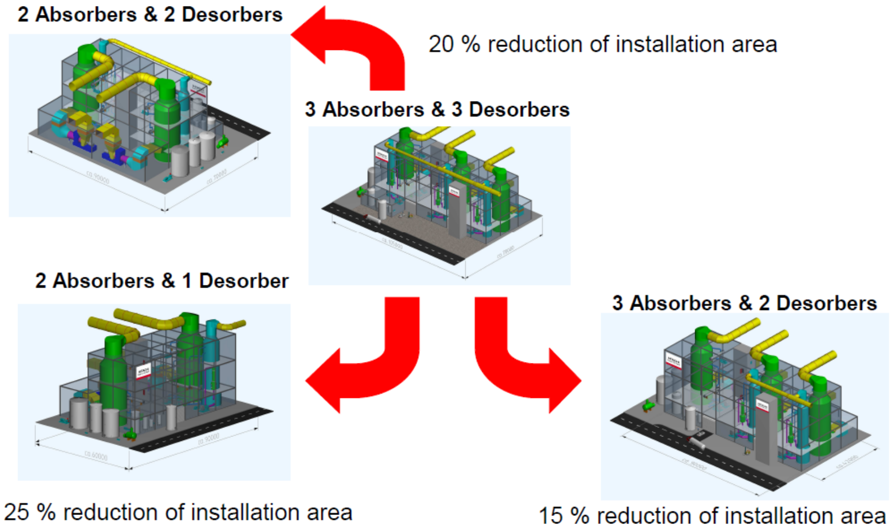


Figure 16. Possibilities for the reduction of installation area

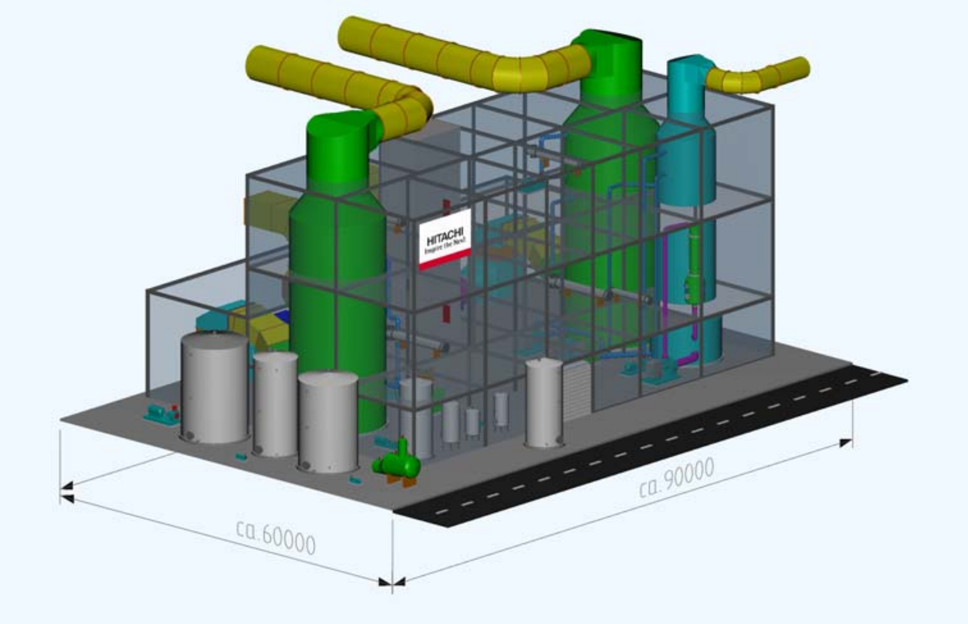


Figure 17. Arrangement for an 800MW with two absorbers and one desorber

In Figure 18 a layout plan for the integrated power plant with a post-combustion capture system is shown. The footprint of the CO_2 capture island is about 60 m x 90 m (197 ft x 295 ft) for the 800 MW electric plant, when using 2 absorbers and 1 desorber.

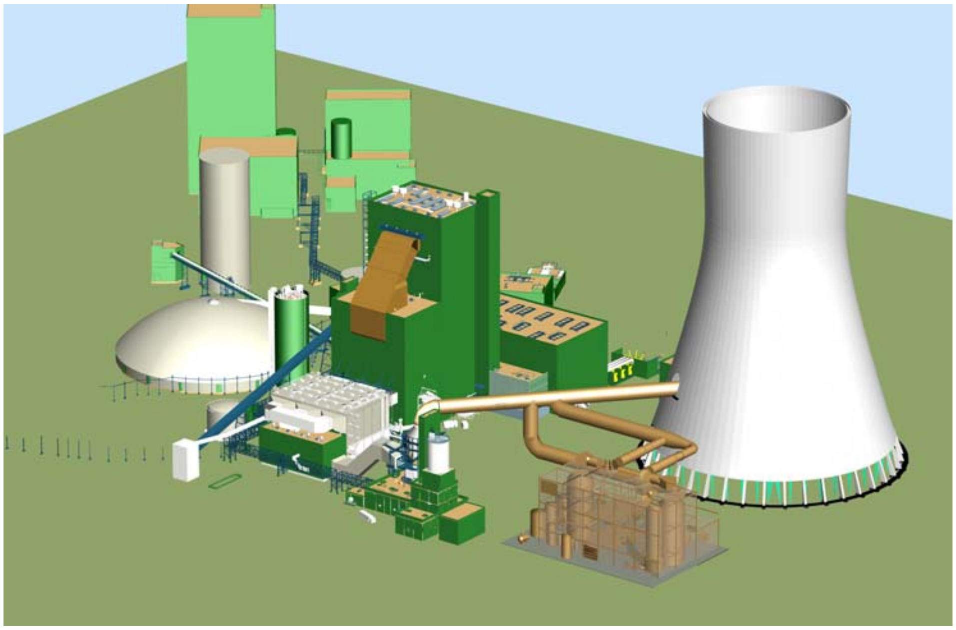


Figure 18. Integration of a post combustion capture plant into a power plant station

POST COMBUSTION CO_2 CAPTURE SUMMARY

For application of post-combustion CO_2 capture at a coal-fired power plant, several technical challenges have to be solved. Use of advanced amine-based solvent can alleviate solvent loss due to impurities in the flue gas. In addition, flue gas cleaning system upstream of the CO_2 absorber has to be optimized. Hitachi has developed an advanced FGD technology capable of ultra-low SO_2 emission levels. A polishing scrubber can also be used when necessary. The Clean Energy Recuperator can effectively remove SO_3 and improve plant efficiency by recovering flue gas waste heat. With these alternatives, SOx outlet concentrations of less than 10 ppm will be achievable for new plant or retrofit applications. Further reduction of amine loss can be realized by using a solvent reclaiming process. With these technical measures the power plant operator will be in a position to find customized and optimum system configuration and operating practices.

Hitachi is building a large mobile pilot plant in cooperation with Electrabel S.A./GDF Suez and E.ON. The pilot plant has a flexible two train, three configuration design that can be operated under a very wide range of operating conditions such as gas flow, residence time, CO_2 removal rate, and allowable solvent carryover. The flexible design also will be able to achieve desirable amine emission levels to meet existing or anticipated regulatory requirements. This large pilot test program will provide valuable data to support the design of demonstration and commercial plants.

A primary target of system level optimization is to reduce efficiency loss, through an approach with three essential elements. The first is the use of the latest solvent with low regeneration energy and resistance to degradation, and continuing development of next generation solvent with further improved performance. The second is rational redesign and modifications of the steam turbine to supply the heat for CCS with minimum loss of electricity output. The third integration area involves boiler heat recovery, combustion air preheating, feedwater heating, and turbine condensate preheating in the context of CCS implementation. By applying these measures, plant efficiency loss can be reduced to less than 5% points (excluding CO_2 compression). Hitachi is currently undertaking extensive global development and demonstration programs to improve basic technologies for CO_2 scrubbing and to accelerate its commercialization for coal-fired power plants worldwide.

3. OXYFUEL COMBUSTION TECHNOLOGY

Oxyfuel combustion produces a flue gas stream containing mostly CO_2, which can be directly compressed and purified without further treatment, assuming upstream removal of other pollutants, such as SO_2, NOx and dust. In the process shown in Figure 19 the CO_2 concentration in the flue gas is greatly increased by using a mixture of recirculated flue gas and pure oxygen instead of air for firing coal. Recirculation of flue gas is necessary to provide sufficient mass flow of flue gas for cooling the flame and also heat capacity and flue gas velocity for convective heat transfer in the boiler.

In the oxyfuel process CO_2 purity is mainly influenced by (a) where the flue gas is recycled in the process (the cleaning that has been done up to this point) (possibilities: 1-6 according to Figure 19), (b) the sealing of boiler and other components to prevent air ingress, (c) the purity of the oxygen from the Air Separation Unit (ASU), (d) the performance of all air quality control system equipment (SCR, FGD, and ESP), (e) additional CO_2 purification during/after compression.

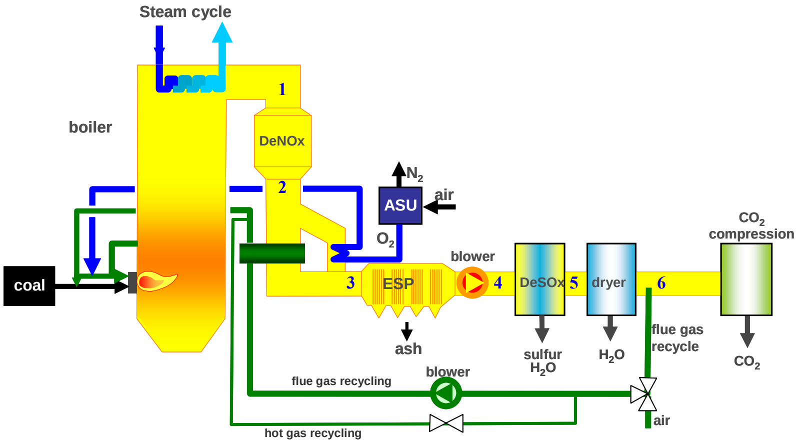


Figure 19: The oxyfuel process

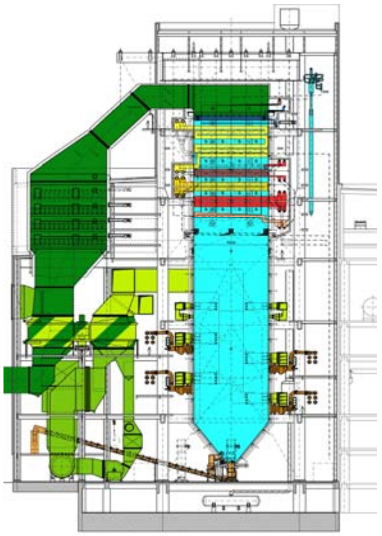
The objective of our current work is to develop a retrofit design that can be applied to the coal-fired power stations commissioned in recent years or in the near future to enable CCS operation. A special characteristic of the retrofit measures is that the power station can be operated both with oxygen and air firing after the retrofit. As a result, the plant can be started and shut down in air-firing mode. Also in the event of operational trouble with the new systems, such as CO_2 compression, transport and storage, switching to air firing can be done quickly to ensure reliable electricity supply.

In the following sections, the required plant modifications, expected boiler performance and plant system impact of oxyfuel conversion, as well as CFD modeling and pilot test studies to support such conversion will be discussed.

DESIGN STUDY AND MODELLING

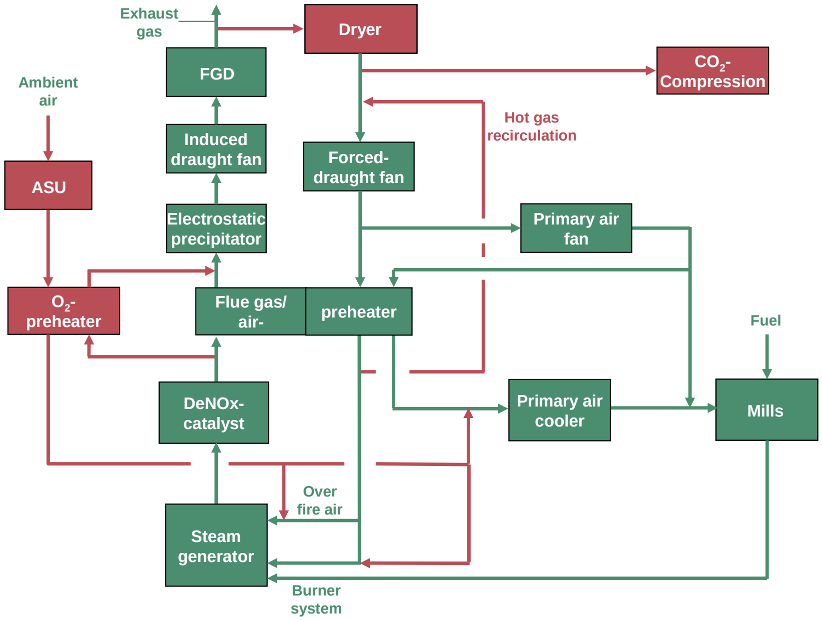
Basis of Design Study

The retrofit design study has been performed based on a state-of-the-art 600°C (1112 °F), 820 MWe power station currently under construction; see Figure 20. During air firing, the combustion air is heated in the flue gas air-preheater and distributed to mill (primary air), burner and over-fire air (OFA) ports. A portion of the primary air enthalpy is used for feedwater preheat in the "mill air cooling cycle". This increases the overall efficiency of the power station process by minimizing cold air to the mill and by reducing the flue gas exit temperature. Additionally, less steam needs to be extracted from the turbine. After combustion the flue gas is cooled in the air preheater, NOx concentration is decreased catalytically, dust is removed in the ESP and SO_2 is removed using limestone in a wet scrubber. In the boiler the steam is mainly heated by radiation in the furnace and by convective heat transfer in superheaters, reheaters, and economizer in the convective pass.



(a) Once-through steam generator, Benson®
820 MWe, 2088 t/h steam
SH: 600 °C / 276 bar a (1112 °F / 4003 psia)
RH: 620 °C / 51 bar a (1130 °F / 740 psia)
Commissioning: 2011/2012
Design coal: Bituminous Coal

LHV:	25100	kJ/kg
Ash:	13	%
Water:	10	%



(b) Firing process

air case = green components and lines

oxyfuel case = red components and lines added

Figure 20: Power station and design coal (a) and firing process (b)

Modifications for Oxyfuel Firing

Looking at the different options for flue gas recirculation (locations 1-6, Figure 19) it is obvious that the complexity of flue gas recycling is reduced from the progress made in flue gas treatment. Recycling high-temperature flue gas (before air preheater) is thermodynamically advantageous but requires a total change of the heat balance and a re-design of the plant (boiler and components). A high dust recirculation upstream of the ESP would increase the erosion of all firing and boiler parts. Without FGD the SO_2/SO_3 concentration would be increased by accumulation (~factor of 3) and additionally SO_3 formation would increase from contacting with catalytic surfaces. Therefore all firing and boiler components (flue gas and recycle gas ducts, blowers, mill, burners, heat exchangers, boiler materials) would be at risk from corrosion.

For all of these reasons the retrofit concept shown below is based on the recirculation of cold, cleaned and partially dried flue gas after FGD and additional flue gas cooling. This would allow all existing components including blowers and ducts – at least inside the boiler house - to be used after the retrofit. The only measure to be taken inside is to preheat the recycled flue gas up to a temperature well above the saturation point so as to avoid condensation, by re-circulating a small amount of hot flue gas. Most of the necessary retrofit measures will be implemented outside the boiler house. Switching between oxyfuel and air operation mode can be done simply by using gas tight dampers at the former air inlet where the recycle duct is mounted.

Other changes outside the boiler house involve pure oxygen oxidation in the FGD plant and the addition of a flue gas cooler/condenser (150 MW_{th}, cooling to 30°C) upstream of flue gas recycling which leads to further reduction of the SO_2/SO_3 content. Downstream of the dryer the flue gas is split and one flow is directed to the boiler house. The duct is connected to the outside air inlet with leak-tight dampers which also provides air in the case of air firing. For oxygen preheating a tubular preheater parallel to the air preheater (now used as gas/gas preheater for the recycle gas) will be installed. By total shutoff of the oxygen preheater, air firing can be enabled even after the retrofit.

The purge gas of the mill is switched to CO_2 and the sealing is retrofitted to ensure that no CO_2 enters the boiler house. The atomizing gas for the aqueous ammonia in the SCR is also replaced by CO_2 and the ash removal at the ESP is replaced by a gas tight system. A retrofit of the boiler ash removal system is not necessary since a minimum ingress of air is already ensured by a wet ash removal system which was implemented during the air firing design.

The modifications minimize the overall leakage/injection of air/nitrogen to the flue gas to 1% of the flue gas mass flow in the furnace. In the oxyfuel mode therefore the flue gas contains about 95 %wt of carbon dioxide. The mass flow is reduced to 25 % of the flue gas flow leaving an air-fired furnace. Purification and compression of this CO_2 rich flue gas is the last step of the oxyfuel process. During compression some of the remaining contamination gases, such as NOx and SOx, are separated and leave the process as condensate from the intermediate coolers as sulfuric and nitric acids. Most of the water is also removed from these intermediate coolers. Normally there is no other purification step since other trace gases such as nitrogen, oxygen and argon can remain in the compression stream for storage. To also remove these gases (depending on the use/storage option) cryogenic separation would be necessary. However, water has to be removed down to very low values to prevent corrosion in pipelines and tanks.

The only commercially available and proven technology today to supply an 820 MWe bituminous coal fired power plant with oxygen is the cryogenic process. In this case up to 13,500 t/d (including 10% reserve) of oxygen is required. The available size and required control range of air separation units (ASUs) will make 4 ASU lines necessary. An important issue is the load change rate of an ASU which typically is limited to 1%/min but the rate of a power plant is up to 5%/min. To compensate for this difference, a temporary buffer storage is necessary.

Combustion Process and Thermal Performance

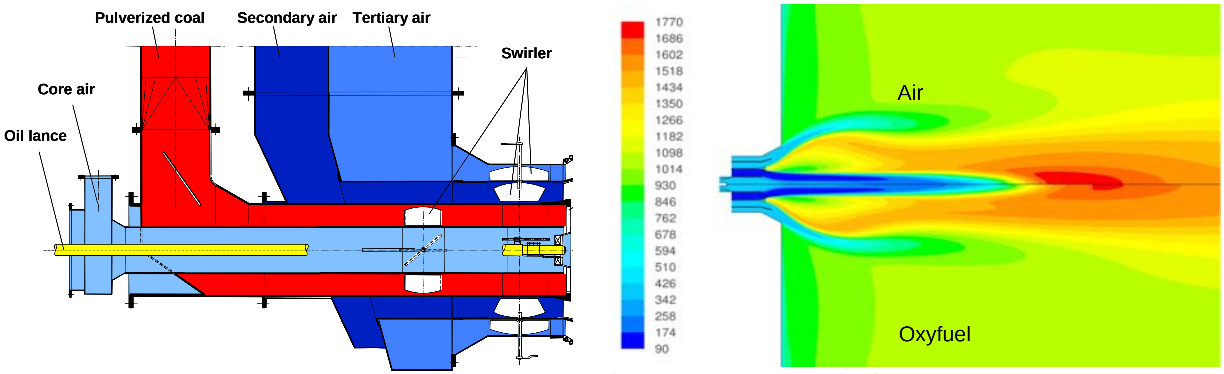
To avoid significant changes to the power plant heat balance, it must be assured that the heat transfer in furnace and heat exchangers in the convective pass match the original design. Additionally, the material temperatures have to be kept in a tolerable range and the steam temperatures and pressures should match the air combustion case. These requirements are fulfilled by the adjustment of the mass flow of recycled flue gas and the split in gas streams for burner (primary gas and other), over-fire and curtain gas as well as the adjustment of the oxygen content in different gas flows. The furnace exit gas temperature is set according to the upper limit given by the ash melting temperature.

Under oxyfuel conditions the firing components (mill, burner) as well as the heat transfer in furnace and heat exchangers are reevaluated with respect to the changed flue gas properties so as to determine the optimal process parameters. For the retrofit case the modifications of existing components are minimized to reduce plant outage time.

Firing System

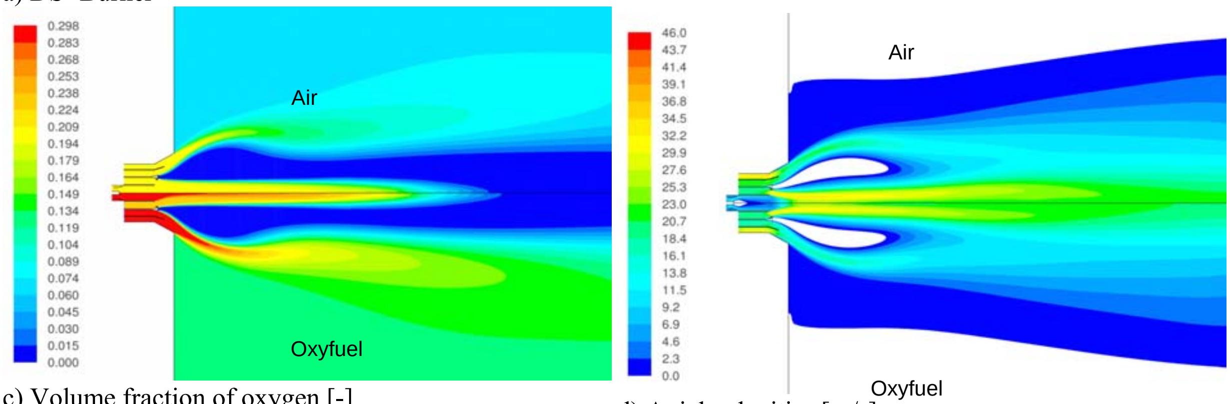
An important criterion for mill operation is the discharge of coal particles by flue gas. The dominating force here is the drag force which depends on temperature and composition of the carrier gas and therefore the mass flow required for the transport of the fuel particles depends on location. The calculations proving that the gas streams can carry the fuel particles have to be carried out at least at the mill nozzle ring, upstream and downstream of the classifier and in the ducts. The volumetric flow has to be the same as for air firing operation to get a reliable and steady flow downstream of the classifier. The lower velocity at the nozzle ring in this case is compensated by the higher gas density.

The power station studied here uses Hitachi Power Europe's low NOx DS burners (Figure 21a). Except for the primary gas flow, which has to be adjusted to the mills' needs, the momentum gas flows at the burner are kept constant in the retrofit case so as to get a flame shape similar to that in the air firing mode.



a) DS- Burner

b) Temperature distribution [°C]



c) Volume fraction of oxygen [-]

d) Axial velocities [m/s]

Figure 21: Low NOx DS burner (a) and results from CFD calculation results (b-d)

The flame temperature and burnout progress are adjusted to fulfill the needs of heat transfer utilizing the oxygen concentration as a variable. This is demonstrated by the findings from a rotational symmetric flame calculation.

Figure 21b depicts the difference between the temperature distribution of air-fired and oxyfuel-fired flames. Whereas the upper half shows the temperature distribution of the air-fired flame, the lower half reveals the findings for oxyfuel conditions. The difference is very small under the examined conditions. This shows that the chosen parameters for the oxyfuel case are appropriate for getting firing conditions inside the furnace similar to those under air-fired conditions.

It should be pointed out that the flame temperature is affected by the volume fraction of oxygen in the oxygen carrying gas. The volume fraction is shown in Figure 21c. Here the oxygen concentration is chosen to match temperature distributions of air and oxyfuel firing as shown in Figure 21b. The often discussed decrease of oxyfuel flame temperature is countered by increasing the volume fraction of oxygen in the oxygen carrier gas.

Comparison of the axial velocities (see Figure 21d) shows that the rule for the momentum for the oxyfuel conditions really does result in similar flow fields for air-fired and oxyfuel flames. In this figure, the white zones represent backflows. The backflow zone behind the tooth-ring of the primary air tube has nearly the same shape in both the air-fired and oxyfuel modes. This backflow zone is essential for burner ignition purposes.

Heat Balance

The mass flow of gas in the OFA system is adjusted so that the heat transfer in the convective pass matches the values of the original design. As is shown in Table 1 the flue gas density in oxyfuel firing is increased by 35%, the heat transfer coefficients for convective and radiative heat transfer by 2.7 and 38.3% respectively and the flue gas mass flow by 3.9%. These increases are partially compensated by a 17K or 5.2% decrease of the logarithmic temperature difference. The flue gas recycling rate is 75.1% and the overall stoichiometric factor is 1.17 (upstream of air heater). This is equivalent to an excess of oxygen at the end of the furnace of 2.86 wt.% wet.

Table 1: Heat Transfer in Convective Pass (average values for entire convective pass).

	Air Mode	**Oxy Mode**	**Unit**	**ΔOxy / Air Mode [%]**
flue gas density	1.33	1.80	kg/m³ (STP)	35.0%
heat capacity of flue gas	1.23	1.29	kJ/kgK	4.7%
dynamic viscosity of flue gas	42.87	40.84	µPa s	-4.7%
heat conductivity of flue gas	0.07	0.07	W/mK	2.8%
mass flow of flue gas	716.7	744.9	kg/s	3.9%
maximum flow velocity	10.70	8.00	m/s	-25.2%
ΔT(logarithmic)	318.2	301.8	K	-5.2%
heat transfer coefficient, outside convective	38.88	39.91	W/m²K	2.7%
heat transfer coefficient, outside radiation	34.98	48.36	W/m²K	38.3%
overall heat transfer coefficient	51.02	60.26	W/m²K	18.1%
total heat transfer in convective pass	705.8	741.5	MW	5.1%

Furnace Modeling

CFD-simulations of both the air firing and oxyfuel modes were made, which allow for the comparison of the different conditions in the furnace. The furnace geometry simulated is shown in Figure 22. In the vertical section of the temperature field in Figure 23a and 23c the positions of the individual burners can be clearly recognized in both operating modes by their relatively cold inlet flows. At the same time the high temperature gradients clearly reveal the burner ignition zones. As the chosen conditions lead to similar flame shapes, the temperature distributions in both operating modes are very similar especially at the burner levels. The lower mass flow through the OFA system in the oxyfuel mode leads to a slight tilt in the temperature field above the burner levels. At the furnace exit the temperature distribution is homogenized by the pressure loss induced by the convective section above the furnace. The furnace exit gas temperature is similar to that in the air combustion case.

The flue gas flow velocities in the furnace are shown in Figure 23b and 23d. The positions of the burners can be clearly identified from the high velocities there with nearly equal values both in oxyfuel and air-fired mode. It can also be seen that in Oxyfuel mode the low mass flow through the OFA system has no significant impact on the flow in the furnace. Therefore due to the staggered burner levels the hot flame gases are directed somewhat more to the front wall than in the air-firing mode. This effect could be overcome by a lower mass flow through the top burner level than through the level below. In air firing mode the highest velocities, which lie beyond the scaling, exist at the OFA system. The high mass flow through the OFA system counteracts the slight tilt induced by the staggered burner levels.

In general the CFD-Simulations show similar conditions in both the air firing and oxyfuel modes, thus the stable operation of the furnace in oxyfuel mode can be expected.

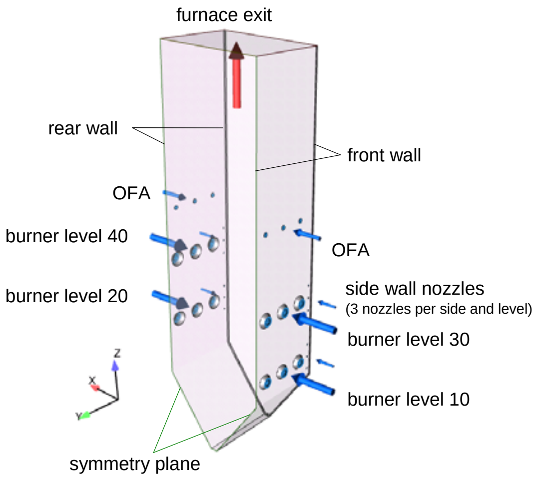
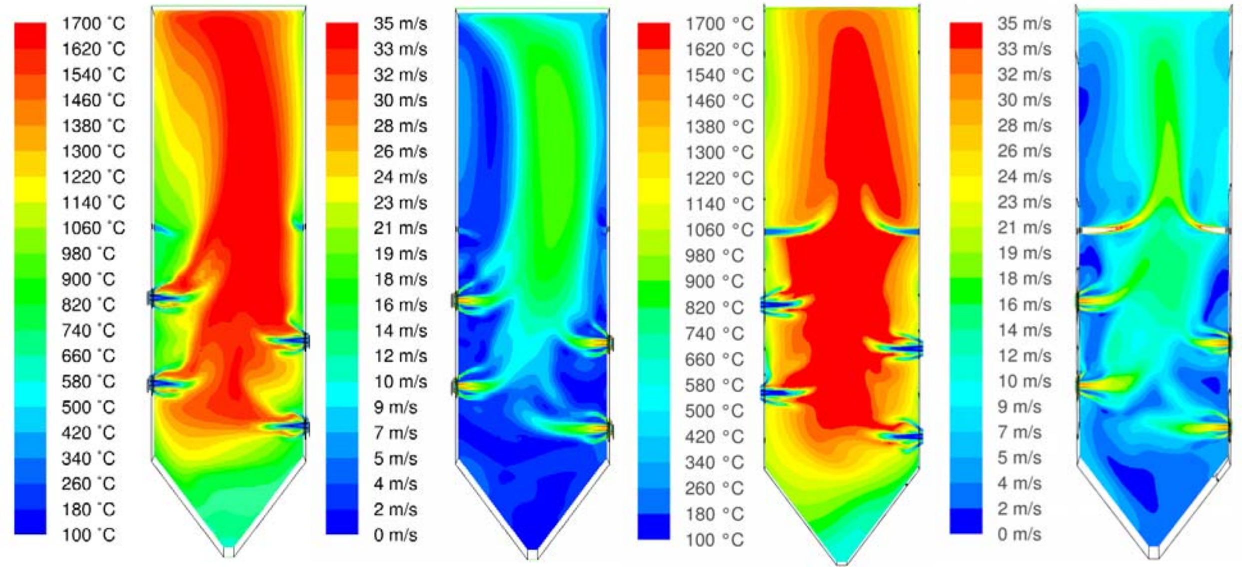


Figure 22: Furnace geometry



a) Temperature distribution - Oxyfuel mode

b) Velocity magnitude - Oxyfuel mode

c) Temperature distribution - air-fired mode

d) Velocity magnitude - air-fired mode

Figure 23: Simulated temperatures and velocity magnitudes in Oxyfuel and air-fired mode

Impact on Power Plant System

The flue gas composition for normal air firing and oxyfuel firing is shown in Table 2. The concentration of CO_2 in the flue gas after cooling is nearly 95 wt% (dry) in the case of oxyfuel firing. This gas can be directly compressed and directed to the storage site without further purification.

Table 2: Comparison of Flue Gas Composition

Gas species	Air firing (Composition after ESP) wt % wet	wt % dry	Oxyfuel-firing (Composition after cooling) wt % wet	wt % dry
H_20	5.9	-	4.8	-
CO_2	19.4	20.6	89.9	94.9
N_2	68.7	73.1	2.0	2.1
O_2	4.5	4.7	2.9	3.0

Table 3: Power Demand, Cooling Duty and Area Required for The Retrofit

	Area Required [m²]	Power Demand [MW]	Cooling Duty [MW]
ASU	26000	107	117
Cooling & AQCS modifications	900	4	150
CO_2 Compression	2000	85	

The required area for the new components as well as the demand for electrical energy and cooling capacity is shown in Table 3. The required area for AQCS modification is rather small compared to the ASU. Nevertheless, the ASU can also be constructed at some distance from the power station when the oxygen is transported to the boiler house by pipeline. Hence, there is normally enough space for the remaining modifications on the site. The arrangement of the new components and plant modifications is shown in Figure 24.

Table 3 also provides the energy requirements for a number of modifications. The ASU and compressors require a large amount of electrical energy. As a result, the gross electrical output of the power plant in the worst case is reduced by more than 24%.

It should be pointed out again that this study is focused on retrofit and with the intention to minimize the equipment modifications and to ensure the power plant after the retrofit can still be operated in oxyfuel or air fired modes. These measures are necessary for early retrofit applications of oxyfuel-based CCS to maintain power grid reliability. Future power plants designed solely for oxyfuel combustion can significantly reduce the penalties in plant efficiency and output by using improved furnace / boiler design and flue gas recirculation arrangement targeting oxyfuel as the only firing mode.

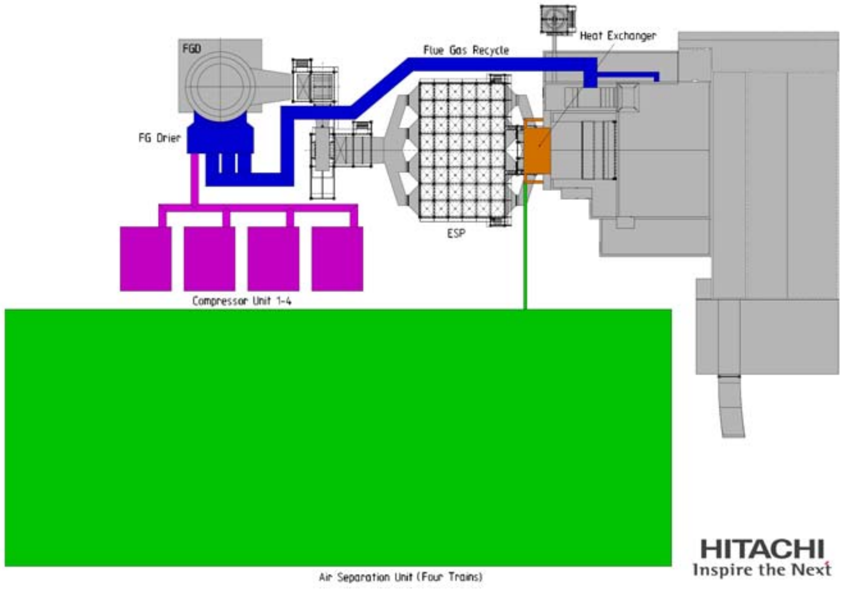


Figure 24: Arrangement for retrofitted power station

COMBUSTION TEST STUDIES

Combustion test studies have been carried out in bench scale and pilot scale to support the oxyfuel retrofit design development, in some cases in cooperation with partners and customers of Hitachi. Combustion experiments are essential to provide data for burning behavior and ash characteristics under oxyfuel conditions. These tests also help verify the numerical simulation tools used extensively in the design study, and generate data for integrated control and operation of the combustion system, air separation unit, and the air quality control system.

A number of small and large pilot plants have been used for this development program, including:

- 0.5 MW_{th} combustion test facility at IVD in University of Stuttgart, Germany
- 1 MW_{th} E.ON combustion test facility in Ratcliff, UK
- 1 MW_{th} combined combustion – air quality control system test facility in Akitsu, Japan
- 4 MW_{th} combustion pilot plant in Kure, Japan

Table 4 gives an example of the pilot test data. The data are from the 1 MW_{th} E.ON pilot plant firing a bituminous coal from South Africa and the 0.5 MW_{th} IVD test facility firing a German lignite. The CO_2 concentrations are approximately 94 vol.% dry from the 0.5 MW_{th} test facility, and only about 80% from the 1 MW_{th} plant due to its higher air in-leakage.

Table 4: Data from 1 MW_{th} and 0.5 MW_{th} Combustion Tests

Test Facility		**1 MW_{th}**	**1 MW_{th}**	**0.5 MW_{th}**	**0.5 MW_{th}**
Test Coal		Bituminous	Bituminous	Lignite	Lignite
Test Mode		Air	Oxy	Air	Oxy
O_2	%vol, dry	2.0	3.3	2.7	5.4
CO_2	%vol, dry	16.7	79.9	17.1	93.7
H_2O	%vol.	7.0	17.2	8.0	31.0
CO	ppm, dry	44	124	7	72
	lb/MMBtu	0.044	0.082	0.007	0.020
NOx	ppm, dry	245	329	451	499
	lb/MMBtu	0.34	0.09	0.67	0.23
SO_2	ppm, dry	513	1738	590	1366
	lb/MMBtu	0.96	0.62	1.18	0.88

NOx and SO_2 concentrations are higher because in both cases there are no post-combustion DeNOx or DeSOx and therefore NOx and SO_2 concentrations are increased due to the accumulation effect of flue gas recirculation. The NOx emission rates are significantly lower for oxyfuel tests, partly because the absence of air nitrogen under oxyfuel combustion conditions. The SO_2 emission rates for oxyfuel are also clearly lower than air firing. It is likely that the much elevated SO_2 concentration during oxyfuel firing leads to enhanced retention of SO_2 by the flyash which contains significant amounts of calcium oxide and magnesium oxide, based on the analyses of both coals. CO is higher for oxyfuel firing both in terms of concentrations in flue gas and the emissions rates.

This increased formation of CO could be partially due to the much higher CO_2 concentration under oxyfuel firing condition which intensifies the gasification reaction between CO_2 and coal char, with CO as a product.

The current focus of the pilot program is to prove and optimize flame temperature, radiation, burnout, and emission behavior in large pilot scale. Working with Vattenfall, Hitachi has installed a dual-firing (air /oxy) capable DS burner (a low NOx, staged combustion swirl burner) at Vattenfall's Schwarze Pumpe Oxyfuel Pilot Plant in Germany. The combustion test facility is equipped with integrated ASU, AQCS, CPU and produces liquid CO_2 for truck transportation to a storage site. The test burner, with a heat input of 30 MW, is nearly the size of those applied in Hitachi's commercial boilers. The combustion test with the DS burner at Schwarze Pumpe has started in early 2010.

OXYFUEL COMBUSTION SUMMARY

In summary, existing state-of-the-art coal-fired power stations can be converted to oxyfuel combustion with no change to the plant water-steam cycle and minimal modifications to the boiler island. Limited alterations to the air quality control system are needed. Major equipment additions for the air separation and CO_2 compression and handling are necessary. The converted power plant will have the flexibility to operate in both air-fired and oxyfuel modes.

While it has been shown that retrofitting existing power plants is technically feasible, all processes have to be further optimized in future to reduce the cost and efficiency penalty of CCS. Hitachi is currently undertaking extensive development work to improve the basic processes for oxyfuel combustion to accelerate the commercialization of this technology for CCS applications.

4. CONCLUDING REMARKS

Coal contributes to about forty percent of the electricity generation in the world today and is expected to maintain its dominant share in the foreseeable future. Clean coal power technologies will play an important role in a sustainable global energy structure in the coming decades. CO_2 capture and sequestration (CCS) based on post-combustion scrubbing or oxyfuel combustion is expected to be an essential part of an overall technology strategy to achieve the carbon reductions required for stabilizing atmospheric CO_2 concentration and global warming.

Hitachi is developing a full portfolio of new clean coal technologies aimed at further efficiency improvement, 90% CO_2 reduction, and near-zero emissions of other pollutants. The next generation ultrasupercritical power plants with oxyfuel combustion or post-combustion CO_2 capture will enable carbon storage applications within a decade. With these technologies, coal will continue to contribute to the well-being of human society.