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OPTIMIZATION OF THERMODYNAMICALLY EFFICIENT NOMINAL 40 MW ZERO EMISSION PILOT AND DEMONSTRATION POWER PLANT IN NORWAY

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ABSTRACT

In Aug 2004 the ***Zero Emission Norwegian Gas (ZENG)*** project team completed *Phase-1: Concept and Feasibility Study* for a 40 MW Pilot & Demonstration (P&D) Plant, that is proposed will be located at the Energy Park, Risavika, near Stavanger in South Norway during 2008.

The power plant cycle is based upon implementation of the natural gas (NG) and oxygen fueled Gas Generator (GG) (1500 ºF / 1500 psi) successfully demonstrated by *Clean Energy Systems (CES) Inc.* The GG operations was originally tested in Feb 2003 and has currently (July 2005) undergone extensive commissioning at the CES 5MW Kimberlina Test Plant, near Bakersfield, California.

The ZENG P&D Plant is an important next step in an accelerating path towards demonstrating large-scale (+200 MW) commercial implementation of zero-emission power plants before the end of this decade. However, development work also entails having a detailed commercial understanding of the techno-economic potential for such power plant cycles: specifically in an environment where the future penalty for carbon dioxide (CO2) emissions remains uncertain.

Work done in dialogue with suppliers during ZENG Project Phase-1 has cost-estimated all major plant components to a level commensurate with engineering pre-screening. The study has also identified several features of the proposed power plant that has enabled improvements in thermodynamic efficiency from ~38% up to ~45% without compromising the criteria of implementation using "near-term" available technology. The work has investigated:

(i) Integration between the cryogenic air separation unit (ASU) and the power plant.
(ii) Use of gas turbine technology for the intermediate pressure (IP) steam turbine.
(iii) Optimal use of turbo-expanders and heat-exchangers to mitigate the power consumption incurred for oxygen production.
(iv) Improved condenser design for more efficient CO2 separation and removal.
(v) Sensitivity of process design criteria to "small" variations in modeling of the physical properties for CO2 / steam working fluid near saturation.
(vi) Use of a second "conventional" pure steam Rankine bottoming cycle.

In future analysis, not all these improvements may necessarily be "cost-effective" when taking into account total overall objectives such as; thermodynamic efficiency, capital investment, operations and maintenance cost, project life, etc. However, they do represent important considerations towards "total" optimization when designing the P&D Plant.

Project Phase-2: Pre-Engineering & Qualification is currently focusing on further improved thermodynamic efficiency and optimization of plant size with respect to total capital investment (CAPEX). We are also collaborating on turbine development for technology migration from the gas turbine environment that will permit raising turbine inlet temperatures (TIT) and attaining "medium-term" thermodynamic efficiency of ~55% (US-DOE, 2005).

INTRODUCTION

The purpose of the ZENG Project Phase-1 was to gather information and propose a "Base Case" zero-emission plant that is appropriate for the pilot and demonstration (P&D) phase of technology development. A main criterion was to use components compatible with an investment decision being made in 3Q-2006 and plant commissioning in 2008.

Furthermore emphasis has throughout been placed on ensuring that such a P&D Plant would provide the necessary knowledge and experience to permit construction for "commercial" power plants of 240-400 MW_e (net export) in the 2010-2014 timeframe.

Such a goal for commercialization in the "medium-term" necessitates attaining power plant thermodynamic efficiency ~55% and ensuring that specific CAPEX is significantly reduced compared with what we estimate for the initial proposed nominal 40 MW P&D Plant.

There still remains considerable scope for optimizing and integrating the CES Gas Generator (GG) within a total balance of plant concept: the "Base Case" described extensively in the Phase-1 Report (Hustad *et al.*, 2004) has already been further developed and improved with respect to thermodynamic efficiency, as described in this paper.

We are also confident that a focussed effort in Project Phase-2 will enable a reduction in CAPEX as we continue to optimize plant integration and work closely alongside the main equipment suppliers.

Furthermore, there continues to be a need for work regarding integration with CO2-handling, interim storage, transportation and commercial sale of CO2 for enhanced oil recovery (EOR), as described by Sæther and Hustad (2005).

DESIGN BASE FOR 40 MW P&D PLANT

Proposed Plant location is on reclaimed "brown field" land made available at the Energy Park, Risavika, shown in Fig. 1. Selection of the P&D Plant nominal design capacity equal to 40 MW_e (net export) corresponds to ~100 MW_t thermal power from the GG.

Fig. 1: *Aerial view of reclaimed land area at the Energy Park, Risavika, nr. Stavanger, South Norway. Highlighted rectangle shows proposed location for the P&D Plant.*

This size of plant was initially chosen as being a reasonable compromise between development risk, economy of scale, CAPEX and technology status. It also provides a useful "next-step" on the path to commercialization from the 5 MW_e Kimberlina Test Plant that CES started commissioning near Bakersfield, Ca. during 4Q-2004.

The GG thermal power output scales with cross-sectional area: for the proposed P&D Plant the current (20 MW_t) GG diameter increases by a factor of 2.4—whilst length remains the same. This is considered to be within practical limits for scaling from the on-going test and operating experience.

Natural Gas (NG) is made available to the Stavanger region by *Lyse Gass AS* through a recently laid 10-inch diameter sub-sea pipeline from Kårstø with shore landing adjacent to the proposed plant site as indicated in Fig. 1.

Fuel Composition	Concentration (%-mol)
Methane (CH_4)	88.54
Ethane	7.71
Propane	0.50
i-Butane	0.03
n-Butane	0.04
Nitrogen (N_2)	0.69
Carbon dioxide (CO_2)	2.49

Table 1: *Summary of Fuel Composition for Natural Gas (NG). For the economic analysis we have assumed NG fuel cost to be 85 øre/Nm^3 (3.29 \$/GJ).*

The fuel gas in Table 1 has heat value (LHV) assumed to be 39.8 MJ/Nm^3 (equivalent to 47.7 MJ/kg) and a line pressure in the range 120-180 bar. With the "Base Case" this will be reduced to 94 bar for the GG and 30 bar for reheat (RH) combustion.

PROCESS DESIGN PHILOSOPHY

The process design was based on "current technology" and required that all major equipment items should be commercially available. We utilize a conventional cryogenic air separation unit (ASU), as shown in Fig. 2, to supply pure oxygen to the GG—this being the most cost-efficient commercial method available to date.

For the power train we employ a conventional steam turbine coupled to an electric power generator; if necessary through a speed reducer.

The high-pressure (HP) turbine inlet steam temperature is restricted to 565 ºC, with an increase to 705 ºC for the intermediary-pressure (IP) turbine (as is acceptable from potential suppliers). The low-pressure (LP) steam turbine exhaust flows to a vacuum condenser with 0.08 bar pressure.

A condenser pressure of 0.04 bar was investigated but this would have considerably increased the condenser size; bearing also in mind that the presence of CO_2 gas in the condensing steam will significantly increase the heat transfer resistance across the condenser compared to a conventional vacuum steam condenser. Furthermore it is advisable to keep the steam conditions upstream of the condenser above saturation level, to avoid corrosion (or erosion-corrosion) on turbine internals.

Fig. 2: *Schematic view of the Air Separation Unit (ASU) adjacent to the main P&D Plant building.*

The mass flow and energy balance data necessary for the selection and dimensioning of the process equipment, fuel feed, utilities consumption, etc., are generated by CHEMCAD (see www.chemstations.net). This includes comprehensive subroutines calculating thermodynamic, physical and transportation properties for the actual mixtures of the fluids involved in the main process, as well as in the utility systems.

We experienced some variation in the results depending on the simulation subroutine models utilized. These originated from differences in the calculated physical properties for the CO_2 / steam mixtures within the lower pressure and temperature regimes. Subsequent discussions have confirmed that there would appear to be limited reliable data available in this region. This means that process data and equipment parameters in the low-pressure (sub-atmospheric) regime should be treated as preliminary for the time being.

Intermediate steam data is based on thermodynamic efficiency specifications obtained from recognized suppliers of steam turbines or "state-of-the-art" efficiency properties for such equipment, as indicated in Table 2. Efficiency factors for gas compressors are based on catalogue values.

Component	Efficiency Factor
HP turbine	0.89
IP turbine	0.90
LP turbine	0.93
Electric Power Generator	0.95

Table 2: *Assumed power train efficiency factors.*

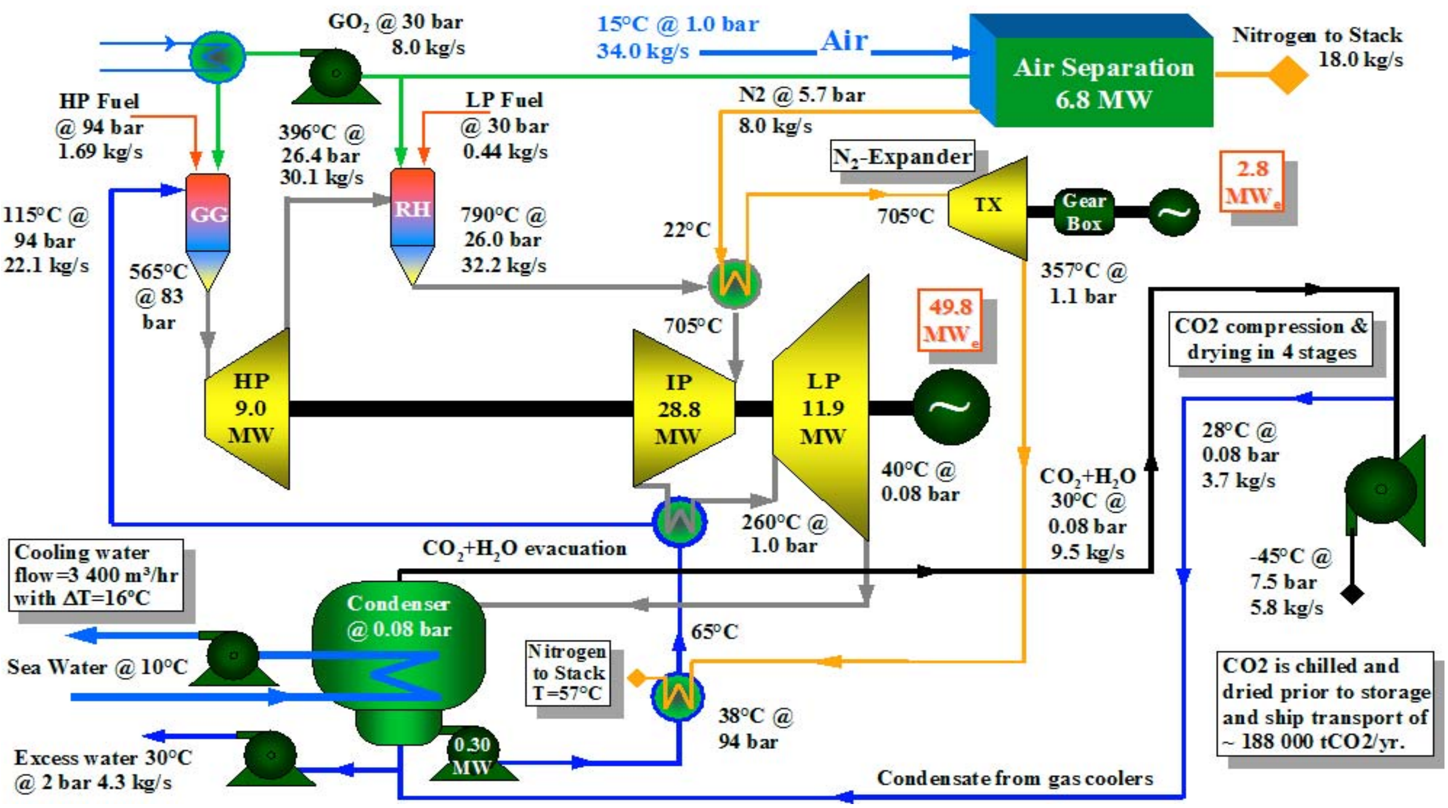

Fig. 3: *Process flow schematic for "Base Case" configuration with 42 MW$_e$ net output and cycle efficiency of ~ 38%.*

DESCRIPTION OF BASE CASE PROCESS

The NG fuel is supplied at 94 bar to the GG injection nozzles through a filtering and pressure reduction control station (see Fig. 3). The gaseous fuel and pure oxygen are combusted in combination with injection of water in a complex manifold and nozzle system; establishing near ideal conditions for stoichiometric combustion and temperature control within the combustor section of the GG shown in Fig. 4.

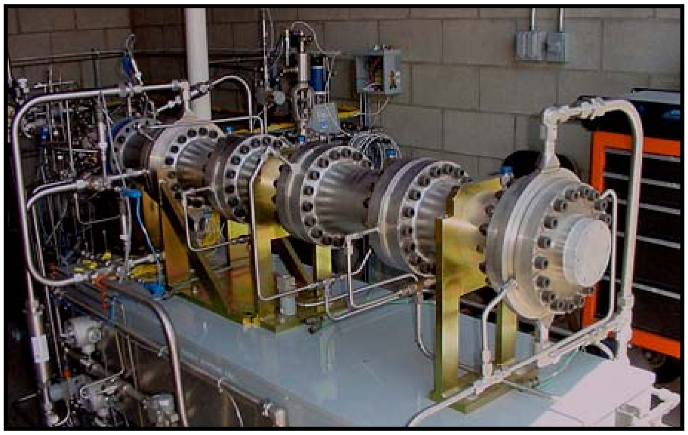

Fig. 4: *The 20 MW_t CES Gas Generator (GG). Combustor section is at far end followed by 4 sequential water-cooldown sections. Closest to observer is the downstream endplate that provided back-pressure during testing 'in lieu' of HP turbine.*

The GG exit pressure is controlled at 83 bar by the rate of fuel and oxygen flow. The process "drive" gas (CO_2 / steam) temperature is maintained at 565 ºC by the water-injection rate in the cooldown sections. The GG wall temperature is controlled by the flow of water through internal cooling passages within the housing.

The process gas stream is routed through the HP turbine and expanded to the outlet pressure at 26.4 bar and 396 ºC. The HP turbine shaft duty is 9.0 MW.

Next the process gas temperature is raised to 790 ºC using a reheat (RH) combustion chamber operating at 26 bar pressure and fed with NG fuel and oxygen at near stoichiometric ratio (Chorpening *et al.*, 2003). The process gas stream at the RH outlet comprises a mixture of 17% CO_2 and 83% H_2O (steam) based on %-weights.

Before the IP turbine the process gas passes through a nitrogen gas heater and is cooled to 705 ºC; close to currently maximum acceptable IP turbine inlet temperature (TIT). The IP turbine expands the process gas from 26 to 1 bar and a temperature of 260 ºC. The IP turbine shaft duty is 28.8 MW.

The nitrogen gas (partially taken from the ASU) is expanded in a N_2-turbine expander from 5.7 bar (705 ºC) to 1.1 bar (357 ºC) producing 2.8 MW_e additional power.

Next the process gas is led to the LP turbine where it is expanded to the condenser pressure of 0.08 bar and a temperature of 40 ºC. This is maintained sufficiently above steam saturation temperature, in order to avoid corrosion problems in the steam turbine and exhaust channels. The LP turbine shaft duty is 11.9 MW. The total turbine duty is 49.8 MW, whilst the electric generator efficiency is assumed to be 95%.

The exhaust steam from the LP turbine is condensed in a seawater-cooled condenser. In addition to CO_2 / steam mixture, the flow to the condenser contains a small amount of oxygen and a trace of carbon monoxide. The concentrations of unburned hydrocarbons and NOx are anticipated to be essentially zero. At an absolute pressure of 0.080 bar, partial pressures of the main components are 0.0065 bar for the CO_2 and 0.0735 bar for the steam (at which pressure the condensation temperature is estimated to be 39.9 ºC).

The seawater flow requirement for the condenser is calculated to be ~3,400 m^3/h with assumed cooling-water inlet temperature of 10 ºC which is standard Norwegian West Coast North Sea.

"Base Case" Cycle Summary Data	
Thermal power input	111.0 MW
Gross power output	52.6 MW
Parasitic power*	10.7 MW
Net power	42.0 MW_e
Overall cycle efficiency	~ 38%
Fuel consumption	7 670 kg/h
Oxygen consumption	28 800 kg/h
Cooling water flow (total)	~ 4 300 m^3/h
Excess water production	15.5 m^3/h
HP Turbine inlet pressure	83 bar
HP Turbine inlet temperature	565 ºC
HP Turbine exhaust temperature	~ 396 ºC
IP Turbine inlet pressure	25.9 bar
IP Turbine inlet temperature	705 ºC
LP Turbine inlet temperature	260 ºC
LP Turbine inlet pressure	1.0 bar
LP Turbine exhaust temperature	40 ºC
Condenser pressure	0.08 bar†

Table 3: *Summary Data for "Base Case" Configuration.*

* The "parasitic" power also includes electric energy consumption for the ASU, oxygen and CO2-compressors, as well as cooling-water supply pumps.

† The condenser pressure was also increased from 0.08 bar to 0.15 bar due to recommendation from the CO2-compressor suppliers. A higher pressure could significantly decrease the dimensions and costs for both the compressors and intercoolers. This increase in condenser operating pressure would have reduced the cycle efficiency from 37.8 to 36.3%.

OPTIMIZED PROCESS DESCRIPTION

To date the practical limit for steam temperature from conventional boilers has been around 565 ºC; and no strong market incentive has existed for the development of steam turbines with higher temperatures. The CES GG presents new possibilities for cycle improvement with increased steam temperature and process pressure. However, steam turbines will not accommodate significant increase of TIT without introduction of secondary flow and internal blade-cooling, together with utilization of sophisticated materials.

But current gas turbine (GT) technology is already operating at significantly higher TIT, albeit at comparatively lower intermediary pressures: these present an excellent opportunity for inclusion as IP turbines in an "Optimized" process scheme as shown in Fig. 5. In such cycles the IP turbine TIT may potentially be elevated to 1,450 ºC thereby resulting in a very substantial increase in cycle efficiency.

However for practical purpose this would require some—but still limited—redesign of a suitable gas turbine. Commercial availability of such GT's is still considered being "a few years" ahead of the initial demonstration goals for the current P&D Plant (but see also US-DOE, 2005) and requires commercial drivers for the equipment suppliers.

To provide an indication of the "near-term" potential for improvement of thermodynamic efficiency we have maintained a TIT of ~700 ºC whilst including here an optional process scheme based on cycle integration using a RR-WR21 recuperated gas turbine as proposed by Phillips (2004).

Included in the "Optimized" configuration is also a "double" Rankine steam cycle, together with further integration of the air compressor and nitrogen expansion from the ASU. (N_2-expansion is here principally the same as for the "Base Case", but now with the total nitrogen flow routed through the expander, thereby increasing power production and cycle efficiency.)

The benefit of the double Rankine cycle is that separation of CO2 occurs at a pressure of 3.0 bar, thus reducing the number of CO2-compressors and dimensions for the CO2-handling equipment. Furthermore the LP "pure" steam Rankine cycle can now have a reduced condensation pressure (0.03 bar) compared with the "Base Case" process (0.08 bar)—this too contributes significantly to overall cycle efficiency.

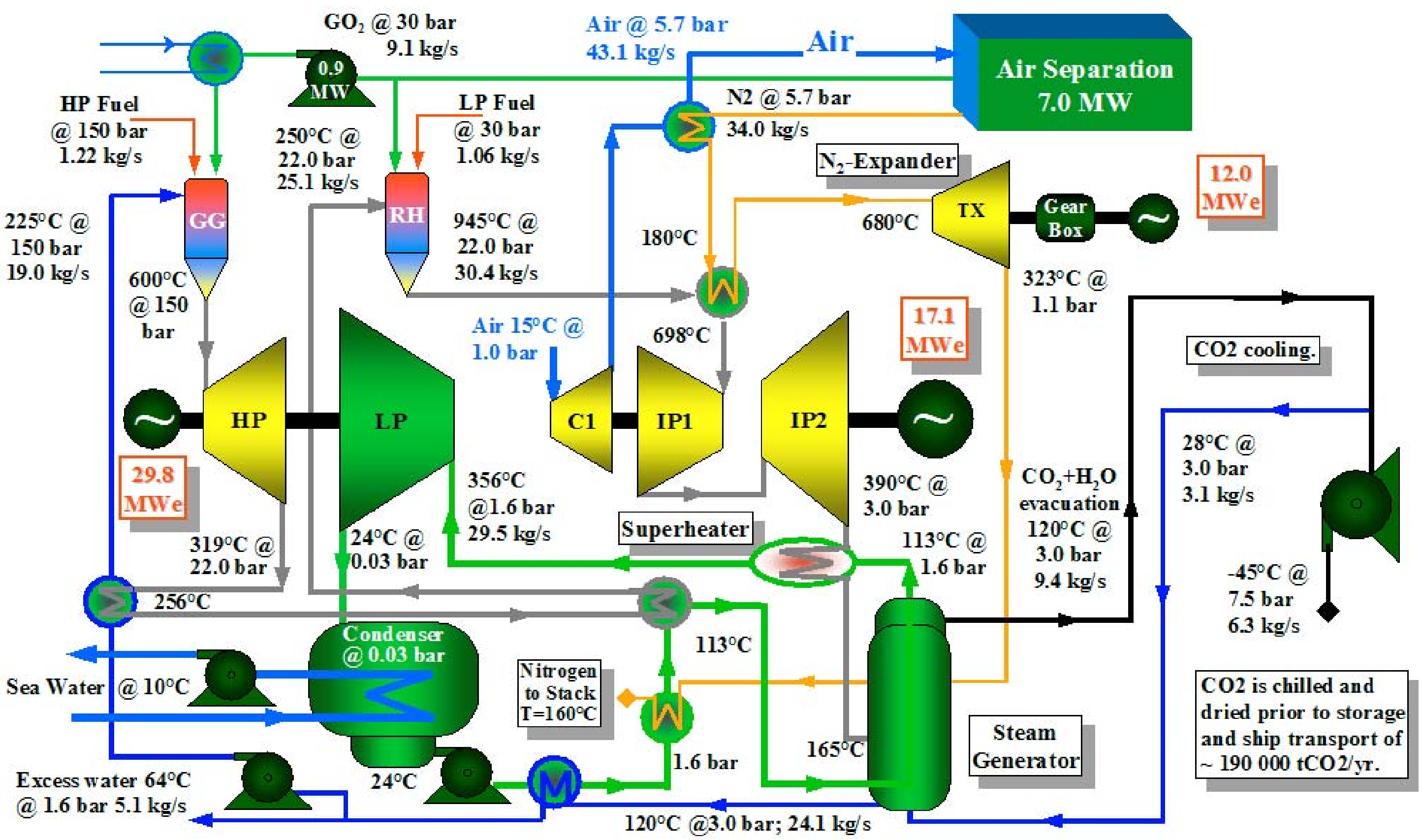

Fig. 5: *Process flow schematic for "Optimized" configuration with 50.5 MW_e net output and cycle efficiency of ~ 45%.*

The GG, HP turbine, reheater (RH), nitrogen heater and turbine expander, feed-water heaters and oxygen compressors are principally the same as in the "Base Case". While the LP cycle is now a conventional "Cogen" condensing steam turbine. Furthermore the compressor (C1) delivers compressed air to the cryogenic ASU (see Fig. 5).

The GG is here operated at 150 bar and therefore a separate pressure reduction station for initial fuel handling is not necessary. Total fuel feed to the GG injection nozzle is 1.22 kg/s. The fuel energy supplied is 61 MW and the outlet energy flux is 80 MW. The process gas at the GG exit contains approximately 5.3 %-mol CO_2 while the combustion generates 3.35 kg/s CO_2 and 2.74 kg/s steam.

The process gas at 150 bar and 600 ºC is routed to the HP turbine where it is expanded to 22 bar and ~ 320 ºC. With turbine efficiency maintained throughout as specified in Table 2 the HP turbine stage shaft duty is now 12.4 MW.

The process gas is passed through a heat exchanger to raise GG feed-water temperature to 225 ºC. The process gas stream also heats feed-water to the LP steam generator up to near vaporization temperature of 113 ºC.

Next the process gas stream is routed to the RH operating at 22 bar and where the process gas temperature is raised to 945 ºC by stoichiometric combustion of fuel gas with oxygen. Fuel consumption is 1.06 kg/s (equivalent to 53 MW fuel energy). The RH combustion produces 2.91 kg/s CO_2 and 2.38 kg/s steam; the gas stream now comprises 6.26 kg/s CO_2 and 24.1 kg/s steam, with the CO_2 concentration being 9.5 %-mol and energy stream flux is 116 MW.

Next the process gas stream flows to the N_2-turbine expander heater where 34 kg/s (all available) nitrogen is heated to 680 ºC whilst the process gas temperature is reduced to ~ 700 ºC in order to be compatible with TIT for the IP1 gas turbine.

The N_2-turbine expander produces 12 MW_e power and has an exhaust temperature of 323 ºC; this is heat-exchanged against feed-water in the LP steam cycle, reducing temperature of the nitrogen exhausting to atmosphere to ~ 160 ºC. (Which is still comparatively high and we should be able to make better use of this with further optimization!)

The IP gas turbine is based on a modified design derived from a recuperated GT (e.g. Rolls-Royce WR-21) where the recuperator is removed and principally replaced by the gas re-heater. The process fluid expands from 22 to 3.0 bar—through two stages (IP1 and IP2)—with temperature decrease from 705 to ~ 390 ºC. Normal exhaust condition for the WR-21 is atmospheric pressure, hence the last turbine stage(s) will need to be modified or removed. The turbine shaft duty is estimated to be ~17.1 MW.

The IP2 exhaust steam is led to the steam superheater for the LP steam cycle, where saturated steam from the steam generator is heated from 113 to 356 ºC. The steam generator is a conventional unit, as normally utilized for production of clean steam from "unclean" steam sources.

The superheater for the produced clean steam is also a conventional free-standing unit, comprising of tube banks in a countercurrent arrangement. The exhaust steam is routed to the steam generator, where the steam fraction is condensed by heat-exchange against the (boiling) feed-water to the steam generator—mol-fraction of steam in the process fluid is 0.90.

The superheated steam (at 1.6 bar and 356 ºC) is routed to the LP turbine where the steam is expanded to condenser pressure at 0.03 bar and ~ 24 ºC. The LP turbine stage duty is estimated to be 17.6 MW_e.

The exhaust steam from the LP turbine is condensed in a seawater-cooled condenser. At an absolute pressure of 0.03 bar the condensation temperature for the steam is 24.1 ºC.

In this preliminary study we have not to date included recompression of CO_2 from 3.0 bar to 7.5 bar followed by chilling to -45 ºC, thereby making it completely ready for interim storage and subsequent ship transportation. However this will only have a small impact on the total cycle efficiency.

"Optimised" Cycle Summary Data	
Thermal power input	114 MW
Gross power output	58.6 MWe
Parasitic power	8.1 MWe
Net power	50.5 MWe
Overall cycle efficiency	~ 45 %
Fuel consumption	8 210 kg/h
Oxygen consumption	32 760 kg/h
Cooling water flow (total)	~ 4 300 m^3/h
Excess water production	19.0 m^3/h
HP Turbine inlet pressure	150 bar
HP Turbine inlet temperature	600 ºC
HP Turbine exhaust temperature	~ 320 ºC
IP Turbine inlet pressure	22.0 bar
IP Turbine inlet temperature	698 ºC
IP Turbine exhaust temperature	~ 390 ºC
CO2 / Steam Condenser pressure	3.0 bar
LP Steam Rankine Cycle	
LP Turbine inlet pressure	1.6 bar
LP Turbine inlet temperature	356 ºC
LP Turbine exhaust temperature	24 ºC
Steam Condenser pressure	0.03 bar

Table 4: *Summary Data for "Optimized" Configuration.*

In recent (unpublished) work we have further increased cycle efficiency by 2-3%-point. And therefore now consider our main focus in Project Phase-2 should be to ensure similar progress in reducing the plant specific CAPEX ($/kW) and ensuring plant availability of ~ 95%, as is achievable with typical steam cycles.

ECONOMIC ANALYSIS

Within the present study we have identified and cost-estimated all major components for the "Base Case" configuration and made comparison with a conventional NG Combined Cycle (NGCC) Power Plant. Included here are also cost-factors based on accumulated project experience in Norway—with high labor costs and strong local currency these can typically lead to estimates that are 20-25% above US Gulf Coast estimates!

The economic model permits input of all main power plant parameters; CAPEX, operating costs, internal rate of return, project duration, efficiency, net power generation, sale of CO2, etc. Model output is derived using annualized cash flow and calculates cost of electricity (CoE) by prescribing a zero net present value. All modeling is pre-tax. We have made the following generalized basic assumptions:

- 10% discount rate and project economic life of 25-yrs.
- Fuel cost is 85 øre/Nm^3 NG (equivalent to 3.29 $/GJ).
- For P&D Plant we assume 60% financed debt at 5% interest. This reflects some "goodwill" from the Norwegian government's interest to help promote development and demonstration of such "new" power generation technology[1].
- For comparison between the conventional NGCC Reference Plant and a "commercial" ZENG-CES plant we revert to assuming 100% equity financing.
- Assume two years for total investment and construction.
- Assume 6 weeks for commissioning during first year.
- Exchange rate is 6.50 NOK/US$.
- CoE is expressed in mills/kWh (1,000 mills/US$) and in Norwegian currency as øre/kWh (100 øre/NOK).

Using Reference CoE from the NGCC without CO2-capture[2] we can also calculate a CO2-capture cost (in US$ per ton of CO2) for comparison with sale of CO2 for EOR to a CO2-aggregator / transporter / oilfield operator.

For the "Base Case" 42 MW$_e$ P&D Plant (inclusive of the ASU) we have total CAPEX of $97.7 million (equivalent to 635 MNOK). With further focus on cost optimization in Project Phase-2 and with economies of scale, we believe there is considerable opportunities for reducing this CAPEX.

The incremental CoE for the "Base Case" is estimated to be +26.0 mills/kWh compared with the 400 MW Reference CoE. Alternatively, the plant would need to recover a CO2-capture cost of $28.0 /ton (at perimeter fence) in order to be competitive with electricity from the Reference Plant[3].

For the "Optimized" Configuration we have estimated total CAPEX to be $109.9 million (equivalent to 714 MNOK). Net export power is 50.5 MW$_e$ resulting in incremental CoE of +19.0 mills/kWh compared with the Reference CoE. Alternatively the "Optimized" P&D Plant must sell the CO2 at a price of $19.3 /ton (at perimeter fence) to cover extra costs.

The CO2-liquefaction plant (with storage facilities) and transportation to offshore platform are outside Scope of Work for the P&D Plant (but see Sæther and Hustad, 2005). However, as described by Hustad and Austell (2004) one may conservatively account for this incremental cost in CO2-handling by assuming an additional ~ $12 /ton whereby delivered price will be ~ $31 /tCO2.

Recent alternative studies have indicated delivered cost for CO2 on North Sea platform to be in the range from $35 /tCO2 as proposed by Elsam / Kinder Morgan, CENS Project (Markussen *et al.*, 2002). Alternatively up to $48 /tCO2 as presented by Statoil for proposed CO2-flooding at Gullfaks.

In the near- to medium-term (2010-2014) we have identified cycle optimization opportunities that could ensure plant efficiency of ~ 51%. Furthermore, we believe cost-optimization, economies of scale and early commercial introduction can contribute to ensure an additional one-third reduction in specific CAPEX. This would entail that a "100% equity financed" commercial 240 MW ZENG-CES Power Plant could have a CO2-capture cost (at perimeter fence) of $17.9 /tCO2 whilst delivering 0.80 mtCO2/yr for EOR.

In this context the key economic variable is the market price of crude oil which determines the sales value of CO2 for EOR. Again we may assume, using larger volumes, that delivered cost of CO2 at the offshore platform from such a 240 MW ZENG-CES Power Plant could be ~ $28 /tCO2. Thus, even with the current fiscal regime in the North Sea—which is not yet optimized to create incentives for CO2-EOR—the pre-requisite crude oil price needed to sustain project economics would be in the range $25-$28 /bbl (see Hustad and Austell, 2004).

In the medium- to longer-term (2012-2015) we foresee technology improvements[4] and economies of scale that should permit a 400 MW ZENG-CES Power Plant to operate with ~ 55% efficiency and have specific investment cost below 1 400 $/kW. Economic modeling for such a plant suggest it would have a CO2-capture price of ~ $10 /tCO2 whilst producing 1.25 mtCO2/yr. The long-term goal is to achieve 60% plant efficiency by 2015 (US-DOE, 2005).

[1] In 2004 the Norwegian government specifically set aside a fund of $310 million to promote P&D Power Plants with CO2-capture & storage (CCS). They have also indicated that as cost-effective technologies emerge, then they shall be willing to further add to this level of support if necessary.

[2] The Reference Plant assumes a new build on the West Coast of Norway with specific CAPEX of 745 $/kW installed. We obtain CoE at 35.2 mills/kWh (22.9 øre/kWh) exclusive of CO2-emissions. We assume that the Reference Plant will need to purchase CO2-credits for an additional cost of $12 /tCO2 starting in 2008 and rising linearly to $24 /tCO2 at end of project economic lifetime. With these assumptions we derive a Reference CoE equal to 40.7 mills/kWh (26.5 øre/kWh). For further details see Hustad *et al.* (2004).

[3] Here we assume "Base Case" P&D Plant will capture 100% of its CO2-emissions at a pressure of 7.5 bar. This will subsequently need to be dried and cooled to -45 ºC (near triple point) for liquefied storage and ship transportation.

[4] Specifically we here foresee commercial introduction of; (i) oxygen membrane technology, and (ii) blade-cooling to permit increased TIT for the HP and—in particular—the IP turbine expansion stages.

Fig. 6: *Schematic sketch of the proposed 40 MW (nominal) Pilot & Demonstration Power Plant at the Energy Park in Risavika.*

In this timeframe we may also assume that the cost of CO2 transportation from the power plant perimeter fence out to an oilfield will be aggregated and handled in a more cost-effective manner through a dedicate CO2-infrastructure. We therefore estimate future delivered price for CO2 to be ~ \$17 /ton. The sustaining market price of crude oil would then need to be ~ \$22 /bbl.

The long-term market expectation is that crude oil will be above \$35 /bbl; highlighting a substantial commercial upside on the basis of EOR. Furthermore, CO2-credits are already trading at ~ \$20 /ton on EU and US exchanges. Thus there are already two strong economic incentives to develop zero-emission fossil power generation.

CONCLUSIONS

Results suggest that a ZENG-CES Power Plant, in combination with sale of CO2 for EOR could provide 3.2 TWh of base load (+8 000 hours per year) zero-emission electricity by 2011. And will, through project economic lifetime in a "carbon-constrained" market be more cost-effective than a conventional power plant having to pay for its CO2 emissions.

Zero-emission power in combination with recognized CO2-EOR potential creates a business opportunity providing an important contribution to the use of NG in Norway, life-extension for the mature oil reservoirs on the Norwegian Continental Shelf, and technology export opportunities.

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