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FIRST GENERATION GRAZ CYCLE POWER PLANT FOR NEAR-TERM DEPLOYMENT

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ABSTRACT

Carbon Capture and Storage (CCS) is a recognized technology pathway to curb the increasing emissions of carbon dioxide (CO_2) from the power generation sector. But most available technologies are still on the study or laboratory-scale level, so that considerable R&D efforts are needed to achieve commercialization level.

The Graz Cycle originally presented in 1995 by Jericha [1] is an oxyfuel technology and promises highest efficiency using state-of-the-art turbine materials and improved thermodynamic developments in a comparatively complex interaction of rotating machinery, condensers and heat exchanger components. But although detailed conceptual design for all main components has been presented, there is still a large step towards a Graz Cycle pilot demonstration plant.

In order to facilitate construction of a demonstration plant we consider the performance of a near-term Graz Cycle process design based on modest cycle data and available turbomachinery components using a simplified flow scheme. The work is supported by on-going development work for a first generation oxyfuel turbine that has already been under-taken by Clean Energy Systems, Inc. [2]. Their further work on a second generation oxyfuel turbine received \$30 million funding support from the U.S. Department of Energy in September 2010 [3].

Two near-term Graz Cycle plants are presented based on basic and advanced operating conditions of the proposed commercially available turbine. Besides the turbine the additional equipment for a first-generation cycle is discussed. The predicted optimum net efficiency is 23.2 % (HHV).

A near-term zero-emission power plant can only be commercially attractive if it will be deployed in a niche market. Therefore an economic analysis commensurate with an early pre-FEED conceptual study is carried out for the U.S. Gulf Coast where revenue from multiple product streams that could include power, steam, CO_2 and water, as well as argon and (potentially) nitrogen from the ASU is provided.

The economic analysis suggests that a capital investment of \$94 million can secure construction of a 13.2 MW_e zero emission oxyfuel power plant and yield a 14.5% (unlevered) return on capital invested.

NOMENCLATURE

ASU	Cryogenic Air Separation Unit for O_2 supply
CCS	Carbon Capture and Storage
CES	Clean Energy Systems Inc.
CES-J79	Cycle for testing the J79 turbine at KPP
CO_2	Carbon dioxide
DOE	U.S. Department of Energy
EOR	Enhanced Oil Recovery
EBITDA	Pre-tax earnings for unlevered analysis
FEED	Front-End Engineering & Design
FG	Fuel Gas
GC-B-J79	Basic Graz Cycle using the J79 turbine
GC-AD-J79	Advanced Graz Cycle using the J79 turbine
GG	Gas Generator (CES combustor)
HPT	High Pressure Turbine
HRSG	Heat Recovery Steam Generator
HTT	High Temperature Turbine
IRR	Internal Rate of Return
J79	A CES modified GE turbine expander
KPP	Kimberlina Power Plant, nr. Bakersfield, Ca.
LPT	Low Pressure Turbine
NG	Natural Gas
NO_x	Nitrogen Oxides
NPV	Net Present Value
Mcf	One thousand cubic feet
O&M	Operations and Maintenance
RG	Reservoir Gas (with high CO_2 and low-btu)

INTRODUCTION

It is now 25 years since publication of the United Nations Brundtland Report [4] on sustainable development, and in the following year the 1988 Toronto Conference on the Changing Atmosphere noted *"... humanity is conducting an unintended, uncontrolled, globally pervasive experiment whose ultimate consequences could be second only to a global nuclear war"*. These were the first indications of media recognition regarding climate change that has subsequently evolved to become the current focus on fuel efficiency, reduced emissions to atmosphere and Carbon Capture & Storage (CCS).

The three main CCS technologies are categorized as pre-combustion, post-combustion and oxyfuel combustion capture. Among the latter, the oxyfuel Graz Cycle can potentially achieve a high efficiency along with 100% capture of CO_2 and no NO_x emissions.

The Institute for Thermal Turbomachinery and Machine Dynamics at Graz University of Technology has worked since 1995 on the Graz Cycle which is an oxygen-based fuel cycle with internal combustion of hydrocarbon fuels that permits cost-effective capture of the combustion generated CO_2 by condensation of the steam / CO_2 working fluid. The cycle has been extensively described through a long list of publications, e.g. [5 - 15].

In [1, 5] thermodynamic studies were presented on a cycle with internal combustion of methane and pure oxygen. In [6 - 8] the cycle was adapted to the firing of syngas from coal gasification, and cycle modifications were proposed leading to a working fluid with two-thirds CO_2 and one-third steam. A layout of the turbomachinery components was presented for a pilot plant of 75 MW net power output. In 2004 and 2005 the cycle scheme was rearranged similar to the original version [1, 5] with a working fluid consisting of three quarter steam and one quarter CO_2 [9, 10]. In 2006 at the ASME IGTI conference the authors published a design proposal for a CO_2 retaining gas turbine of 400 MW_e where condensation of the working fluid takes place at atmospheric pressure [11]. In 2007 stress and rotor dynamic design improvements to the high speed compressor shaft where published at CIMAC conference in Vienna [12] and a design comparison to competing proposals was presented at the ASME IGTI conference in Montreal [13]. In 2008 an updated cycle using advanced turbomachinery was presented demonstrating a power increase to 600 MW_e [14]. Also the integration of a Graz Cycle plant and a coal gasification unit showed advantages compared to conventional IGCC plants with pre-combustion decarbonization [15].

The highest efficiency of the Graz Cycle can be achieved using state-of-the-art turbine materials and thermodynamic developments in a comparatively complex interaction of rotating machinery, condensers and heat exchanger components. Furthermore, it is well recognized that a significant development effort and further research is needed before construction of such an optimal Graz Cycle plant.

However, here it is suggested to use commercially available plant equipment for a near-term demonstration plant following the example of Clean Energy Systems, Inc. (CES). CES has been operating a prototype oxyfuel power plant located near Bakersfield, California since 2005 [16]. The main difference between the Graz and the CES cycle lies in the composition of the steam / CO_2 working fluid due to the inclusion of a recycle compressor in the Graz Cycle. This increases the overall concentration of CO_2 in the working fluid. Thermodynamically such a regenerative process will improve thermal efficiency but is technically more challenging. Invariably there is an optimization process regarding the extent to which one should recycle while changing the thermodynamic properties of the working fluid.

CES is currently working on the implementation of a GE J79 turbine as a steam-driven turbine expander [2, 17]. Further work for a next generation oxy-turbine that is based on the Siemens SGT-900 (formerly Westinghouse 251B) has also received funding from the U.S. Department of Energy [3].

Because the main components of the CES cycle are similar to the ones needed for the Graz cycle, we here investigate how a Graz Cycle demonstration plant can benefit from the CES work. We present a first generation Graz Cycle plant using a simplified flow scheme and modest cycle data based on the J79 turbine expander. The additional turbomachinery components are discussed in detail and power data and cycle efficiencies for a pilot plant are presented.

An economic analysis for niche market application providing power and CO_2 for enhanced oil recovery (EOR) using oil field reservoir gas – having a high CO_2 content – as fuel concludes this study. In the power sector, investment strategy has to be conservative because multi-billion dollar projects will bind capital over an economic life of between 20 to 50 years, thereby creating a barrier to entry for new technologies. To address concerns regarding introduction of "unproven hardware", the economic analysis is structured in such a manner that we reduce project size and use commercial contracts to mitigate risk across multiple stakeholders that all benefit through successful deployment of the technology.

This roadmap also identifies a pathway for early commercial introduction of a near-term Graz Cycle power plant by minimizing costs and risks through utilizing predominantly conventional turbomachinery equipment.

DESCRIPTION OF THE GRAZ CYCLE

The basic principle of the Graz Cycle was presented by H. Jericha in 1985 [18] for solar generated oxygen-hydrogen fuel. In 1995 this was extended to fossil fuels [1, 5] and was at that time the first proposal for this type of oxyfuel power cycle having CO_2 capture. Any hydrocarbon-based fuel gas is proposed combusted with oxygen so that, neglecting small impurities, only the two combustion products CO_2 and H_2O are generated. The cycle working fluid comprising of CO_2 and steam allows for comparatively simple (and cost-effective) CO_2 separation through conventional condensation. Furthermore, a closed-cycle oxyfuel process design can recover much of the latent heat of vaporization from the combustion process that open cycle plants emit in the flue gas; this partly compensates for the additional work needed for oxygen production.

Fig. 1 shows the principle flow scheme of the Graz Cycle with the main cycle data as presented at the ASME IGTI conference 2005 [10], since this variant is better suited to be modified for a first generation plant.

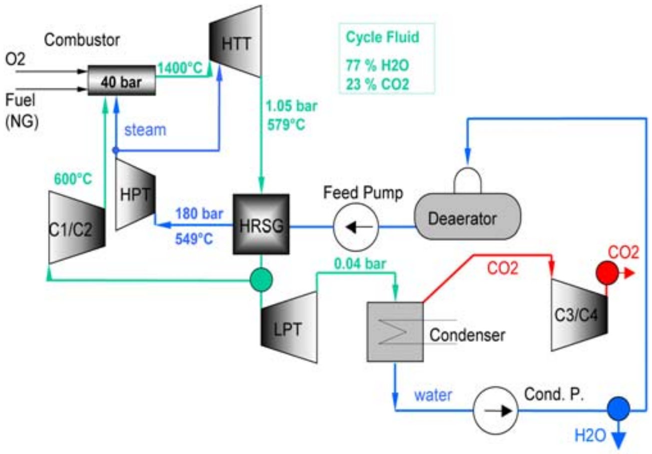

Fig. 1: Principle Process Design for the Graz Cycle Plant [10].

The cycle consists of a high temperature Brayton cycle (compressors C1/C2; Combustor; High Temperature Turbine, HTT) in combination with a low temperature Rankine cycle (Low Pressure Turbine LPT; Condenser; Heat Recovery Steam Generator, HRSG; High Pressure Turbine, HPT). The fuel, together with a nearly stoichiometric mass flow of oxygen, is fed to the combustor which is operated at a pressure of 40 bar. Steam, as well as a CO_2 / H_2O mixture, is supplied to cool the combustor burners and liner.

A mixture of about 74% steam, 25.3% CO_2, 0.5% O_2 and 0.2% N_2 (mass fractions) leaves the combustion chamber at a mean temperature of 1,400°C. The fluid is then expanded to a pressure of 1.05 bar and 579°C in the HTT. Cooling is performed with steam coming from the HPT at about 330°C (13.7% of the HTT inlet mass flow), thereby increasing the steam content to 77% at the HTT exit. It is quite clear that a further expansion down to condenser pressure would not end at a reasonable condensation point for the water component, so that the hot exhaust gas is cooled in the following HRSG to vaporize and superheat steam for the HPT; the temperature difference is 25°C at the superheater exit. But after the HRSG only 45% of the cycle mass flow is further expanded in the LPT. With a cooling-water temperature of 8°C (i.e. North Europe) the LPT exit and thus condenser pressure is 0.04 bar.

Gaseous and liquid phases are separated in the condenser. Following, the gaseous mass flow which contains the combustion CO_2 and about half of the combustion water, is compressed to atmosphere by C3 and C4 with inter-cooling and further extraction of condensed combustion water before being supplied for use and / or storage. At atmosphere the CO_2 purity is 96% while additional drying, for liquefaction or to pipeline specifications, is achieved during the final compression phase.

After segregating the remaining combustion H_2O, the water from the condenser is pre-heated, vaporized and super-heated in the HRSG. The steam is delivered to the HPT at 180 bar and 549°C. After expansion it is used to cool the burners and the HTT stages.

The majority of the working fluid – the return flow after the HRSG – is compressed using the main cycle compressors C1 and C2 with inter-cooling, and then fed to the combustion chamber at a maximum temperature of 600°C.

This cycle offers several advantages by allowing heat input at very high temperature, while expansion occurs to vacuum conditions so that a higher thermal (Carnot) efficiency can be achieved. Furthermore, less than half of the steam in the cycle releases its heat of vaporization by condensation. The major part is compressed in the gaseous phase and so recycling its heat content back to the combustion chamber.

The detailed flow sheet used for the thermodynamic simulation can be found in [10] and gives mass flow, pressure, temperature and enthalpy of all streams. The thermodynamic investigation of the plant resulted in a net efficiency of 52.6%, including oxygen supply and CO_2 compression.

For the Graz Cycle plant of 400 MW_e net power output the layout of turbomachinery was presented at the ASME IGTI conference 2006 [11]. Several components for this cycle are not standardized, primarily because of the unusual working fluid consisting of three parts H_2O and one part CO_2. However, the high temperature turbine (HTT) is similar to the expander of an air-breathing gas turbine, but the working fluid has a higher enthalpy drop given the same pressure ratio, and acts differently on the high-temperature materials. Furthermore the combustion chamber must burn the fuel gas with a nearly stoichiometric amount of oxygen in an atmosphere formed by CO_2 and steam, whereas in a conventional combustor the oxidizing and cooling medium is air. On the other hand low-NO_x technology is not needed because any NO_x formed (from small traces of nitrogen) is captured together with the CO_2. However, the working fluid compressors C1 and C2 have to cope with the higher enthalpy rise of the steam / CO_2 mixture so that a greater number of stages (or higher circumferential speed) is required, as has been described in [11].

The HRSG also needs to cope with the changed exhaust gas. A first layout by Giglmayr *et al.* [19] showed that it is comparable with a conventional HRSG regarding costs and technical efforts. The condenser is a critical component as condensation occurs in the presence of a non-condensable gas which reduces the heat transfer. Special care must also be addressed regarding increased corrosion from the combination of CO_2 and water which forms carbonic acid.

THE CES ZERO EMISSION POWER PLANT

Clean Energy Systems, Inc. (CES) has pursued the concept of a zero-emission power plant based on the oxyfuel technology for over 15 years [20, 21]. Fig. 2 shows a simplified schematic diagram of the process [2], the main cycle data are here taken from a variant presented in 2004 [22].

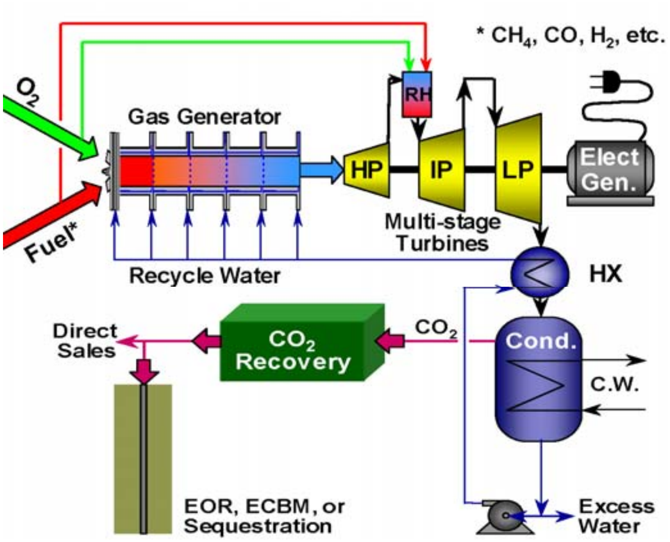

Fig. 2: Principle Flow Scheme of the Oxyfuel CES Cycle [2].

The gas generator (GG) is a key component of the cycle and much experience has already been acquired though operations of a 20 MW_{th} unit since 2005 and, more recently, of a commercial size 170 MW_{th} unit. The GG has been demonstrated to burn a wide variety of gaseous hydrocarbon fuels (including low-btu content) using pure oxygen at nearly stoichiometric conditions. Recycled water is used to cool the gas generator which typically produces a working fluid of about 88% steam and 12% CO_2 (mass fractions) at high temperature and pressure (design max. is 1,700°C and 103 bar). These combustion products can then drive a conventional or advanced high pressure (HP) steam turbine. After leaving the HP turbine, the mixture can be re-heated to conventional gas turbine temperatures (1,200°C) in a re-heater (RH) by internal combustion of additional fuel with near stoichiometric oxygen. After the RH the working fluid entering the intermediate-pressure (IP) turbine typically comprises 79% steam and 21% CO_2. The working fluid leaves the IP turbine at ~1 bar and enters a low-pressure (LP) turbine for its final expansion to vacuum conditions (~0.06 bar). The exhaust from the LP turbine flows through a feedwater pre-heater (HX) to heat the water that is separated from the process working fluid in the condenser and used to cool the GG. Also, in the condenser / separator section, water and CO_2 are separated by water condensation.

Whereas the HP turbine can be a conventional steam turbine, because of the high steam content of the working fluid and its operating parameters, the IP turbine is – similar to the Graz Cycle HTT – a modified gas turbine expander operating with a modified working fluid. Furthermore, the re-heater is a gas turbine combustion chamber which ensures a complete and nearly stoichiometric combustion of the fuel in the steam / CO_2 environment. The low pressure turbine has a CO_2 content of ~21%, but nevertheless it can be regarded as a conventional LP steam turbine (similar to the Graz Cycle LPT). The water pre-heater also uses standard heat exchanger technology, whereas the condenser has to take into consideration water condensation in the presence of the non-condensable CO_2 (again, similar to the Graz Cycle). Finally the GG is a non-conventional component, but its deployment has already been extensively demonstrated by CES up to a power level of 170 MW_{th}.

Comparing both cycles, many similarities are apparent, e.g. the flow from the re-heater/combustion chamber to the high temperature IP turbine and recirculation of the segregated water. But in the Graz Cycle the cooling of the working fluid takes place immediately after the HTT, before only about half of the mass flow is fed to the LP turbine. The remaining working fluid is recompressed and directly sent to the combustion chamber. In both cycles the water has to be pressurized and evaporated. In the CES Cycle the evaporation takes place in the GG which has a similar role as the HRSG of the Graz Cycle. The high-pressure steam of both cycles (in the CES Cycle with a small amount of CO_2) is then expanded in a similar HP turbine upstream of the combustion chamber. The higher complexity of the Graz Cycle – due to recirculation of the working fluid using recycle compressors – is balanced against higher cycle efficiency. This is achieved because less working fluid is fed to the condenser and so less heat is extracted from the cycle.

We therefore observe that the main components of both cycles have similar engineering specifications. Looking at the technical challenges and operating conditions, all turbines, the condenser and the combustion chamber can be considered as almost identical components. Only the Graz Cycle HRSG and the recycle compressors have no counterparts in the CES Cycle. This similarity suggests that both can benefit from related R&D done for each cycle.

CES has already acquired considerable experience with operations of their oxyfuel GG on multiple fuels. Starting in 1998 they built and demonstrated a laboratory scale 110 kW_{th} prototype and subsequently in 2002 a 20 MW_{th} GG. While in 2008 they demonstrated their 170 MW_{th} commercial offering. CES have also been operating since 2005 – with insurance – their Kimberlina Power Plant (KPP), near Bakersfield in California using the 20 MW_{th} GG in combination with a small steam turbine acting as the HP turbine of Fig. 2, the feed system and a condenser able to separate the non-condensable gases (mainly CO_2) from the turbine exhaust. This plant is a fully proven proof-of-concept and can also be considered to be a major milestone towards commercial deployment of the oxyfuel technology.

The GE J79 turbine was originally developed for the F104 Phantom fighters in the 1960's. It has been specifically re-engineered by CES for power generation [2, 17]. The original 17-stage axial compressor has been removed from the engine and a thrust bearing system added to compensate for the loss of the axial compressor loading. The working fluid is fed into the first stage nozzles through a volute manifold that is closely coupled to a set of turning vanes. The modified turbine expander provides approximately 32 MW (shaft) power output. While an additional single-stage turbine (based on the aero-derivative LM1500) expands the flow from 2.3 bar to atmosphere providing an additional 11 MW shaft output.

The combination of Main and Exhaust turbine will be employed in the first generation plant as an IP turbine. Power output will be similar to the original gas turbine, but the flow conditions will be different, with an inlet temperature of typically 760°C and pressure of 11.6 bar, compared with 927°C / 12.3 bar originally. This initially eliminates the need for blade-cooling which is another critical point in the development of oxyfuel turbines. Whereas in the full CES Cycle the steam / CO_2 mixture stems from the re-heater, the working fluid of the first generation plant will be provided directly by the 170 MW_{th} CES GG.

In order to improve cycle efficiency for future plants, a reheat combustor is currently being developed by CES based on the original GE J79 combustor cans. The steam / CO_2 working fluid at 315°C from an upstream GG is re-heated up to 927°C, the original inlet temperature of the GE J79. In the first test-runs stable efficient combustion has been achieved. But the oxygen ratio needed for a stable burn was higher than the stoichiometric ratio and the temperature distribution at combustor exit was non-uniform with a hot core [2]. These issues are being addressed by CES through further testing and development work.

FIRST GENERATION GRAZ CYCLE PLANT

CO2-Global is a zero emission power plant development company that is currently proposing deployment of first generation oxyfuel technology in the power range of 20 up to 150 MW_e at suitable location where revenues using CO_2 for enhanced oil recovery (EOR) can be obtained [23]. Such deployment would benefit by using a mixture of NG and available Reservoir Gas (RG) that has up to 85% CO_2 content.

Table. 1: Fuel Gas Specifications assuming 5% mix of Pipeline NG with the CO2-EOR Reservoir Gas.

GAS SPEC mol-%	Res Gas	Pipeline NG	Fuel Gas
CH4	7.50%	93.9 %	11.8 %
C2H6	2.50%	3.20%	2.54%
C3H8	1.50%	0.70%	1.46%
C4+	2.00%	0.30%	1.92%
N2	1.50%	0.90%	1.47%
CO2	85.0 %	1.00%	80.8 %
HHV (Btu/scf)	224	1033	265

Typically one would stabilize fuel input by mixing the RG with 5% NG. This yields a low-btu Fuel Gas (FG) with HHV of about 265 Btu/scf (5.75 MJ/kg/ LHV 5.25 MJ/kg) (see Table 1). The GG has been tested to combust as low as 250 Btu/scf.

Fig. 3 shows the CES Cycle that is compatible with the configuration used for testing the J79 turbine at KPP (cycle CES-J79) firing natural gas as fuel. Inlet conditions, mass flow, intermediate and exit pressure are taken from [2], all other values have been calculated using the thermodynamic simulation software IPSEpro [24]. The thermodynamic simulation was performed to calculate representative efficiencies, volume flows and working fluid composition for the turbine so that differences to utilization in a Graz Cycle operated with the modified FG could be estimated.

Isentropic efficiencies of the Main and Exhaust turbines were evaluated in order to obtain a turbine shaft power of 32.4 and 10.8 MW respectively in accord with the power estimates of CES [2]. Recycled water is used for cooling and to ensure that the Main turbine inlet temperature (TIT) is 760°C and thus compatible with the un-cooled turbine.

Results are presented in Table 2 where the first column summarizes results for the CES Cycle of Fig. 2. The working fluid leaving the GG is composed of 84.5% steam, 14.5% CO_2 and 1.0% O_2. The inlet flow to the Main turbine is 23.4 m^3/s, while the outlet flow is 95 m^3/s and the Exhaust turbine outlet flow is 163 m^3/s at 1.1 bar.

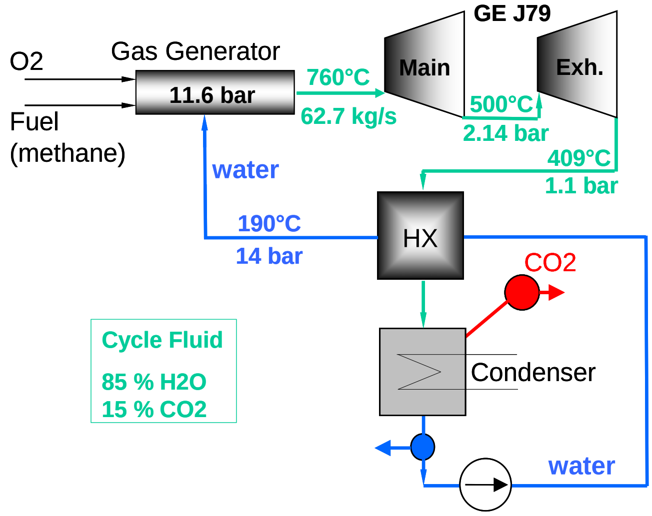

Fig. 3: Flow configuration of J79 testing at KPP (CES-J79)

Table 2: Operating Parameters for the Oxyfuel J79 Turbine when using Natural Gas (NG) in the CES configuration and Fuel Gas (FG) (CO_2 contaminated Reservoir Gas (RG) mixed with 5% NG) in the GC configuration

Operating Parameters	i) CES-J79: NG	ii) GC-B-J79: FG	iii) GC-AD-J79: FG
Mass flow [kg/s]	**62.7**	**77.9**	**74.1**
Inlet pressure [bar]	**11.6**	**11.6**	**12.3**
Inlet temperature [°C]	**760**	**760**	**815**
Speed [rpm]	**7460**	**7460**	**6890**
Main Turbine [MW]	32.4	32.6	33.0
Exhaust Turbine [MW]	10.8	10.9	10.7
% Composition (H_2O/CO_2)	85/15	54/45	51/48
Inlet Volume [m^3/s]	23.4	23.3	21.5
Intermediary Vol. [m^3/s]	95.0	96.1	93.9
Exit Vol. [m^3/s]	163	166	162.7
Spec. Enthalpy drop [kJ/kg]	710	576	607
Main Turbine Exit Temp. [°C]	500	514	552
Exh. Turbine Exit Temp. [°C]	409	426	463

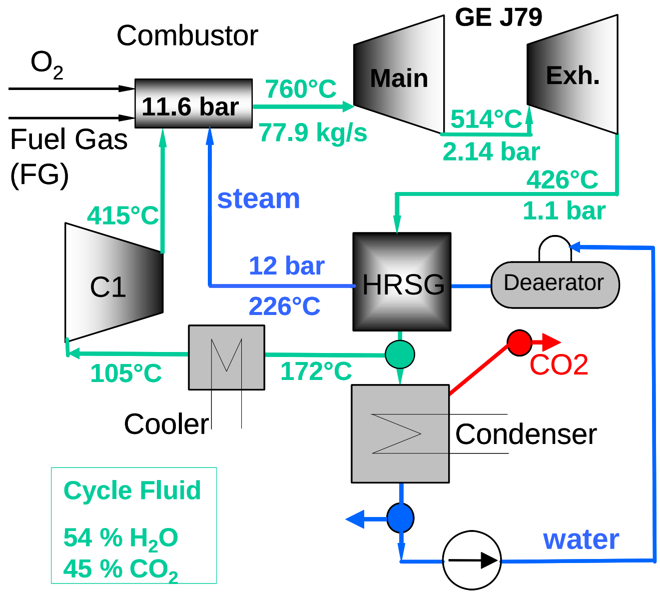

Fig. 4: Flow Configuration for a Near-Term Basic Graz Cycle Plant with Fuel Gas Firing (GC-B-J79)

The respective turbine exit temperatures of 500°C and 409°C are well below the original design data specifications for air-breathing operations at 635°C and 465°C.

Fig. 4 shows the principle flow scheme of a first generation Graz Cycle plant GC-B-J79 operating with the J79 Main and Exhaust turbines similar to the conditions tested by CES. The flow parameters shown are with FG as fuel input.

Comparing with the higher efficiency Graz Cycle in Fig. 1, the steam generated in the HRSG is now fed directly to the combustor, while the HP turbine is omitted for simplicity reasons. The steam superheating is relatively low with steam at 226°C. Calculations showed that a higher temperature of 401°C (25°C temperature difference in the superheater) has negligible effect on the efficiency, and the lower steam temperature is suggested to reduce thermal loading. The condenser operates at near atmospheric conditions, while final expansion into an LP turbine is also omitted. So the segregated CO_2 is provided at atmospheric conditions for further use. Because of the lower cycle peak pressure (11.6 bar compared to 40 bar), only the C1 compressor is needed for working fluid recompression. The recycled fluid is cooled to 105°C before C1 so as to have lower inlet and exit temperatures thus reducing compression power requirements and increasing overall efficiency.

In order to prevent the accumulation of oxygen and CO_2 in the boiler water, and thus the risk of increased corrosion, a deaerator is arranged prior to the HRSG. It uses small amounts of superheated steam for degassing.

The three columns in Table 2 compare the main parameters of the J79 turbine for three different cycle configurations, i) is the cycle shown in Fig. 3 (CES-J79), ii) is a Basic Graz Cycle as shown in Fig. 4 using the Fuel Gas shown in column 3 of Table 1 (GC-B-J79), iii) is similar to cycle ii) but with optimized turbine flow conditions (see below) (GC-AD-J79).

The volume flows were kept the same as in the CES Cycle, but we observe a significant reduction in specific enthalpy (from 710 to 576 kJ/kg) between the two cycles CES-79 and GC-B-J79. This is due to the lower heat capacity of the working fluid with the changed composition of the FG containing now 54% steam and 45% CO_2. This will lead to a poor matching of the velocity vectors with the blade angles.

In the third column we therefore present an optimized cycle (GC-AD-J79) with slightly higher TIT at 815°C (thereby requiring some blade cooling), but with reduced turbine speed (92.4% nominal). At this speed volume flow and enthalpy drop match according to the nominal conditions and this leads to a specific enthalpy drop of 607 kJ/kg at an inlet volume flow of 21.5 m^3/s.

For the two Graz configurations the working fluid contains almost equal volumes of CO_2 and steam. This increases the total CO_2 produced by the power plant and, as we show later, significantly improves overall project economics.

The CES-J79 Cycle requires 170 MW_{th} (HHV) of heat input to produce 43.2 MW shaft power, thereby indicating a thermal (Carnot) efficiency of 25.4%. Table 3 presents the power balance for the CG-B-J79 cycle (column 1) where we observe that required thermal input is now only 65 MW_{th} to produce 43.5 MW from the turbine, although net shaft power is 20.0 MW after deduction of 23.5 MW for the recycle compressor (mechanical efficiency of 97 % is assumed).

Table 3: Power Balance of a Near-Term Graz Cycle Plant with Basic(B) and Advanced (AD) J79 Turbine Operating Conditions.

Fuel Gas is as shown in Table 1 column 3	**GC-B-J79 (760°C/ 11.6 bar)**	**GC-AD-J79 (815°C/ 12.3 bar)**
Total Heat Input (HHV) [MW]	65.3	62.6
Total Turbine Power [MW]	43.5	43.6
Compressor Power [MW]	23.5	22.5
Net shaft power [MW]	20.0	21.1
Thermal cycle efficiency [%] (1)	**30.6**	**33.7**
Auxiliary Losses [MW] (2)	0.6	0.6
Electrical cycle efficiency [%] (3)	**29.2**	**32.2**
Power for ASU [MW] (4)	5.9	5.7
Net Power Output [MW_e]	**13.2**	**14.5**
Net Efficiency (HHV) [%](5)	**20.2**	**23.2**
Net Efficiency (LHV) [%](5)	22.2	25.4

Notes: *(1) Defined as shaft output divided by heat input (3% mechanical losses are considered).*
(2) Includes power of pumps, fans for cooling, etc.
(3) After deduction of auxiliary losses and generator / transformer losses of 1.5%.
(4) Deduct utility power to "over-the-fence" ASU.
(5) All the CO_2 and N_2 compression is done by the oilfield operator as part of the field centralized gas processing plant and is not considered here.

Thermal shaft efficiency is 30.6% and higher than the CES-J79 configuration as would be expected for a regenerative cycle. After consideration of generator / transformer losses of 1.5 % and the deduction of auxiliary losses, the electrical efficiency is reduced to 29.2%.

Oxygen with a surplus of 5% is provided for the combustion process. The power effort for oxygen generation in an ASU and compression to 14 bar is estimated to 1,225 kJ/kg. This reduces the net power output to 13.2 MW and the net efficiency to 20.2% (22.2 % based on LHV).

In Table 3 we also summarize performance of the more advanced cycle (GC-AD-J79) having 815°C TIT and 92.4 % nominal speed in order to optimize the flow velocity vectors. We observe that the net efficiency increases by 3.0 %-points, delivering a net power output of 14.5 MW.

Equipment for the First Generation Graz Cycle Plant

The testing currently being undertaken by CES on the re-engineered GE J79 turbine with the LM 1500 exhaust turbine and reheat combustion cans, will facilitate deployment of a near-term Graz Cycle plant to the extent that only the recycle compressor needs additional considerations.

In Fig. 5 from [12] the design for the C1 compressor of a 400 MW_e Graz Cycle plant was presented. The compressor runs at 8500 rpm compressing the working fluid from 1 to 10.4 bar. The drum rotor carries 6 axial and one radial stage. The blade length of the first stage is 367 mm and the last axial stage 112 mm with an inlet volume flow of 309 m^3/s.

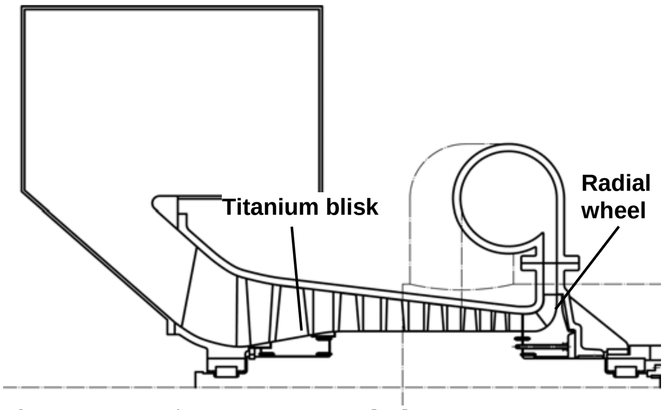

Fig. 5: Graz Cycle Compressor C1 [12]

The enthalpy rise of the C1 compressor is 590 kJ/kg, which is compatible with the Basic Graz Cycle compressor enthalpy rise of 474 kJ/kg when the compressor runs at the speed of the J79 turbine of 7,460 rpm. The volume flow is about 18% that of the original compressor.

If the same compressor design is used, then dimensions are as shown in Table 4. If the turbine drives the compressor at the nominal design speed of 7,460 rpm, then a slightly larger mean diameter (+ 1.2 %) is necessary to compensate for the reduced rotational speed. The blade lengths reduce to 73 and 24 mm respectively, with a mean diameter of about 1.055 m at the inlet. These dimensions appear to be feasible and lower the centrifugal load compared to the original Graz Cycle design.

Table 4: Main Dimensions for the Recycle Compressor of a First Generation Graz Cycle Demonstration Plant.

		Inlet	Outlet
Volume flow	m^3/s	55	9
Mean diameter (D)	m	1.055	0.897
Inner diameter	m	0.982	0.873
Blade length (l)	mm	73	24
Inlet D_{in}/D_{out}	–	0.87	0.95
Enthalpy rise	kJ/kg	474	
Stage number	–	6+R	
Speed	rpm	7460	

ECONOMICS

Any near-term zero emission power plant will need to be deployed commercially in a niche market preferably providing revenue from multiple product streams. A unique feature of the CES GG is that it enables stoichiometric combustion of almost any composition of gaseous hydrocarbon fuel that has a low heat value and high CO_2 content, thereby also enabling use of fuels that other conventional power generation technologies cannot utilize.

We here present a preliminary economic analysis for the Basic Graz Cycle[1] in conjunction with enhanced oil recovery (EOR) as is applicable for deployment of standalone power and injectant gas to oilfield operators in the Permian Basin or onshore along the Gulf Coast.

The field operators in these regions have reservoirs that have been extensively depleted through primary production and secondary water-flooding over the past half century. Tertiary recovery using miscible[2] gas injection is a well understood production method to extract the final incremental oil before abandonment. Depending on reservoir characteristic, depth and temperature, the type of miscible gas used will be determined by availability of supply and total cost for production of each incremental barrel of oil.

In the North Sea tertiary recovery has been extensively practiced since the 1980's using available hydrocarbon gas from adjacent fields. This was primarily because it was not possible offshore to get the gas to market at a pace commensurate with the demand for oil production.

In the Permian Basin there emerged during the 1970's an opportunity for using CO_2 as injectant gas because; i) an availability of supply (initially from natural gas processing plants) and, ii) a favorable depth / characteristics for the reservoirs (typically lighter oils above 28 API and at depths from 3,000

[1] Our economic analysis for this paper showed that there were only marginal differences in economic performance between the three cycles we have considered.

[2] Miscibility is the ability of the injectant gas and the reservoir oil to effectively combine and form a single-phase interface that releases more of the oil that would otherwise be trapped in the pore space. The Minimum Miscibility Pressure (MMP) is a key indicator regarding suitability of the injectant gas for use with EOR.

down to 7,000 feet) provided ideal conditions for miscibility with CO_2. Furthermore during the 1980's an extensive pipeline infrastructure was developed that currently secures the supply of around 34 million tonne per year (mt/yr) of predominantly naturally occurring CO_2 into the Permian Basin [25].

Finally in deeper fields below 10,000 feet it is also recognized that nitrogen can have similar miscibility characteristic as carbon dioxide [26, 27]. Although nitrogen is currently not extensively used for EOR – because of a preference for CO_2 – this remains a question of relative price and availability, because the field operator is motivated to acquire as cheap an injectant gas as is compatible with the reservoir characteristics.

The situation for many operators is that they have declining oil production and would like to expand their existing secondary or tertiary flood operations, but this requires a significant capital investment to upgrade the field, combined with availability of power for additional compression requirements and supply of suitable injectant gas. The concept of a standalone power plant that can provide both of these is therefore recognized as a solution that also enables the operator to freely develop the field in accord with the needs of the reservoir oil production – and in regions where there is currently no pipeline infrastructure. Furthermore we observe that the size of plant capital investments, available power and volume of injectant gas produced, is also compatible with the typical scale of such oil field operations.

In the following analysis we have assumed deployment of a plant where there exists opportunity for sale of both CO_2 and N_2 as injectant gas into the same or two adjacent oil zones at different depths. Such locations exist in both SE New Mexico and along the Gulf Coast. We have used costs that are appropriate for the Gulf Coast area as of 3Q-2010.

Power Plant Capital Investment and Development Costs

The overall power plant capital investment is estimated to be \$41.5 million based on a preliminary assessment at pre-FEED level that draws on our historic experience from earlier studies for the same region.

The power plant is basically a single cycle power train with a recycle compressor, combustor, turbine expander and electric generator. This is combined with a HRSG in combination with the condenser and balance-of-plant equipment. Overall we have specific cost of \$3,147 /kW installed (excluding ASU).

Plant oxygen requirement is 414 tonne per day (tpd) while the energy penalty for delivery at 14 bar is 0.34 kWh/kg. The commercial arrangement for supply of oxygen can differ, e.g. fully integrated, "over-the-fence" or standalone. We here assume that the ASU is fully integrated requiring 5.9 MW_e of auxiliary power (plus utilities). The incremental capital investment cost is \$44.0 m representing an effective cost for capital of \$46.5 per tonne of O_2 produced. These values are based on dialogue we have had with several industrial gas providers in the Gulf Coast region. [Our general perception is that the capital cost for oxygen will vary from roughly \$30 - \$40 per tonne as the size of ASU scales down from 3,000 to 800 tonne per day unit plant size.]

Table 5: Estimation of total project costs for a Basic Graz Cycle Plant (GC-B-J79)

Graz Cycle - Basic Reservoir Gas with 5% NG	
INVESTMENT SUMMARY	**$mill.**
Power Plant Capital Cost	41.54
Air Separation Unit Cost	44.05
Total CO2 Plant Capital Cost	**85.6** $mill.
Development Costs	2.77
Pre-FEED + Contingency	1.08
Interest During Construction	4.49
Govt. Grant Support	0.00
TOTAL PROJECT COST	93.9 $mill.

Compression and recycling of produced gas is a significant part of oilfield operations requiring major investment in centralized facilities by the field operator. Integration of power plant compression with field operations can provide a good opportunity for overall process optimization – however this will also be very field dependent. For a generic study, as presented here, it is difficult to include such details. Instead we choose a base case scenario where we exclude power plant compression equipment from our capital investments, choosing instead to sell injectant gas at a significantly reduced price that conservatively recognizes an incremental cost of \$0.60 /Mcf (equivalent to \$11.4 per metric tonne) to the field operator for compression. Total project investment (including development cost and interest during construction) is therefore estimated in the order of \$94 million as summarized in Table 5 above. (Note: Contingency here relates to development costs and not total investment cost.)

Table 6: Summary of Plant Output

PLANT CAPACITY SUMMARY		
Power Output		13.2 MWe
Efficiency (HHV)	15,405 Btu/kWh	20.2 %HHV
Availability		92.0 %
Annual Production Hours		8,065 hrs
Annual Electrical Output		0.106 TWh
Annual CO2 Output		0.394 mt/yr
Annual N2 Output		0.454 mt/yr
Annual H2O Output		0.068 mt/yr

A summary of plant capacities is shown in Table 6 above. Overall annual power production is 106 GWh while 0.39 mt/yr of CO_2 is produced. Furthermore the ASU produces an additional 0.45 mt/yr of N_2. The combustion process yields 68,000 t/yr (~54,000 US gall/day) of industrial quality de-ionized water. We do not for time being include the sale of any steam nor produced water although for some locations these will have an intrinsic value.

Financial and Revenue Assumptions

Our basic financial assumptions are summarized in Table 7. These are comparable with typical project investment criteria. Although we do not include any direct government support, we do assume the possibility for 70% debt financing which would most probably require some form of loan guarantee based on federal or state incentives to help promote zero-emission technologies. Several states and the DOE already have such mechanisms in place, but the project would need to qualify on the basis of merit.

Table 7: Financial Assumptions

FINANCIAL ASSUMPTIONS	
Interest Rate	7.5%
Equity Percent	30%
Debt Term	12 yrs
Project Duration	15 yrs
Construction Duration	1.5 yrs
Business Tax Rate	34.0%
Discount Rate	15.0%

Specified project duration of 15 years inherently assumes that adjacent fields will help extend power plant operation beyond the 5 to 8 years that is typical for tertiary oil recovery floods. Historically this is how regions have expanded in the past. Once "in-situ" it is plausible that the plant may both expand in size and have an extended economic lifetime.

The revenue assumptions presented in Table 8 are representative of prevailing market conditions; but subject to more detailed commercial discussions. The fuel cost is based on a market price of \$4.50 /MMBtu (HHV) for pipeline quality natural gas (NG). However, we purchase Reservoir Gas directly from the field operator (at a 25% discount on market price in recognition that it is delivered prior to gas processing) and then mix this with 5 mol-% of NG to maintain a satisfactory btu-value for combustion in the Gas Generator. The resulting Fuel Gas cost is \$3.59 /MMBtu (HHV).

Table 8: Revenue Assumptions in US\$

REVENUE ASSUMPTIONS		
Electricity Tarrif	60.0	\$/MWh
Fuel Gas Cost	3.59	\$/MMBtu
CO2 Sales Price	19.01	\$/tonne
CO2 Credit Price	0.00	\$/tonne
Oxygen Cost	0.00	\$/tonne
N2 Sales Price	14.93	\$/tonne
H2O Sales Price	0.00	\$/tonne

We assume a power price of \$60 /MWh. Regional cost of electricity varies significantly throughout the Gulf Coast region due to a complex historical mix of regulated and free-market utilities. Although in some regions, like West Texas, there is an over capacity of wind power that has driven power prices down, the above quoted price is most probably lower than any long-term contract the field operators may be able to secure from their local utility. A typical market spread would be from \$55 to \$95 /MWh – although the longer-term contracts (e.g. 10+ years) would be towards the higher end of this spread.

The net produced power (13.2 MW_e) is comparable with that used by the field operator for compression, gas processing and utilities. This can simplify project deployment, reduces additional cost of grid connections and removes the need for the field to be dependent upon access to a local utility power infrastructure.

The CO_2 sales price is here proposed to be \$1.00 /Mcf (equivalent to \$19.0 per metric tonne) uncompressed. This arrangement of outsourcing compression simplifies overall project investment decisions and reduces total power plant capital investment cost. A comparable price for CO_2 when delivered fully compressed would be \$1.60 /Mcf (equivalent to \$30.4 per metric tonne). Our experience suggests that this would be commensurate with oil at \$80 /bbl. In most commercial contracts the sales price of CO_2 is indexed to the price of oil so that both the power plant and field operator may benefit if oil prices rise.

The sale of nitrogen (N_2) is not standard practice, but does occur [26] and most operators are well aware of its favorable characteristics for gravity drainage, pressure maintenance or miscible/ immiscible displacement. In one of the deeper floods that we have evaluated the combined use of CO_2 and N_2 (mixed) maintained MMP and effectively doubled the volume of injectant gas available. This accelerated incremental oil production while requiring fewer production wells so that payback period for field investment was shorter. The analysis suggested that N_2 delivered at \$0.50 /Mcf (equivalent to \$14.9 per tonne) uncompressed had satisfactory commercial value for incremental recovery when oil remained above \$80 /bbl.

Table 9: Evaluation of Net Operating Income

ANNUAL RESULTS	\$mill.
Electricity Sales	6.39
CO2 Sales	7.48
Emission Credits	0.00
N2 Sales	6.78
Steam Sales	0.00
H2O Sales	0.00
Gross Revenues	**20.65 \$mill.**
Oxygen Costs	0.00
Fuel Gas Costs	6.46
O&M Costs	0.83
Operating Costs	**7.29**
Net Operating Income	**13.36 \$mill.**

Based on the aforementioned assumptions the power plant annual results are shown in Table 9 above. One encouraging feature is that gross revenue is evenly distributed between sales of three distinctly different product streams; electricity (\$6.4 m), CO_2 (\$7.5 m) and N_2 (\$6.8 m). Furthermore, albeit of lesser significance, we do not here assume income from sales of produced steam /water nor emission credits (e.g. NO_x, sulfur or, if recognized, CO_2 for zero emission power).

For expenditure we have chosen to leave the ASU as a capital investment so that oxygen is not an operating cost – while fuel (\$6.5 m) and O&M (\$0.8 m) are. Overall the net annual operating income is \$13.4 m suggesting a payback period of 7 years. The unlevered (EBITDA) return on invested capital (IRR) is 14.5% while our estimate for the levered return is 21.8%. These results are summarized in Table 10 below and are compatible with financial market expectations.

Table 10: Financial Summary

Graz Cycle - Basic Reservoir Gas with 5% NG		
FINANCIAL SUMMARY		**Project Analysis**
Startup Date: 2014 - Q2	Project Life: 15 yrs	
Total Project Investment		93.9 $mill.
Net Annual Operating Income		13.4 $mill.
Investment Return	Unlevered	14.5% IRR
	Levered	21.8% IRR

Although these results are based on very preliminary cost estimates and some of the assumptions regarding capital investment, operating costs and plant performance will differ as the project moves into more detailed pre- and full-FEED studies. The important point to note is that the project has multiple revenue streams and sensitivity to cost escalation in one area may be suitable compensated by adjusting other assumptions thereby balancing economic risk.

This confirms that there may be opportunities for near-term implementation of oxyfuel CCS technology using the hardware that has already been demonstrated by Clean Energy Systems. Furthermore there are medium-term opportunities for process optimization using the regenerative Graz cycle that can improve performance as next generation oxyfuel turbines [3] and recycle compressors become commercially available.

Another observation from the present study is that for the near-term there is no unique optimal process design. This can in fact vary depending upon regional costs, demand for power, and need for injectant gas.

For brevity we do not include results for the Advanced Graz Cycle (GC-AD-J79) beyond stating that IRR is practically identical despite the slightly higher electrical efficiency. This simply highlights the important economic feature of such near-term plants when electricity sales are not the primary source of revenue. In some cases NPV increases as efficiency is reduced because there is more income from the sale of injectant gases compared with power sales.

CONCLUSIONS

In order to achieve near-term development of oxyfuel technology for CCS it is suggested to utilize existing hardware already proven by Clean Energy Systems. We have configured a basic (GC-B-J79) and an advanced (GC-AD-J79) "near-term" Graz Cycle plant using data for the GE J79 turbine expander that is currently being deployed by CES at their Kimberlina Power Plant. The fuel source is based on low-btu reservoir gas typical for CO2-floods following breakthrough of CO_2 into the production wells.

When maintaining the main turbine parameters (inlet temperature and pressure, volume flows and power output) as suggested by CES, a net cycle efficiency of 20.2% (HHV) and a net power output of 13.2 MW can be expected. However a lower enthalpy drop leads to poorer flow efficiencies. For this reason an advanced cycle is also proposed having a higher TIT that necessitates some cooling but ensures closer compliance with the GE J79 data leading to a net efficiency of 23.2% (HHV) and 14.5 MW power output.

We have performed an economic evaluation for early introduction of such zero-emission power generation technology based on the basic cycle GC-B-J79. The approach and methodology is significantly different from that which is usually adopted for such project assessment in that we focus on revenue streams rather than thermodynamic efficiency. We have identified a unique market location (EOR) and deploy a smaller size power generation capability (65 MW_{th}) where there are oilfield operators that specifically need multiple products (power, CO_2 and N_2) at costs and volumes that are comparable with the proposed power plant economics and configuration.

We acknowledge that some of the assumptions regarding capital investment, operating costs and plant performance will differ as the project moves into more detailed pre- and full-FEED studies. However the economic analysis suggests that with multiple product streams there is greater opportunity to absorb economic risk, thereby making the project more robust.

This strategy for early commercialization can enable proof-of-concept demonstration of oxyfuel CCS hardware that may subsequently be improved through commercial deployment in a broader competitive market while providing acceptable return on invested capital already with the near-term projects.

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