

# A Thermodynamic Study of a Novel Zero-Emission Power Plant, Based on Hydrogasification of Coal

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## Abstract

This paper analyzes various aspects of a zero-emission power plant, based on a coal gasification process using recycled steam and hydrogen as gasifying agents, with a calcium chemical looping for CO<sub>2</sub> capture. The synthetic fuel produced, a mixture of steam and hydrogen, is burned with oxygen in a semi-closed high temperature steam cycle. The paper presents a rather comprehensive analysis of various critical issues (e.g. the calciner process and the cooling of the high temperature turbine), often neglected in previous works from the literature. A realistic performance estimation was developed, with reference to advanced components, whose production can be foreseen for a mid-term future with known technologies, if not present in today's marketplace. The predicted net efficiency is 50% with a maximum cycle temperature of 1500°C: it is an unsurpassed value for a zero-emission coal power plant, justifying further relevant R&D efforts.

## 1. Introduction

In the last few years, the carbon capture and storage (CCS) technologies, applied to coal-fed power plants, are encountering an increasing popularity. Demonstration plants are now under final planning and development. Being the first-of-a-kind, low-risk capture technologies are adopted, namely: (i) pre-combustion capture, including syngas production from coal or natural gas, water-shift reaction, CO<sub>2</sub> absorption, hydrogen-rich syngas combustion in a gas turbine, (ii) amine capture from flue gases of conventional power plants. A third technology (oxy-combustion) is also considered in many studies, particularly applied to pulverized coal boilers. These processes, even if introducing new milestones in the power industry, are characterized by a relevant loss of efficiency with respect to the reference power plants without capture (e.g. 35-38% vs. 43-45% for coal power plants), yielding to elevated costs of the CO<sub>2</sub> sequestered (30 to 50 €/ton for coal, much more for natural gas).

In a scenario projected to mid-long term applications, a number of alternative processes can be proposed, aimed to improve the performance (efficiency, emissions, economics) of coal power plants with CCS [1]. The process here proposed is based on the use of steam as the oxidizing agent for coal (hydrogasification – HG) and on the chemical looping of calcium oxide – calcium carbonate for CO<sub>2</sub> capture. It is derived from the basic idea of Johnson and Ziock [2], but it was deeply modified, especially as far as the power cycle is concerned, to make it feasible with advanced but realistic components. In the Zeca [2] process, the fuel produced by the HG section (a mixture of steam and hydrogen, deprived of carbon compounds) feeds a high temperature solid-oxides fuel cell (SOFC), whose exhausts are the heat carrier for the calciner. This is a very brilliant concept, potentially achieving unsurpassed efficiency, but it is not likely to work in the real world for two main reasons: (i) the utilization of SOFCs in multi-hundreds MW power plants, in the perspective of reducing their cost to be competi-

tive in such a market, is very far from any predictable development, (ii) calcination at high pressure requires heat at very high temperatures (well over 1250°C at the pressure quoted by [2]), a figure out of any capability of SOFCs. In the scheme previously studied by [3] and here analyzed, the fuel is burned by pure oxygen within a semi-closed, high temperature steam cycle, while coal oxy-combustion provides the necessary heat to calcination.

These two solutions should ensure the technical feasibility of the thermodynamic process with potentially sustainable economics. However, we will outline the necessity of significant advances, with respect to today's best technology, to make the system viable for industrial application while conserving superior performance. Therefore, the approach used in this paper is to make reference to technological advancement, reasonable foreseeable for the next 10-20 years (e.g. blade cooling systems resembling the transpiration concept), including components not available in today's power industry, provided that they can be developed by resorting to known technologies (e.g. oxygen-hydrogen burners, elevated temperature steam compressors), but excluding unproven devices whose feasibility cannot be anticipated today (e.g. SOFC operating at 1300+°C, uncooled turbines at 1500+°C...).

As we will discuss, the proposed system offers very attractive advantages (zero emissions, elevated efficiency), but their evaluation is not straightforward, because the performance of novel processes and of novel components cannot rely on manufacturers' or users' experience. The present paper, following the work of Calabrò et al. [3], presents a thermodynamic study of the process. It includes a general assessment of the chemical section and a rather detailed analysis of the power plant, based on (i) appropriate models for predicting the performance of some key components, (ii) an evaluation of the effects of cooling requirements of the hot parts and (iii) a careful optimization of the heat recovery sections.

## 2. General plant description

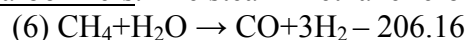
Let us briefly describe the concept behind the process. Two main sections can be depicted (fig.1): the chemical island, producing a clean, hydrogen-rich syngas and a CO<sub>2</sub> stream ready for sequestration, and the power island. A third section is the air separation unit (ASU): even if very relevant in terms of power consumption and investment cost, it consists of a conventional double-column process, including air and oxygen compression (nitrogen is wasted), and does not require a particular description. A fourth section is devoted to CO<sub>2</sub> compression and liquefaction. The chemical island is made of three processes:

- The hydrogasification (HG). It is a coal-slurry fed gasifier basically using steam as the oxidizer. It operates at moderate temperatures (800°C), whereas both endothermic and exothermic reactions occur, such as:
 

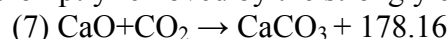
(1) $C+H_2 \rightarrow CH_4 +74.9$	(3) $C+H_2O \rightarrow CO+H_2 -131.28$
(2) $C+CO_2 \rightarrow 2CO -172.44$	(5) $CO+H_2O \rightarrow CO_2+H_2 +41.2$
(4) $2C+O_2 \rightarrow 2CO +221.1$	

(heat of reaction in MJ/kmol, H<sub>2</sub>O as steam, at 15°C, 1.013 bar). Some oxygen can be used to rise the reactants temperature to improve the kinetics. Large steam excess is needed to complete the carbon conversion, therefore the oxidizer is the recirculated gas from the first decarbonizer (see below). The product gas is mainly composed of CH<sub>4</sub>, CO, CO<sub>2</sub>, H<sub>2</sub> and steam.

- The decarbonizers. The steam-methane reforming is here accomplished:



The CO produced is converted to CO<sub>2</sub> by the water-gas shift (reaction 5), provided that CO<sub>2</sub> is promptly removed by the strongly exothermic reaction:



Therefore, reactions (5) and (6) are pushed rightwards: carbon is removed from the gaseous species and adsorbed by calcium. To minimize the steam needs it is convenient

to use two decarbonizers (Dec1 and Dec2): a large amount of the gas treated by Dec1 is recirculated to HG, not needing a complete carbon removal, while the smaller gas flow in Dec2 is deeply decarbonized by large steam and calcium excess.

- The calciner (Cal). It regenerates calcium oxide by reversing reaction (7) thus requiring large heat supply at an elevated temperature, related to  $\text{CO}_2$  partial pressure as shown by fig.2. Given the pressure range of 20-30 bar here used, and excluding calciner operation at a pressure much lower than the one in the decarbonizers<sup>1</sup>, temperatures in the range of 1200°C can be anticipated.

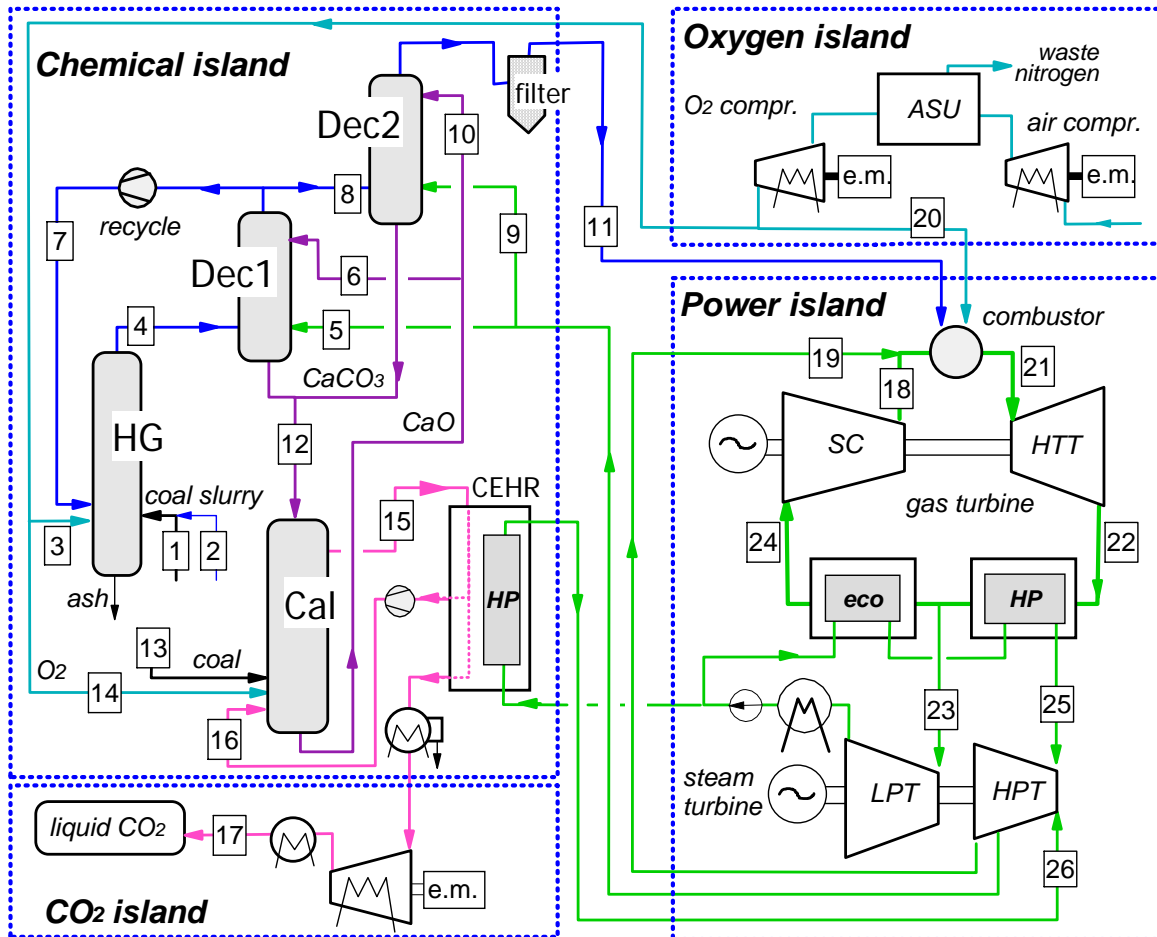


Fig.1: Simplified scheme of the complete process.

The only viable solution to provide heat at such temperatures is direct combustion. Oxy-combustion of coal is the simplest solution, to obtain a final stream of  $\text{CO}_2$  and  $\text{H}_2\text{O}$ , ready for sequestration after water condensation. The presence of coal ash and of sulfur compounds (i.e. CaS), polluting the product CaO, should represent a workable problem because the circulating solid charge has to be replaced progressively, to avoid degradation of its sorption-desorption capabilities (see later).

The plant scheme of fig.1 is not significantly different from the one discussed in [3]. However we introduced a more detailed analysis of some crucial components, yielding to different operating parameters while keeping the same zero-emissions concept. As well, we re-

<sup>1</sup> Calcination at atmospheric pressure would be preferable due to the lower operating temperature. However, a system introducing solid charges in pressurized vessels starting from atmospheric conditions (for instance, lock hoppers pressurized by steam) seems quite unrealistic due the very large quantities of solids here involved and due to the losses of mass and pressure, relative to the very large amount of steam needed for pressurization.

moved some relevant simplifying assumptions to obtain a more reliable efficiency prediction and technological assessment, as described in the next chapter.

For the power cycle, the high temperature syngas produced (mainly composed by hydrogen and steam) is mixed with IP steam available at the IP turbine outlet, to reduce the temperature of the fuel-rich mixture. However, much more dilution steam is needed to limit the combustor outlet temperature to acceptable values for the High Temperature Turbine (HTT), i.e. to values similar to the ones used in modern industrial gas turbines. Such dilution steam comes from recycling part of the steam at near-atmospheric pressure, available after the HTT expansion and heat recovery. A steam compressor is needed for this task: therefore the power cycle is basically a semi-closed Joule cycle. The excess steam, not needed for dilution, is expanded in a LP condensing steam turbine. Condensate is pumped to supercritical pressure, superheated and re-heated by recovering heat from the HTT discharge. It is interesting to notice that the reheat steam turbine does not have an IP section (from 30 to 1 bar) because: (i) a large part of 30 bar steam is needed by the hydrogasification process, (ii) the remainder is used as dilution steam in the combustor and then expanded in the HTT.

### **3. Method of calculation and solutions used**

Calculations were completely performed by means of the GS computer code developed by the authors' working group at Politecnico di Milano, with the exception of the CO<sub>2</sub> compression and liquefaction calculated by commercial software Aspen<sup>TM</sup>. GS is devoted to calculate the energy/exergy balance of virtually any type of power plant configuration, built-up by assembling a number of components, like turbines, compressors, heat exchangers, chemical reactors, and so on. The key features of GS are: (i) the possibility of reproducing power plants with extended chemical sections (like IGCCs, plants with CO<sub>2</sub> capture, etc.) in a single tool, mainly by means of chemical equilibrium calculations; (ii) the implementation of some models for an appropriate evaluation of the performance of some key components: in particular the prediction of the efficiency of steam turbines, cooled gas turbines, axial compressors and of the cooling needs of the high temperature turbines can be performed by proper correlations, often following a 1-D first design of the turbomachines, accurately calibrated to reproduce the performance of commercial engines and power plants. These models are described in [4-6] and used in a number of previous papers (e.g. [7-8]). The limit of GS is that it can just handle: (i) mixtures of ideal gases, (ii) incompressible flows (liquids and solids), (iii) water/steam as the only real gas. This is relevant for the present paper: see section 3.3.

When analyzing the present plant scheme, the models used to predict the behavior of a number of components deserve some descriptions, because they strongly affect the overall performance and the technical feasibility of the system.

#### **3.1 Hydrogasifier and decarbonizers**

The HG is calculated by assuming chemical equilibrium conditions. A moderate oxygen injection (point 3 in fig.1) was used to keep a sufficiently high operating temperature, here assumed as 800°C, to obtain acceptable kinetic aspects. The rate of recirculation (oxidant flow / HG total flow) was determined to assure a proper steam excess for hydrogasification. The recycle compressor is a very particular and delicate machine, operating at extreme temperatures (about 800°C): its technical feasibility is a key point for the realization of the plant.

Also decarbonizers are calculated at chemical equilibrium. The amount of steam used was determined in order to obtain a reduced content of carbonated compounds in the final syngas (this affects the power cycle thermodynamics: high steam consumptions are detrimental to efficiency). Sulfur and fly ash removal is not addressed in this paper, assuming that: (i) hot-gas clean-up systems will be available in a next future (for instance, metal oxides adsorption, ceramic filters...), with negligible impact on the system heat balance, and/or (ii) sulfur

capture can be performed by contacting acid gas with calcium oxide in the carbonizers. This is a very simplifying assumption, deserving more attention in the future. A low sulfur coal is here utilized to make the results more realistic.

### 3.2 Calciner

The calciner operates in pressurized conditions, at about 30 bar like the remainder of the chemical section, to make the solid circulation feasible with normal fluidized or fixed bed techniques (the calciner is supposed to be a fixed bed type reactor, while the circulating bed option can be studied for the decarbonizers: their basic design is a very challenging task, to be addressed in the future). Pressurized operation call for higher operating temperatures (fig.2: 1170°C for the actual CO<sub>2</sub> partial pressure), requiring a larger coal and oxygen consumption than for atmospheric operation. However, the CaO at very high temperature brings about a relevant sensible heat

contribution to the syngas, thus enhancing the temperature of the final syngas. In other words, the excess heat required by the calciner is transferred to the power cycle by means of a warmer syngas: assuming as technically feasible the handling of syngas at about 800°C, the process is very efficient and no losses are caused by a larger calcination temperature.

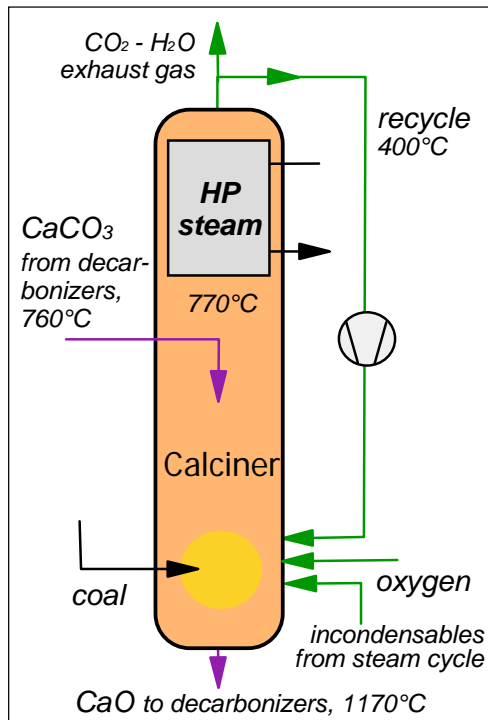


Fig.3: Conceptual scheme of the calciner arrangement.

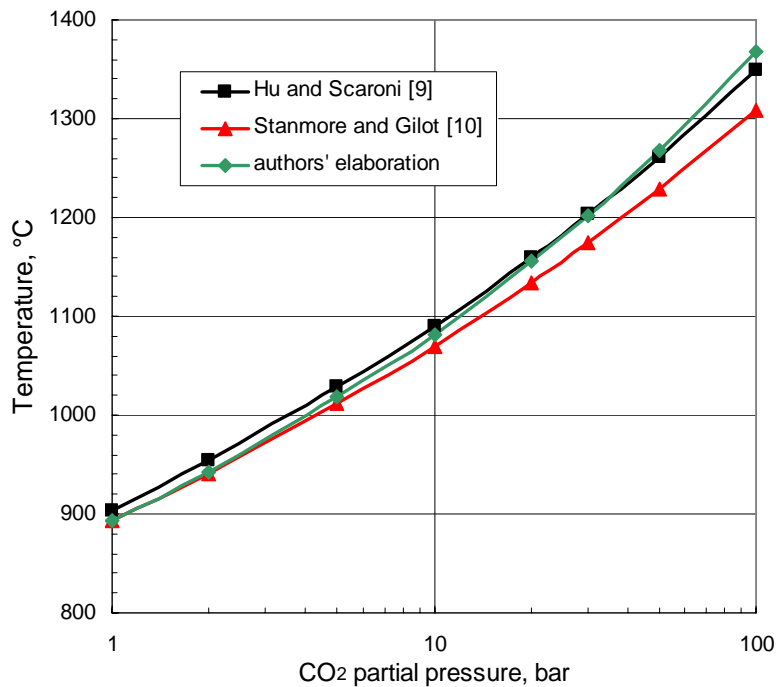


Fig.2: Calcium carbonate equilibrium decomposition temperature, as a function of CO<sub>2</sub> pressure. In this paper we will refer to the Stanmore and Gilot data [10].

Another feature, useful to reduce the coal-oxygen requirement, is that CaCO<sub>3</sub> is preheated, before entering the reacting zone, by contacting the combustion gas (see fig.3 for a qualitative schematization). The system is similar to one used in single tower ovens for CaO industrial production, where limestone is introduced at the tower top and its temperature increases by cooling the product gas, which flows counter-current. Hence, combustion heat is used primarily to provide heat of reaction at high temperature, rather than to preheat limestone, as required by good thermodynamics: as well, the CO<sub>2</sub> rich gas exits the calciner at moderate temperature, more appropriate for steam generation.

To obtain more realistic results, it was considered that 5% of the circulating solid needs to be replaced by fresh limestone, to prevent a dramatic loss

of sorption capability. The real rate of replacement is one of the big ‘question marks’ in the practical feasibility of the system: even if a number of literature indications are now available [11-12], no experience exists in large scale systems operating in an industrial environment with coal combustion products. The figure of 5% here assumed is just a tentative to make our calculations more realistic. Heat is recovered from the spent limestone by rising high pressure steam, so that the performance penalization is pretty reduced.

### 3.3 Power cycle working fluid

The syngas, consisting of hydrogen and steam, with about 2-4% of CH<sub>4</sub>, CO and CO<sub>2</sub>, is burned by 98% purity oxygen and diluted by re-circulated steam, thus producing a mixture of steam (97-98%), CO<sub>2</sub>, N<sub>2</sub>, Ar and O<sub>2</sub> as the working fluid for the Joule power cycle<sup>2</sup>. In such a mixture, operating between 1–30 bar and 100–1600°C, some real gas effect takes place, while our computer tool considers steam as an ideal gas when mixed to other gases.

By some preliminary calculation we estimated that, considering steam as an ideal gas instead of a real gas, in the cycle operating conditions very limited differences take place, less than 1% on the cycle heat balance, uniquely due to some discrepancy relative to the compression phase, operated at lower temperatures with stronger real gas effects. Such inaccuracy is tolerable in the context of the present analysis. In fact, steam is considered pure (and therefore treated as a real gas) in the high pressure part of the cycle, between the supercritical pressure of 250 bar and the process pressure of 30 bar. The here assumed HP steam conditions of 250 bar, 600/610°C, with condensation at 4 kPa, are typical of recent steam power plants.

### 3.4 Utilization of incondensable species

The incondensable species CO<sub>2</sub>, N<sub>2</sub>, Ar, O<sub>2</sub>, entrained from the expanding flow to condenser, are removed from the condenser itself or from the deaerator, together with some steam. Rather than dissipating these gases to the atmosphere, it is useful to use them as oxidizer for the calciner (fig.3), after a proper compression (incondensable recovery and compression is a usual practice in geothermal power plants, using geothermal steam with relevant CO<sub>2</sub> content). In such a way, excess oxygen is fully recirculated in the system and CO<sub>2</sub> capture is complete, even if some carbonated compounds are present in the syngas. This improves the plant efficiency and virtually no carbon escapes the system.

### 3.5 HTT blade cooling

The cooling of the HTT blades is a crucial point in the analysis, not only because cooling affects the cycle thermodynamics and the plant performance at a given cycle maximum temperature, but especially because the selection of such temperature is completely arbitrary if cooling is not addressed, making comparisons meaningless.

For the cycles here selected, the HTT operates at gas conditions not far from the more advanced gas turbines now present in the industrial market (turbine inlet temperature of 1400-1500°C). However, there are significant differences between combustion gases and steam, as far as the heat transfer properties are concerned. In fact, the higher thermal conductivity and specific heat of steam [7], together with the higher inlet pressure (30 bar vs. about 20 of recent industrial gas turbines), bring about a higher heat transfer coefficient and, therefore, at similar temperatures, a higher heat flux (thermal power exchanged per unit surface of cooled blade). One relevant consequence is that the temperature drop due to heat conduction across the blade thickness becomes a critical issue. Considering that a typical heat transfer coefficient in the first stator row is 2500 W/m<sup>2</sup>K for a modern conventional gas turbine (according

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<sup>2</sup> The content of N<sub>2</sub>, Ar and O<sub>2</sub> are determined by the assumptions of: (i) 98% oxygen purity (a value which can be obtained by means of conventional double column air separation units without relevant penalties on the process energy requirement), (ii) an oxygen excess of 1% with respect to stoichiometric combustion.

to our calculation) and that it becomes about  $5600 \text{ W/m}^2\text{K}$  for the present application, the conduction  $\Delta T$  more than doubles (at equal blade thickness and material). The situation is depicted in tab.1, comparing the simplified cooling situation in the first nozzle of a modern gas turbine (a FB unit with  $TIT=1350^\circ\text{C}$  – case A) to the situation that would occur for the present cycle with the same cooling technology, but taking advantage of a much lower coolant temperature (case B)<sup>3</sup>. The latter presents very elevated cooling requirements and strongly affects the efficiency. A more manageable situation can be accomplished (case C) if we assume that the blade maximum temperature can be risen from  $845$  to  $895^\circ\text{C}$  (a realistic position for a near-future technology) and that the blade conductivity can be halved, by resorting to a lower thickness (very difficult to obtain in the practice!) or by introducing a kind of transpiration cooling (representing a possible development for advanced machines of a next generation, a position quite coherent to the general context of the present analysis). By this assumptions, the cooling situation is quite favorable and the achievement of a  $TIT$  as high as  $1500^\circ\text{C}$  seems to lead to a feasible situation (case D).

*Tab.1: Calculated cooling situation for the first nozzle of an advanced commercial gas turbine (case A) and of three different cases for a semi-closed steam cycle. The calculation method and the assumptions are described in [6].*

Case	A	B	C	D
Expanding fluid, coolant	Gas, air	Steam, Steam (98% approx.purity)		
Turbine inlet temperature, °C	1350	1350	1350	1500
Gas side blade temperature, °C	845	845	895	895
$\Delta T$ across the blade wall, °C	165	275	134	191
Coolant inlet temperature, °C	405	250	250	250
Coolant outlet temperature, °C	588	405	495	480
Coolant flow, % of inlet gas flow	9%	11.6%	6.8%	9.2%

*Tab.2: Details of the cooled high temperature turbine relative to the case with  $TIT=1500^\circ\text{C}$ .*

Stage number		1	2	3	4	5	6	7
Rotor mean diameter, m		2.475	2.573	2.671	2.769	2.867	2.965	3.063
Rotor exit blade height, m		0.167	0.224	0.319	0.438	0.573	0.706	0.814
Degree of reaction		0.050	0.125	0.200	0.275	0.350	0.425	0.500
Isentropic enthalpy drop, kJ/kg		332.4	321.7	306.1	285.4	259.3	227.3	189.3
Specific speed		0.297	0.371	0.492	0.655	0.888	1.235	1.824
Efficiency		0.883	0.894	0.908	0.920	0.927	0.934	0.852
Coolant inlet temperature, °C	S	250.0	250.0	511.6	411.7	330.7	0	0
	R	250.0	250.0	471.9	368.7	277.3	0	0
Coolant outlet temperature, °C	S	476.1	482.0	652.1	698.4	787.5	0	0
	R	537.8	519.1	654.9	686.1	756.6	0	0
$\Delta T$ across blade wall, °C	S	188.1	102.4	68.1	24.4	3.4	0	0
	R	99.7	60.7	37.5	14.2	1.5	0	0
$\Delta T$ across thermal barrier coatings, °C	S	227.1	123.7	0	0	0	0	0
	R	125.5	76.4	0	0	0	0	0
Average heat transfer coeff., $\text{kW/m}^2\text{-K}$	S	5.749	4.489	3.183	2.185	1.460	0	0
	R	3.619	2.902	2.189	1.629	1.181	0	0
Coolant mass flow, kg/s	S	38.19	27.87	35.35	11.97	4.68	0	0
	R	18.90	17.15	20.39	9.14	4.11	0	0
Rotor inlet mass flow, kg/s		429.1	475.9	528.4	560.7	574.5	578.6	578.6

<sup>3</sup> The coolant steam for the high pressure parts of the HTT, taken from the steam compressor discharge, cannot be used directly, because its temperature is too elevated ( $612^\circ\text{C}$ ). Therefore, it is mandatory to cool it as much as possible ( $250^\circ\text{C}$ , close to the saturation temperature of  $234^\circ\text{C}$  at 30 bar).

The results of tab.1 are just relative to the first blade row. However, the considered industrial gas turbine has 6 cooled rows (out of 4 stages), while the HTT has 10 cooled rows (out of 7 stages, with comparable aerodynamic load) and therefore larger coolant flows are required in the present case. The cooling calculation method [6] includes a 1D design of the turbine, reported in tab.2 together with some relevant information about cooling. Notice that the first four rows are cooled by means of the high pressure steam flow taken from the compressor outlet and cooled to 250°C. The following six rows are cooled by steam taken at intermediate compressor ports at higher temperatures. Extensive use is made of thermal barrier coatings in the first two stages. For those rows, the  $\Delta T$  across the blade wall is very large in spite of the favorable assumptions: this reduces the allowable coolant  $\Delta T$ , enhancing the cooling requirements. The total cooling flow is one third of turbine outlet flow rate, a very elevated figure, depicting the cruciality of the cooling problem.

### 3.6 Heat recovery system

An efficient heat recovery from the HTT exhausts is essential for the cycle efficiency. From fig.4 showing a better detail of the power section than fig.1, we can see that:

- Supercritical steam is evaporated, superheated and reheated (250 bar, 600/610°C) in the higher temperature heat recovery section;
- Steam exceeding the mass flow handled by the compressor (determined by the dilution needed in the combustor) is diverted to the condensing section of the steam cycle (point 23), by splitting the atmospheric steam flow before it enters the low temperature (economizers) section of the heat recovery system;
- Steam to be compressed must be cooled as much as possible, e.g. 110°C; this is mainly done by HP condensate pre-heating, but, if it is not sufficient, some IP steam (30 bar) can be produced and then used for cooling, mixed to the coolant extracted from SC outlet;
- Steam taken from the compressor outlet to cool the HTT is cooled from 612 to 250°C by rising HP steam from condensate at intermediate temperature (about 200°C);
- Therefore, supercritical steam entering the HP steam turbine comes from four heat sources: (i) the HTT exhausts, (ii) cooling of the HTT coolant, (iii) the CO<sub>2</sub>-H<sub>2</sub>O stream from the calciner, (iv) the spent limestone. Steam at point 25 in fig.4 is generated from sources (i) and (ii), point 26 from sources (iii) and (iv).

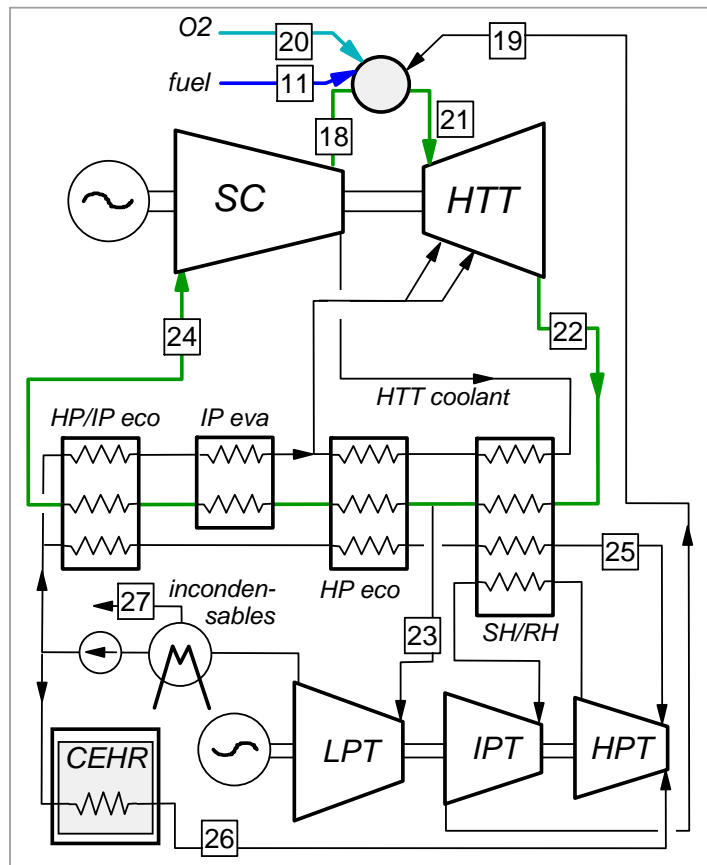


Fig.4: Detailed arrangement of the power section.  
CEHR: calciner exhausts heat recovery.

## 4. Results

A large number of cycle parameters should be optimized to obtain the best performance from a complicated cycle configuration as the one here considered. Let us recall some of

the most important: (i) the pressure of the chemical section, (ii) the rate of recycle to the HG, (iii) the steam flow to Dec1, (iv) the steam flow to Dec2, (v) the maximum cycle temperature (we call it TIT, with same meaning – first rotor total inlet temperature – used for gas turbine cycles), (vi) the HTT outlet pressure, (vii) the HP steam pressure and temperatures, (viii) the temperature at which part of the LP steam is diverted to the condensing section, and many more. We did not perform a full optimization, but we made a reasonable selection of the mentioned parameters to obtain two final solutions (with two different TIT) that, in our experience, should be pretty close to the optimized ones. In particular: (i) 30 bar was assumed as the system pressure, a proper value for both chemical and power sections: however it deserves further analyses; (ii) recycle rate and steam rates were selected to assure sufficient steam excess for both gasification and reforming, thus obtaining low concentration of carbonated species in the final syngas; (iii) two TIT of 1400 and 1500°C were considered, following the discussion of 3.5, (iv) a near-atmospheric pressure was set at HTT outlet. An exhaustive analysis of the influence of the above factors will be included in [13].

Tab.3 reports the power breakdown and the net efficiency of the two cycles, including a first estimate of the auxiliary power requirement (coal handling was estimated at 1% of coal LHV input – see also [8] for other assumptions) and of the CO<sub>2</sub> compression and liquefaction. The most interesting figure is the net efficiency, close to 50%. This is an outstanding result, because today's best reference power plants, either USC or IGCC, can provide at best 45-46% efficiency without CO<sub>2</sub> capture, while the application of capture technologies will abate efficiency in the 33-38% range, according to literature [14-16].

*Tab.3: Power breakdown and efficiency of the proposed plant, with two TIT.*

TIT (°C)	1400	1500
Power (electric-mechanical), MW		
High Temperature Turbine	887.0	844.2
Steam Compressor	-318.7	-267.2
HTT+SC electric-mechanical loss	-9.4	-9.5
Steam Turbine	217.9	224.6
Power cycle auxiliaries	-14.3	-14.5
Gasification recycle compressor	-11.5	-11.5
Coal handling and auxiliaries	-12.5	-12.5
ASU air compressor	-83.4	-83.2
O <sub>2</sub> compressor	-36.8	-36.7
Incondensable gas compressor	-4.0	-4.2
CO <sub>2</sub> compression/liquefaction	-12.6	-12.6
Net power output (MW <sub>EL</sub> )	601.8	617.0
Coal input LHV (MW <sub>TH</sub> )	1227.3	1225.1
Net LHV efficiency (%)	49.04	50.36

Tab.4 shows the main flow characteristics in the most relevant points (see figs.1 and 4), for the case at TIT=1500°C. From the two tables, it can be noticed that:

- About 30% of the coal input is used by the calciner, even if limestone enters at a very high temperature of 767°C;
- The final syngas is produced at 820°C: this is one of the main reasons for the superior thermodynamic performance of this process (the product gas remains at high temperature, without efficiency losses related to cooling), but cleaning, handling and controlling of this stream is quite questionable in a real plant;
- The final syngas contains about 3.5% of carbonated species, which are supposed to be extracted from the steam condenser (after being oxidized in the combustor) and recycled to the

calciner (point 27, after compression). Lower carbon content can be achieved by using larger steam excess, but this would lower the efficiency: even in the present situation, 102 kg/s of steam (points 5+9) are used in the decarbonizers, out of 176 kg/s entering the HP turbine (points 25+26);

- Oxygen production and compression requires 120 MWe out of 792 of gross power produced (about 15%): a careful optimization of the ASU is therefore of the highest relevance, as for any type of oxy-combustion power plant.

*Tab.4: Mass flow, temperature, pressure and chemical composition (% molar) in the most significant points of the proposed plant (from fig.1 and 4), with a TIT of 1500°C.*

N.	G (kg/s)	T (°C)	p (bar)	CH <sub>4</sub>	CO	CO <sub>2</sub>	H <sub>2</sub>	H <sub>2</sub> O	Ar	N <sub>2</sub>	O <sub>2</sub>	Ash	CaO	Ca- CO <sub>3</sub>
1	34.52	15.0	coal (%w: 64.4 C, 4 H, 7.4 O, 1.5 N, 0.8 S, 9.2 H <sub>2</sub> O, 12.7 ash – LHV=24.62 MJ/kg)											
2	11.50	90.0	35	0	0	0	0	100	0	0	0	0	0	0
3	3.64	223.9	35	0	0	0	0	0	1.46	0.54	98	0	0	0
4	129.37	800.0	33.24	16.94	8.26	5.29	43.23	25.66	0.05	0.57	0	0	0	0
5	52.07	510.5	33.24	0	0	0	0	100	0	0	0	0	0	0
6	88.13	1169.8	33.24	0	0	0	0	0	0	0	0	1.38	98.62	0
7	84.61	890.7	35	6.26	5.12	2.10	58.96	27.10	0.04	0.43	0	0	0	0
8	42.43	858.3	31.58	6.26	5.12	2.10	58.96	27.10	0.04	0.43	0	0	0	0
9	50.05	510.5	33.24	0	0	0	0	100	0	0	0	0	0	0
10	24.70	1169.8	33.24	0	0	0	0	0	0	0	0	1.38	98.62	0
11	77.25	820.5	30	1.03	1.23	1.20	49.00	47.26	0.02	0.26	0	0	0	0
12	182.26	767.3	30	0	0	0	0	0	0	0	0	0	18	82
13	15.24	15.0	coal (%w: 64.4 C, 4 H, 7.4 O, 1.5 N, 0.8 S, 9.2 H <sub>2</sub> O, 12.7 ash – LHV=24.62 MJ/kg)											
14	32.13	198.7	32	0	0	0	0	0	1.46	0.54	98	0	0	0
15	158.04	777.3	28.50	0	0	83.63	0	11.70	1.43	1.35	1.77	0	0	0
16	27.71	407.7	32	0	0	83.63	0	11.70	1.43	1.35	1.77	0	0	0
17	123.55	30.0	150	0	0	94.71	0	0	1.62	1.53	2.00	0	0	0
18	162.98	612.7	30	0	0	2.06	0	97.35	0.26	0.25	0.09	0	0	0
19	73.25	494.5	30	0	0	0	0	100	0	0	0	0	0	0
20	64.45	198.7	30	0	0	0	0	0	1.46	0.54	98.00	0	0	0
21	390.89	1605.2	29.4	0	0	2.06	0	97.35	0.26	0.25	0.09	0	0	0
22	612.60	581.3	1.04	0	0	0	0	100	0	0	0	0	0	0
23	227.90	350.0	1.01	0	0	2.06	0	97.35	0.26	0.25	0.09	0	0	0
24	353.39	110.0	1.00	0	0	2.06	0	97.35	0.26	0.25	0.09	0	0	0
25	136.99	590.0	247	0	0	0	0	100	0	0	0	0	0	0
26	40.41	640.0	247	0	0	0	0	100	0	0	0	0	0	0
27	13.52	118.9	30	0	0	77.59	0	0	9.65	9.23	3.54	0	0	0

## 5. Conclusions

The predicted overall net efficiency (including CO<sub>2</sub> compression and air separation consumptions) is in the range of 50%, an impressive value for a coal power station with carbon capture, even if lower than some over-optimistic early predictions [2]. The system can be proposed as ‘zero-emissions’ because: (i) 100% CO<sub>2</sub> is captured, by assuming that incondensables produced by the steam semi-closed cycle can be recycled to the calciner, (ii) the carbon content of the fuel is negligible, so carbon monoxide, unburned organic compound and particulates would be virtually absent, (iii) nitrogen oxides should also be absent, because negligible nitrogen quantities are present in the combustion chamber.

However, many formidable technical challenges must be faced to demonstrate the feasibility of the system, as well as many novel components must be developed. In the authors’

opinion, a crucial point is the clean-up and the handling/controlling of the syngas at very high temperature (never falling below 800°C!): ceramic candles filtering and metal-oxides sulfur removal are the key components, but they are not, at present, proven technology. Another key point is the capability of the limestone of withstanding chemical looping with a limited rate of fresh charge requirement. On the point of view of the components, we can outline that: (i) no experience exists in large industrial hydrogen oxy-burners with characteristics similar to the present ones, (ii) the turbine blade cooling requires some serious advancement to reduce the coolant needs, (iii) the raw syngas recycle compressor also operates in extreme conditions, (iv) the reactors system is fully unproven, (v) the handling of huge quantities of solids at high pressure and extreme temperatures is another tremendous problem.

Therefore, this study should be simply considered as an evaluation of the (large!) potential of the system, useful to outline the key areas for further R&D activities, devoted to a better understanding of the stimulating technical problems brought about by a future generation of zero emissions fossil fuel power plants.

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