A qualitative reliability and operability analysis of an integrated reforming combined cycle plant with CO_2 capture

Lars Olof Nord [a,*], Rahul Anantharaman [a], Marvin Rausand [b], Olav Bolland [a],

[a] *Department of Energy and Process Engineering, the Norwegian University of Science and Technology, NO-7491 Trondheim, Norway*

[b] *Department of Production and Quality Engineering, the Norwegian University of Science and Technology, NO-7491 Trondheim, Norway*

Abstract

Most of the current CO_2 capture technologies are associated with large energy penalties that reduce their economic viability. Efficiency has therefore become the most important issue when designing and selecting power plants with CO_2 capture. Other aspects, like reliability and operability, have been given less importance, if any at all, in the literature.

This article deals with qualitative reliability and operability analyses of an integrated reforming combined cycle (IRCC) concept. The plant reforms natural gas into a syngas, the carbon is separated out as CO_2 after a water-gas shift section, and the hydrogen-rich fuel is used for a gas turbine. The qualitative reliability analysis in the article consists of a functional analysis followed by a failure mode, effects, and criticality analysis (FMECA). The operability analysis introduces the comparative complexity indicator (CCI) concept.

Functional analysis and FMECA are important steps in a system reliability analysis, as they can serve as a platform and basis for further analysis. Also, the results from the FMECA can be interesting for determining how the failures propagate through the system and their failure effects on the operation of the process. The CCI is a helpful tool in choosing the level of integration and to investigate whether or not to include a certain process feature. Incorporating the analytical approach presented in the article during the design stage of a plant can be advantageous for the overall plant performance.

Key words: CO_2 capture, Pre-combustion, Reliability, FMECA, Operability, Control degrees of freedom

1 Introduction

Capturing the CO_2 from fossil fueled power plants can be part of an overall mitigation strategy to reduce the rise in atmospheric temperature. There are several approaches for capturing CO_2 from power generation. One is pre-combustion capture, where the fossil fuel is decarbonized to produce a syngas. The carbon, as CO_2, is separated out before the combustion takes place. For coal, one could implement pre-combustion CO_2 capture in the integrated gasification combined cycle (IGCC). IGCC plants exist, but none of them employs CO_2 capture. There are, however, a number of IGCC plants with CO_2 capture in the planning phase (Scottish Centre for Carbon Storage, 2009). For natural gas pre-combustion capture, the integrated reforming combined cycle (IRCC) that reforms natural gas into a hydrogen-rich fuel (Andersen et al., 2000),

* Corresponding author.
Email address: lars.nord@ntnu.no (Lars Olof Nord).

could be attractive. This technology has yet to be implemented in practice. The gas turbines in an IGCC or IRCC plant would fire a hydrogen-rich fuel.

The IGCC cycle has been studied extensively in terms of thermodynamical analyses to arrive at a cycle efficiency, and also economical analyses (e.g., Bohm et al., 2007; Descamps et al., 2008). To a lesser extent, aspects such as reliability, availability, and maintainability (RAM) have been studied for the IGCC cycle (Higman et al., 2006). Limited literature is available on reliability analyses of pre-combustion natural gas cycles. However, as large-scale implementation of CO_2 capture from power plants draws nearer, there will likely be more focus on areas such as RAM and operability.

A main issue related to pre-combustion techniques is that the plant becomes more complex with the significant integration between the power cycle and the gasification (for the IGCC case) or reforming (for the IRCC case) process. In addition, some of the technology is less mature than for a pulverized coal plant or for a natural gas combined cycle (NGCC) plant. The gas turbine (GT) technology is, for example, much more mature for natural gas firing than for firing a hydrogen-rich fuel. Chiesa et al. (2005) address issues related to using hydrogen as fuel for GTs. Also, a GT designed for an IGCC or IRCC plant typically needs to be more fuel flexible, which requires special attention to the burner design (Bonzani and Gobbo, 2007) and the control system (Shilling and Jones, 2003). The less-mature technology and the integration present in IGCC plants are some of the reasons for the initially low availability of such plants (Higman et al., 2006; Beér, 2007). However, the availability of IGCC plants have steadily been improving since first introduced to the market.

In the RAM field, more literature is found if one looks for analyses of power plants in general and do not limit oneself to CO_2 capture plants. Examples of RAM analyses in the literature include Eti et al. (2007) and Åström et al. (2007). Another related area is reliability analysis of chemical systems. A thorough literature review related to chemical system reliability is given by Dhillon and Rayapati (1988). An international standard for production assurance and reliability management has recently been published (ISO 20815, 2008). In this standard, the term "production assurance" is used with the same meaning as operability in this article.

Failure modes, effects, and criticality analysis (FMECA) is a widely used qualitative method for reliability analysis (e.g., see Rausand and Høyland, 2004; IEC 60812, 2006). Teng and Ho (1996) discuss the use of FMECA for product design and process control. Teoh and Case (2004) describe, among other topics, the connection between system functional diagrams and FMECA. FMECA can be used to identify critical areas during the design stage of the system. When the criticality of failures is not investigated, the FMECA is sometimes called failure mode and effect analysis (FMEA).

The complexity and efficiency of a process plant normally increase with the degree of integration. While the increase in efficiency is a desired result, the increased complexity can give rise to operability and risk issues (e.g., see Perrow, 1999). The degree of integration in a process plant should therefore be determined based on a trade-off between efficiency and complexity. Operability is dependent on plant design and efforts have been made to incorporate process operability and control at an early stage of the design process (Barton et al., 1991; Blanco and Bandoni, 2003). The procedures presented in literature are computationally intensive and provide a level of rigor not required for the purposes of this work. A new index called the comparative complexity indicator (CCI) presented here is a parameter for comparing complexity of processes that provides a simple guide to the engineer on the extent of integration. As the name suggests, this indicator is useful only when comparing two processes and the absolute value of the indicator for a single process has no significance by itself.

The main objectives of this article are: (i) To illustrate and discuss the use of qualitative reliability and operability analyses in the field of CO_2 capture as a first step in developing a methodology for the design of a power plant with pre-combustion CO_2 capture, and (ii) to introduce a new concept, the comparative complexity indicator, as a tool for choosing the level of process integration and to gauge the complexity of a CO_2 capture plant.

The remainder of the article is divided into the following sections: Section 2 describes the process with functional descriptions of the building blocks. Section 3 describes the details of the methodologies used in the article. The results are shown and analyzed in Section 4, and concluding remarks are given in Section 5.

2 Functional description of process

A functional diagram of the cycle studied is shown in Fig. 1. The purpose of the plant is to generate fossil fueled power with low CO_2 emissions. The process has a defined system boundary as shown in Fig. 1. Inputs to the system include natural gas, ambient air, make-up water, and cooling water. Outputs across the system boundary include compressed CO_2, water that has been separated out, cooling water, exhaust from the heat recovery steam generator (HRSG)

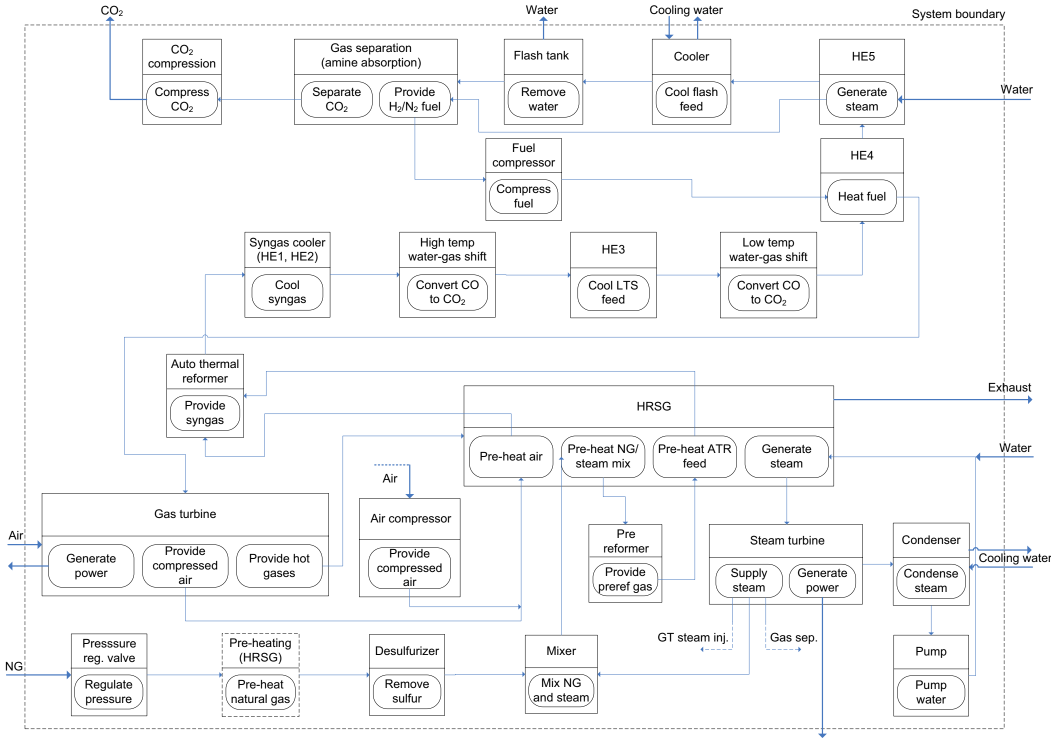

Fig. 1. Functional block diagram of an integrated reforming combined cycle plant.

that originated in the gas turbine exhaust, as well as power generated in the generator connected to the power train. In Fig. 1 the generator is incorporated into the gas turbine and steam turbine blocks.

In addition to the functional diagram in Fig. 1, a process flow sheet of the system is shown in Fig. 2. This representation of the system gives further insight and will prove helpful in the operability analysis.

2.1 Description of system inputs and outputs

The system inputs and outputs crossing the system boundary in Fig. 1 are described below.

Natural gas

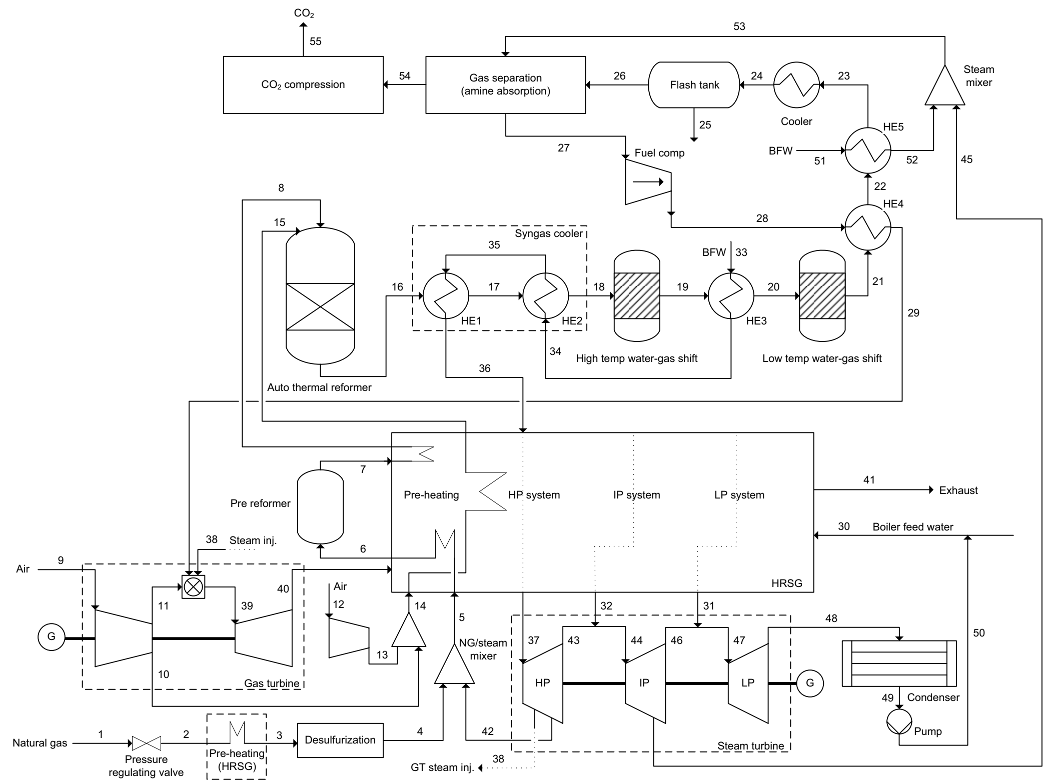

Fig. 2. IRCC process flow sheet.

The supplied natural gas has an assumed pressure of 3.1 MPa and a temperature of 16°C with a mass flow of 19 kg/s. The stream composition is given in Table 1.

Ambient air

The ambient air is assumed at 0.1013 MPa and 15°C with 60% relative humidity and a total mass flow (air to gas turbine and to air compressor) of 648 kg/s. The air composition is given in Table 2.

Exhaust

The exhaust originating from the gas turbine exhaust, passing through the HRSG, and exiting through the stack has a temperature of about 90°C and a pressure of 0.1013 MPa with a mass flow of 650 kg/s.

Table 1

Natural gas composition in model.

Component name	Chemical formula	Unit	Value
Methane	CH_4	vol%	79.84
Ethane	C_2H_6	vol%	9.69
Propane	C_3H_8	vol%	4.45
i-Butane	C_4H_{10}	vol%	0.73
n-Butane	C_4H_{10}	vol%	1.23
i-Pentane	C_5H_{12}	vol%	0.21
n-Pentane	C_5H_{12}	vol%	0.20
Hexane	C_6H_{14}	vol%	0.21
Carbon dioxide	CO_2	vol%	2.92
Nitrogen	N_2	vol%	0.51
Hydrogen sulfide	H_2S	ppmvd	5

Table 2

Ambient air composition in model.

Component name	Chemical formula	Unit	Value
Oxygen	O_2	vol%	20.74
Nitrogen	N_2	vol%	77.30
Argon	Ar	vol%	0.92
Carbon dioxide	CO_2	vol%	0.03
Water	H_2O	vol%	1.01

Water

Make-up water has an inlet temperature of 49°C and a pressure of 0.19 MPa.

Cooling water

The incoming cooling water for the condenser and cooler has an assumed temperature of 15°C with a temperature increase in the heat exchangers of 10 K. Direct cooling by sea water is assumed.

CO_2

The compressed CO_2 stream has above 99 vol% CO_2 and a pressure of 11.0 MPa with a temperature of about 41°C. The mass flow is 47 kg/s.

Power

The net power output from the plant is approximately 362 MW.

2.2 Functionality and description of equipment

The functional blocks in Fig. 1 are described below.

Pressure regulating valve

Function: To reduce the natural gas pressure from a delivery pressure of 3.1 MPa to approximately 1.9 MPa.

The pressure is set in order to match the compressed air pressure at the entrance of the auto thermal reformer (ATR).

Desulfurizer

Function: To reduce the H_2S content in the natural gas to 2 ppmvd.

Sulfur removal is necessary to protect the catalysts in the reforming and water-gas shift reactors. Because of the low sulfur content in the selected natural gas composition, 5 ppmvd H_2S, a ZnO desulfurizer is selected. The sulfur is removed by flowing of the natural gas through a bed of ZnO granules according to the reaction

$$H_2S + ZnO \rightarrow H_2O + ZnS \tag{1}$$

Mixer

Function: To mix the desulfurized natural gas with steam extracted from the steam turbine.

The steam to carbon ratio is set to 1.5 on a molar basis.

Gas turbine

Function: To generate power; to provide compressed air to the ATR; to provide hot flue gases to the HRSG.

The power cycle consists of a General Electric (GE) 9FA H_2–fired gas turbine (GT). The fuel fed to the GT combustor in principle consists of a mixture of H_2 and N_2. Because of the air-blown ATR, the water-gas shift reactors and the CO_2 capture processes, the fuel consists of approximately 50 vol% hydrogen. This enables use of available IGCC-type diffusion combustors (Todd and Battista, 2000; Shilling and Jones, 2003). The nitrogen acts as a fuel diluent. For further NO_x control, steam is injected into the flame. From the gas separation stage the fuel mix is passed on to the gas turbine via a fuel compressor. The GT turbine inlet temperature has been reduced because of the high steam content in the turbine. The hydrogen fuel together with the injected steam lead to an H_2O content entering the turbine of about 18.2 vol%. This leads to a higher heat transfer rate to the blades compared to a natural gas fired turbine. As a result, the metal temperature of the turbine blades is higher for the same turbine inlet temperature as in a conventional gas turbine. To obtain similar life of the turbine parts, the turbine inlet temperature reduction is necessary. Chiesa et al. (2005) report TIT decreases of 10-34 K for hydrogen combustion with nitrogen or steam diluent (VGV operation cases). As a model assumption, a TIT reduction of 30 K has been assumed for this work. In addition to running the GT on a hydrogen-rich fuel, the idea is to be able to operate on natural gas as a back-up fuel if the pre-combustion process is shut-down. This requires fuel flexibility for the combustor system (Shilling and Jones, 2003; Bonzani and Gobbo, 2007). In addition, start-up of the GT would be with natural gas fuel. It is also possible to run with a mixture of natural gas and the hydrogen-rich fuel. The gas turbine exhaust stream passes through the HRSG for pre-heating of the process streams and steam generation before emitted to the atmosphere through the stack.

Air compressor

Function: To provide compressed air to the ATR.

The external compressor is introduced in order to better utilize the operation of the gas turbine. If too much air is removed prior to the combustion chamber in the gas turbine, the effect on the performance and temperature profile can be negative.

Heat recovery steam generator

Function: To pre-heat the compressed air, the natural gas/steam mixture, and the pre-reformed ATR feed; to generate steam.

A triple pressure steam cycle was selected. The HRSG includes pre-heating for the various process streams. The pre-heated streams include the NG/steam feed to the pre-reformer, the ATR feed stream coming from the pre-reformer, and air extracted from the compressor discharge stream of the gas turbine combined with an additional compressor air stream before supplied to the ATR. The steam cycle is designed for pressure levels of approximately 8.3/1.0/0.3 MPa for the high, intermediate, and low pressure (HP/IP/LP) systems respectively. The pre-heating makes the HRSG design more complex and a lot of heat is removed from the gas stream at the hot part of the HRSG due to the high temperature requirements of some of the process streams. Note that the pre-heating is not entirely in the hot end of the HRSG but instead intermixed with the low, intermediate, and high-pressure sections. Equipment such as pumps for the different pressure levels, drums, valves, and so on, are not shown in the functional diagram.

Steam turbine

Function: To supply steam for the reforming process, the gas turbine, and the gas separation sub-system; to generate power.

The steam turbine (ST) has extractions for the GT steam injection, the reforming process steam, and for the reboiler in the amine absorption system.

Condenser

Function: To condense the steam.

After exiting the last low pressure turbine stage the steam is condensed in the condenser.

Pump

Function: To pump the water up to feed water pressure.

Pre reformer

Function: To convert the higher hydrocarbons into hydrogen and carbon monoxide.

Adiabatic pre-reforming of hydrocarbons is described by Vannby and Winter Madsen (1992). In the pre-reforming reactor the hydrocarbons higher than methane are converted to protect against coking in the primary reformer according to the reactions

$$C_xH_y + xH_2O_{(g)} \rightarrow xCO + (x + \frac{y}{2})H_2 \qquad -\Delta H^0_{298} < 0\,\mathrm{kJ/mol} \qquad (2)$$

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O_{(g)} \qquad -\Delta H^0_{298} = 206\,\mathrm{kJ/mol} \qquad (3)$$

Also, the exothermic water-gas shift reaction (4) converting the CO into CO_2 takes place in the pre-reforming reactor.

$$CO + H_2O_{(g)} \rightleftharpoons CO_2 + H_2 \qquad -\Delta H^0_{298} = 41\,\mathrm{kJ/mol} \qquad (4)$$

Auto thermal reformer

Function: To reform the stream from the pre-reformer into syngas.

Auto thermal reforming is described by Christensen and Primdahl (1994); Dybkjær (1995); Christensen et al. (1998). In the ATR the exothermic reaction (5) provide heat to the endothermic reaction (6).

$$CH_4 + \frac{1}{2}O_2 \rightarrow CO + 2H_2 \qquad -\Delta H^0_{298} = 36\,\mathrm{kJ/mol} \tag{5}$$

$$CH_4 + H_2O_{(g)} \rightleftharpoons CO + 3H_2 \qquad -\Delta H^0_{298} = -206\,\mathrm{kJ/mol} \tag{6}$$

As in the pre-reformer, the water-gas shift reaction (4) converts some of the CO into CO_2.

Syngas cooler

Function: To cool the syngas supplied by the ATR.

The syngas is cooled in the syngas cooler before entering the water-gas shift reactors. As a secondary function the hot stream supplied by the ATR is generating high-pressure steam in the syngas cooler. This steam is then supplied to the HP superheaters in the HRSG. The reason for using the syngas cooler as an evaporator rather than as a superheater is due to the risk of metal dusting. Metal dusting is further discussed in Section 3.1.2.

Water gas shift reactors

Function: To convert CO to CO_2.

The rest of the CO is converted to CO_2 according to reaction (4). The reasons behind dividing the water-gas shift reaction into a high temperature reactor and a low temperature one (HTS and LTS) are due to conversion rate and catalysts. To get a higher degree of conversion of the CO to CO_2, two reactors are favorable compared to a one-reactor setup. Also, there is a need for a more active catalyst at the lower region of the temperature range (Moulijn et al., 2007). It can therefore make sense to use a standard catalyst at the higher temperature range and then have a separate reactor with a more active catalyst for the low end temperature.

Heat exchanger 3

Function: To cool the stream from the HTS going to the LTS.

HE3 is also, together with the syngas cooler, producing high-pressure saturated steam to be added to the high-pressure superheater in the HRSG.

Heat exchanger 4

Function: To pre-heat the hydrogen-rich fuel for the gas turbine.

Heat exchanger 5

Function: To cool down the gas for the gas separation process.

Heat exchanger 5 (HE5) is also producing some of the steam necessary for the reboiler in the amine absorption process.

Cooler and flash tank

Function: To cool down the stream from HE5 and remove the water before the gas separation stage.

Gas separation (amine absorption)

Function: To separate out CO_2; to provide H_2-rich fuel.

In this model the gas separation stage is using the chemical absorbent activated MDEA (van Loo et al., 2007).

CO_2 compression

Function: To compress CO_2 up to delivery pressure.

The CO_2 is passed on to the compression section where the gas is compressed in the four compressor/intercooler stages and excess water is removed. To achieve the exit pressure of 11.0 MPa a pump is used at the end of the compression train.

3 Methodology

The plant model in Figs. 1 and 2 was analyzed from several angles, as illustrated in Fig. 3, in order to determine reliability and operability aspects of the plant design. As basis for the reliability analysis the process was first thermodynamically analyzed. This is important to be able to define the functional requirements and reveal the part load behavior of the plant. Some of the failure modes may affect the ability of the plant to operate at full load and the reliability of the plant will depend on the part loads. Even though the aim is to operate the plant at full load, it is also necessary to be able to operate the plant at part load. The thermodynamic analysis is not documented in this article, but indicates that part load operation down to 60% relative gas turbine load is possible. The relative load is here defined as the actual load of the GT divided by the full GT load at actual ambient conditions.

The reliability analysis was carried out as a functional analysis followed by an FMECA. The operability analysis is based on the new comparative complexity indicator (CCI). In the following sections, the reliability and operability analyses are described.

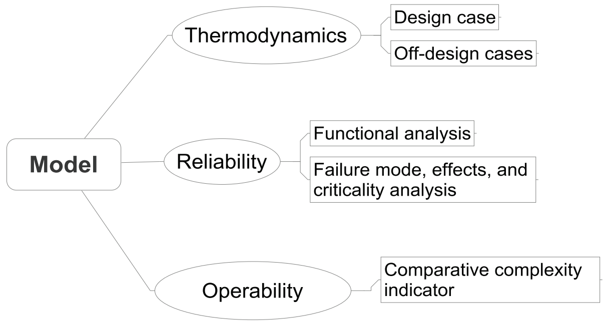

Fig. 3. Analytical approach to process model study.

3.1 Reliability analysis

The first step of the reliability analysis was a detailed functional analysis that was carried out to reveal and define all the required functions of the plant elements. For each function, the associated performance criteria were determined. A thorough understanding of all required functions and their associated performance criteria is a prerequisite for the FMECA.

The FMECA involves analyzing all the potential failure modes of the system elements (components and subsystems) and identify the causes and effects of these failure modes. The FMECA is also used to determine how failures may propagate through the system, and to reveal the failure effects on the operation of the plant. Another purpose of the FMECA was to identify the most critical components/integration points for further and more detailed analyses at later stages of the project.

3.1.1 Functional analysis

The functional analysis was carried out at the equipment level of the system, as shown in Fig. 4. The different subsystems and their equipments are listed in Table 3 together with the functional requirements (e.g., see Murthy et al.,

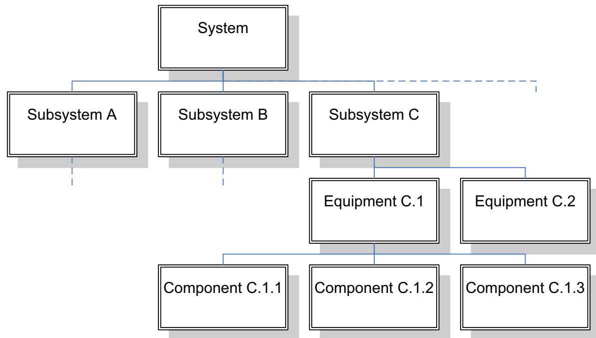

Fig. 4. Functional levels of a system.

2008). On system (plant) level the functional requirements are: Plant power output $\geq$ 300 MW (ISO); CO_2 capture rate $\geq$ 90%. The CO_2 capture rate is defined as the fraction of the formed CO_2 that is captured. The functional analysis that is documented in this article only includes the essential functions, meaning that auxiliary functions, protective functions, and so forth, are not covered.

3.1.2 FMECA

The FMECA approach that was selected for this project is illustrated in Fig. 5. In this approach, a risk, or criticality, number is assigned to each and every failure mode as a risk priority number (RPN). The RPN of a failure mode is calculated based on an evaluation of the factors: detection, failure rate, and severity, of a failure mode. Each of these three factors are typically assigned numbers ranging from 1 to 10. There are several approaches for assigning these numbers, one is described by Bevilacqua et al. (2000) where a Monte Carlo simulation approach is used for testing the weights assigned to the RPNs. In this article, the normal 1 – 10 scale was modified to the more limited 1 – 3 scale. The reason for this modification was to more readily being able to identify the numbers the RPN are based upon.

Table 3

Functional requirements of the system. Subscript numbering in accordance with Fig. 2 stream numbering.

Subsystem	Equipment	Function	Functional requirement
NG processing	Pressure regulating valve	Decrease line pressure down to system pressure	1.8 MPa $\leq p_2 \leq$ 2.0 MPa
NG processing	Desulfurizer	Remove sulfur	Exhaust $H_2S \leq$ 2 ppmv
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$
Power cycle	Gas turbine	Provide air	$m_{10} \geq$ 67.5 kg/s, $T_{10} \geq 350^\circ$C
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^\circ$C
Power cycle	Steam turbine	Generate power	$P_{ST} \geq$ 125 MW
Power cycle	Steam turbine	Supply steam to pre-reformer	$S/C = 1.5 \pm 0.1$
Power cycle	Steam turbine	Supply steam to reboiler in amine system	$p_{45} \geq$ 0.32 MPa. Heat flow provided $\geq$ 70 MJ/s
Power cycle	Generator	Generate power	Power output $\geq$ 300 MW (ISO)
Pre-heating	NG pre-heater	Pre-heat NG	350°C $\leq T_3 \leq 425^\circ$C
Pre-heating	NG/steam pre-heater	Pre-heat NG/steam mix	$T_6 \geq 480^\circ$C
Pre-heating	Air pre-heater	Pre-heat air	$T_{15} \geq 450^\circ$C
Pre-heating	ATR feed pre-heater	Pre-heat ATR feed gas	$T_8 \geq 450^\circ$C
HRSG	LP	Generate LP steam	$m_{31} \geq$ 10 kg/s
HRSG	IP	Generate IP steam	$m_{32} \geq$ 20 kg/s
HRSG	HP	Generate HP steam	$m_{37} \geq$ 40 kg/s
Reforming	Pre-reformer	Convert higher hydrocarbons. Provide preref gas	$T_6 - T_7 \geq$ 40 K, $T_7 \geq 430^\circ$C
Reforming	ATR	Convert methane. Provide syngas	900°C $\leq T_{16} \leq 1000^\circ$C
W-G shift	HTS	Convert CO to CO_2	$\Delta T \geq$ 75 K
W-G shift	LTS	Convert CO to CO_2	$\Delta T \geq$ 30 K
HX network	Syngas cooler (HE1, HE2)	Cool ATR product	300°C $\leq T_{18} \leq 450^\circ$C
HX network	HE3	Cool LTS feed	180°C $\leq T_{20} \leq 250^\circ$C
HX network	HE4	Heat fuel	$T_{29} \geq 180^\circ$C
HX network	HE5	Generate steam	$x_{52} = 1.0$
HX network	Cooler	Cool flash feed	$T_{24} \leq 30^\circ$C
HX network	Condenser	Condense steam	$p_{49} \leq$ 0.0044 MPa
HX network	Condenser	Condense steam	$p_{50} \geq$ 0.18 MPa
Pre-comb capture	Gas separation	Separate out CO_2	Remove $\geq$ 95% CO_2
Compression	Air compressor	Compress air for ATR	$p_{13} = p_{10}$. $m_{13} \rightarrow T_{16} = 950^\circ$C
Compression	CO_2 compression	Compress CO_2	$p_{55} \geq$ 10.0 MPa
Compression	Fuel compressor	Compress fuel	$p_{28} \geq$ 1.8 MPa

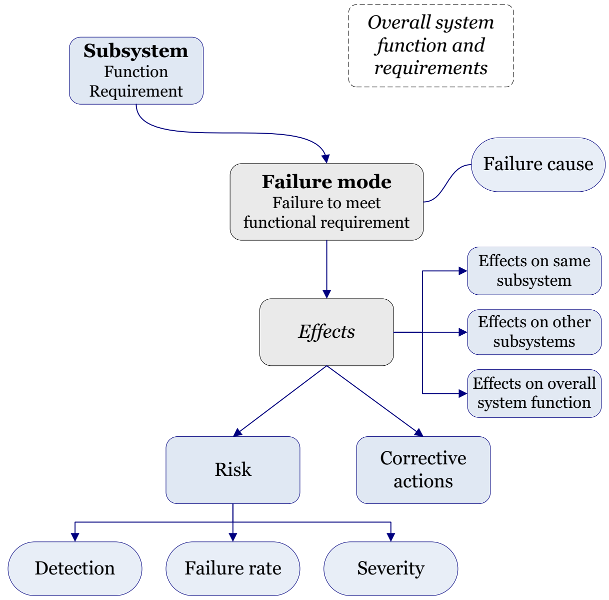

Fig. 5. Graphical representation of the FMECA approach.

The detection scale was defined as: 1 = highly detectable, almost certain detection; 2 = moderately detectable; and 3 = non-detectable.

The failure rate scale was defined as: 1 = failure unlikely; 2 = occasional failure; and 3 = frequent failure.

The severity scale was defined as: 1 = no, or very small effect; 2 = plant operating at part load or bypassing CO_2 capture; and 3 = plant shutdown.

As a basis of the analysis, it is assumed that the plant is operating at full load when a failure occurs. Furthermore, potential human errors are not considered in the analysis.

A failure mode is defined as a failure to meet a functional requirement of a specific equipment. Once a failure mode has been specified, the causes and effects of the failure need to be identified. Regarding failure effects, the effects on the same equipment where the failure occurred were first analyzed. Secondly, the effects on other equipment in the system were investigated, and finally, the overall system effects were identified. One example of failure causes and their effects is coking, or metal dusting, in the reactors and heat exchangers (Grabke and Wolf, 1986; Grabke et al., 1993). Coking in pre-reformers is investigated by Sperle et al. (2005). Several failure causes, including metal dusting in a heat exchanger for synthesis gas, are investigated by Grabke and Spiegel (2003). Catalyst degradation due to coking in reactors is analyzed by Rostrup-Nielsen (1997).

Some of the failure causes for the gas turbine were listed as a protective load shed (PLS) cause or a trip cause. A protective load shed is described as an automatic deload of the GT due to an abnormal situation such as an elevated temperature. A trip occurs when a more critical event takes place. The reason for listing a failure cause as a PLS or trip cause is because the reasons for the PLS or trip can be many.

The most common protective load shed causes are found to be:

- Thermo-acoustic instabilities
- Abnormal exhaust temperature
- Controls and instrument problems
- HRSG trip

The most common trip causes are found to be:

- Thermo-acoustic instabilities
- Flame monitor
- Abnormal exhaust temperature
- Controls and instrument problems
- Bearings (temperature, vibration)

- Manual trip

The detection rating was, for the most part, derived based on knowledge in instrumentation and controls. For example, an abnormal temperature or pressure change is easy to detect, whereas a change in a gas composition can be more difficult to sense. With the 1 – 3 scale, the numbers were fairly easy to assign. To determine the failure rate numbers, several data sources were consulted (OREDA, 2002; NERC, 2007). The severity ranking was established based on studying the effects of the various failure modes. The RPNs were computed by multiplying the detection, failure rate, and severity numbers, and must therefore range from 1 to 9.

3.2 Operability analysis

Main contributors to operability problems are (i) component and subsystem failures and (ii) system complexity and coupling between subsystems. The first aspect was discussed in the previous section.

The complexity of a plant and its control system is directly related to the number of manipulated variables. A *manipulated variable* is the variable that is changed, in a control strategy, to achieve a certain process condition. It is desirable that the complexity of a control system is as low as possible (Skogestad, 2004). The main aim is thus to have a system with a small number of manipulated variables for better operability.

As a qualitative measure of the complexity of a process we introduce the new *comparative complexity indicator* (CCI), as the number of variables that can be manipulated in a process while accounting for integration between different process areas.

The CCI is based on a well-established concept in control system design - the control degrees of freedom (CDOF), defined to be the number of manipulated variables that can be used in control loops. The CDOF of a process is therefore the number of process variables: temperatures, pressures, compositions, flow rates, or component flow rates, that can be set by the control system once the non-adjustable design variables, such as vessel dimensions, have been fixed.

It is important to distinguish between the CDOF and the design degrees of freedom, even tough the CDOF is the same as the design degree of freedom for some classes of processes (Luyben, 1996). If there are C components, then there are $(C + 2)$ *design* degrees of freedom. This implies that the designer has choice over feed stream composition, pressure, and temperature. This is true during the design of a process. In an actual control scenario, the only manipulation possible is to change the stream flow. Whatever may be the nature of the control loop (flow, level, pressure, temperature, or composition), ultimately the manipulated variable is the flow rate of a process stream.

3.2.1 Procedure for calculating control degrees of freedom

To determine the CDOF of a process is the most important step in evaluating the CCI. The procedure used in this article is adapted from Murthy Konda et al. (2006) and further expanded in Vasudevan et al. (2008). As mentioned above, the manipulated variables will always be process stream flows. The motivating question behind calculating CDOF is whether it is possible to manipulate all the process streams and, if not, what are the restrictions? This leads to:

- CDOF of a unit $\leq$ Total number of streams associated with that unit, or
- CDOF of a unit + Restraining number = Total number of streams associated with that unit.

The *restraining number* is the number of streams that cannot be manipulated. Murthy Konda et al. (2006) and Vasudevan et al. (2008) list the restraining number of commonly used units in process plants. To find the CDOF for a process, the following formula is used:

$$\text{CDOF} = N_S - N_R \tag{7}$$

where N_S is the total number of streams in the process and N_R is the sum of restraining numbers for all units in the process.

A simple utility heater or cooler has a CDOF of 2 (Murthy Konda et al., 2006). A heat exchanger implies a more complex and tightly integrated process. In this analysis, a heat exchanger should therefore have a higher CDOF than the value of 2 proposed by Murthy Konda et al. (2006). In practice, many heat exchangers have by-pass streams that usually are not shown on process flow diagrams. The number of streams for a process/process heat exchanger would then be 6, rather than 4, leading to a CDOF of 4 (compared to 2). In this article, this is included in the procedure to calculate the CDOF of heat exchangers.

Fig. 6 shows a simple Westerberg process with ten process streams (including the energy stream). The restraining numbers for each of the units in the process are shown in Table 4.

The CDOF of the Westerberg process is $10 - 4 = 6$.

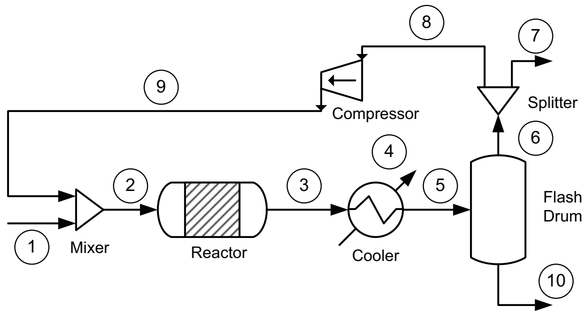

Fig. 6. Westerberg process

Table 4

Restraining numbers for units in the Westerberg process.

Unit	**Restraining no.**
Mixer	1
Reactor	0
Cooler	1
Flash drum	0
Splitter	1
Compressor	1
Total	**4**

3.2.2 Evaluating the comparative complexity indicator

The CDOF does not sufficiently represent how tightly a plant is integrated and particularly, integration between different process areas. The CCI adds a level of realism to the CDOF procedure by considering the way the different process areas of a plant are integrated.

The procedure for evaluating the CCI is shown by the flow diagram in Fig. 7. The first step involves decomposing the plant into *functional* process areas. For example, in the IRCC plant the reforming section is one process area and the CO_2 compression section another. The CDOF of each process area is then calculated as described in the previous section. If the flow between two process areas is a manipulated variable then an extra degree of freedom is added. This

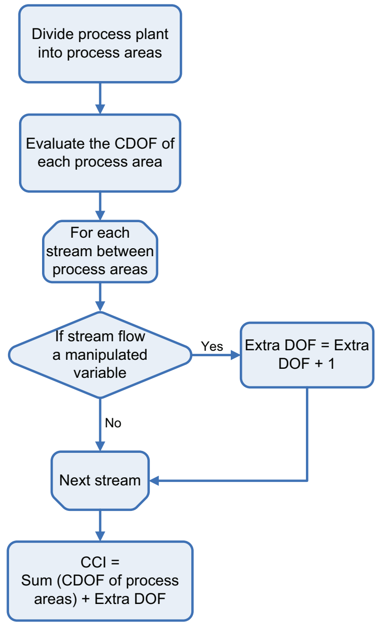

Fig. 7. Calculating the comparative complexity indicator (CCI) of a process.

check is repeated for each stream between the different process areas in the plant. The CCI is then calculated as the sum of the CDOFs of the process areas and the "extra degrees of freedom". This means the CCI is an addition of the total number of CDOFs and the, between process areas, connecting streams that are manipulated variables.

The calculation of the CCI for different IRCC configurations, as well as, for an NGCC plant with and without post-combustion capture are presented in the next section.

4 Results and discussion

The documentation of the analysis and of the results of the FMECA is comprehensive. Therefore, only a part of the results is shown in this article. Table 5 includes the failure modes with an RPN greater than 6. As seen from the table, many of the high risk results are linked to the gas turbine. This is not unexpected. In a regular NGCC plant the gas turbine and its auxiliaries are also responsible for the largest part of the forced outages (NERC, 2007).

For an IRCC, there may be additional GT failures stemming from issues related to the supply of the hydrogen-rich fuel and because of the lower level of experience with hydrogen-fired GTs compared to NG-fired GTs.

One may criticize the risk priority rankings and argue that some of them should be changed. Certainly, if another person performed the FMECA, different results would arise, but the key results, such as what equipment is most critical in the plant, should be similar if performed by someone else.

Table 5

FMECA: highest risk failure causes. Subscript numbering in accordance with Fig. 2 stream numbering.

Subsystem	**Equipment**	**Function**	**Functional requirement**	**Failure mode**	**Failure cause**	**Effects on same equipment**	**Effects on other equipment**	**Effects on overall system function**	**Detection (1-3)**	**Failure rate (1-3)**	**Severity (1-3)**	**Risk (DxFxS)**
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$60\% \leq P_{rel,GT} < 90\%$	Fuel supply	Part load operation	Reduced steam production in HRSG. Reduced power output from steam turbine	Reduced plant load	2	2	2	8
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}C$	$T_{40} < 560^{\circ}C$	Fuel supply	Part load operation	Reduced steam production in HRSG. Reduced power output from steam turbine	Reduced plant load	2	2	2	8
Reforming	Pre-reformer	Convert higher hydrocarbons. Provide preref gas	$T_6 - T_7 \geq 40$ K, $T_7 \geq 430^{\circ}C$	$T_6 - T_7 < 40$ K, $T_7 < 430^{\circ}C$	Catalyst issue	Lower conversion rate	Higher hydrocarbons to ATR (coking)	Reduced plant load. Decreased CO_2 capture rate	2	2	2	8
Reforming	ATR	Convert methane. Provide syngas	$900^{\circ}C \leq T_{16} \leq 1000^{\circ}C$	T_{16} outside range	Catalyst issue	Lower conversion rate	Hydrocarbons to HTS	Reduced plant load. Decreased CO_2 capture rate	2	2	2	8
Reforming	ATR	Convert methane. Provide syngas	$900^{\circ}C \leq T_{16} \leq 1000^{\circ}C$	T_{16} outside range	Burner issue	Possibly lower temperature. Flame shape distortion → mechanical damage to reactor walls	Hydrocarbons to HTS. Lower temp to HE1	Reduced plant load. Decreased CO_2 capture rate	2	2	2	8
W-G shift	HTS	Convert CO to CO_2	$\Delta T \geq 75$ K	$\Delta T < 75$ K	Catalyst issue	Lower conversion rate	Higher CO content to LTS	Reduced plant load. Decreased CO_2 capture rate	2	2	2	8

Subsystem	**Equipment**	**Function**	**Functional requirement**	**Failure mode**	**Failure cause**	**Effects on same equipment**	**Effects on other equipment**	**Effects on overall system function**	**Detection (1-3)**	**Failure rate (1-3)**	**Severity (1-3)**	**Risk (DxFxS)**
W-G shift	LTS	Convert CO to CO_2	$\Delta T \geq 30$ K	$\Delta T < 30$ K	Catalyst issue	Lower conversion rate	CO content to gas separation stage	Reduced plant load. Decreased CO_2 capture rate	2	2	2	8
NG processing	Pressure regulating valve	Decrease supply pressure down to system pressure	1.8 MPa $\leq p_2 \leq 2.0$ MPa	$p_2 > 2.0$ MPa	Valve malfunction	-	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$P_{rel,GT} < 60\%$	Trip cause	GT trip	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$P_{rel,GT} < 60\%$	Protective load shed cause	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$P_{rel,GT} < 60\%$	Combustion problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$P_{rel,GT} < 60\%$	NO_x emissions	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Generate power	$P_{rel,GT} \geq 90\%$	$P_{rel,GT} < 60\%$	Other gas turbine problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	2	1	3	6
Power cycle	Gas turbine	Provide air	$m_{10} \geq 67.5$ kg/s, $T_{10} \geq 350°$C	$m_{10} < 67.5$ kg/s, $T_{10} < 350°$C	Trip cause	GT trip	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide air	$m_{10} \geq 67.5$ kg/s, $T_{10} \geq 350°$C	$m_{10} < 67.5$ kg/s, $T_{10} < 350°$C	Protective load shed cause	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6

Subsystem	**Equipment**	**Function**	**Functional requirement**	**Failure mode**	**Failure cause**	**Effects on same equipment**	**Effects on other equipment**	**Effects on overall system function**	**Detection (1-3)**	**Failure rate (1-3)**	**Severity (1-3)**	**Risk (DxFxS)**
Power cycle	Gas turbine	Provide air	$m_{10} \geq 67.5$ kg/s, $T_{10} \geq 350^{\circ}$C	$m_{10} < 67.5$ kg/s, $T_{10} < 350^{\circ}$C	Combustion problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide air	$m_{10} \geq 67.5$ kg/s, $T_{10} \geq 350^{\circ}$C	$m_{10} < 67.5$ kg/s, $T_{10} < 350^{\circ}$C	NO_x emissions	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide air	$m_{10} \geq 67.5$ kg/s, $T_{10} \geq 350^{\circ}$C	$m_{10} < 67.5$ kg/s, $T_{10} < 350^{\circ}$C	Other gas turbine problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	2	1	3	6
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}$C	$T_{40} < 560^{\circ}$C	Trip cause	GT trip	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}$C	$T_{40} < 560^{\circ}$C	Protective load shed cause	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}$C	$T_{40} < 560^{\circ}$C	Combustion problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}$C	$T_{40} < 560^{\circ}$C	NO_x emissions	GT shutdown	Shutdown of all subsystems	Plant shutdown	1	2	3	6
Power cycle	Gas turbine	Provide hot gases	$T_{40} \geq 560^{\circ}$C	$T_{40} < 560^{\circ}$C	Other gas turbine problems	GT shutdown	Shutdown of all subsystems	Plant shutdown	2	1	3	6

Table 6

CDOF evaluation for process areas in IRCC plant

Area	Total streams	Restraining no.	CDOF
Reforming area	36	7	29
CO_2 capture area	24	9	15
CO_2 compression area	24	10	14
GT fuel preparation area	5	2	3
CCPP area	79	28	51
Total			**112**

For the operability analysis, the IRCC process can be decomposed into the following five process areas:

(1) Reforming area
(2) CO_2 capture area
(3) CO_2 compression area
(4) Gas turbine fuel preparation area
(5) Combined cycle power plant area

The CDOF of the five areas are calculated and shown in Table 6.

The total "extra degrees of freedom" in the system equals 3. Thus the comparative complexity indicator for the IRCC plant shown in Fig. 2 is 115. The overall efficiency of the process is 41.9%.

Process modifications will affect both efficiency and the CCI of the overall process. The subsequent paragraphs briefly analyse two process modifications with regard to the efficiency and CCI of the process and identify if the modification is favorable or not.

Process modification 1: Streams 33 and 51 are extracted from the deaerator (not shown in Fig. 2) at 105°C (pre-heated in low-temperature economizer before entering the deaerator). The low temperature heat in stream 23 could be used to pre-heat the boiler feed water from 30°C to 105°C for HP and LP steam generation in the reforming process (rather than pre-heating in low-temperature economizer). The efficiency increase by including this modification is negligible, whereas the CCI of this modified process is 118. This implies this process modification is not favorable as the complexity of the process increases without any corresponding improvement to efficiency, the decision variable.

Thus for processes with the same efficiencies, the heuristic is to select the one with least CCI.

Process modification 2: If the LP steam generator HE5 in Fig. 2 were ignored, the cooling water requirement would increase and the stream extraction from the steam turbine to the CO_2 removal section would increase. This reduces the overall efficiency to 41.5%. The CCI for this modified process is 111. The efficiency drop of 0.4%-point is significant in the context of this process. Thus, even though the complexity of this option is less than the original design, the efficiency drop causes this process modification to be disregarded.

In processes where efficiency improvements are essential, increasing complexity is acceptable within limits. For example, a process modification causing the efficiency to increase by 0.5%-points while increasing the CCI by 15 can be deemed less favorable compared to a modification that causes an efficiency increase by 0.4%-point with a CCI increase of 7.

For reference, the CCI for a natural gas combined cycle power plant without CO_2 capture is 48. Process areas such as reforming, CO_2 capture, and CO_2 compression are not included in an NGCC plant without CO_2 capture. The

CCI for a natural gas combined cycle power plant with post combustion CO_2 capture is 82.

5 Conclusions

Functional analysis and FMECA are important steps in a system reliability analysis, as they can serve as a platform and basis for further analysis. Also, the results from the FMECA can be interesting in themselves. From the FMECA performed in this work, it is clear that the gas turbine is the most critical equipment in an IRCC plant. One of the reasons for this is the significant integration present. The gas turbine feeds air to the ATR, receives fuel from the pre-combustion process, and the steam turbine supplies steam to the GT combustor. This integration has an effect on the overall reliability of the system and shows up in the FMECA, not the least in the "Effects on other equipment" column in Table 5. In addition to the integration issues, the gas turbine technology is less mature for hydrogen fuels than for natural gas fuels. It should also be mentioned that even in a natural gas fired combined cycle plant the gas turbine is the most critical equipment. With all this said, the strong dominance of gas turbine failures in a list with the highest risk priority numbers such as in Table 5 is not unexpected. Operability analysis is another important tool during the design stage. The CCI is a helpful tool in choosing the level of integration and when investigating whether or not to include a certain process feature. Incorporating the analytical approach presented in the article and displayed in Fig. 3, during the design stage of a plant, can be advantageous for the overall plant performance.

6 Acknowledgements

This work was supported by the Norwegian Research Council and StatoilHydro. The authors are grateful for valuable comments received from the reviewers and editor.

References

Andersen, T., Kvamsdal, H. M., Bolland, O., 2000. Gas turbine combined cycle with CO_2 capture using auto-thermal reforming of natural gas. In: ASME Turbo Expo 2000. Munich, Germany.

Åström, K., Fontell, E., Virtanen, S., 2007. Reliability analysis and initial requirements for FC systems and stacks. Journal of Power Sources 171 (1), 46–54.

Barton, G. W., Chan, W. K., Perkins, J. D., 1991. Interaction between process design and process control: The role of open-loop indicators. Journal of Process Control 1 (3), 161–170.

Beér, J. M., 2007. High efficiency electric power generation: The environmental role. Progress in Energy and Combustion Science 33 (2), 107–134.

Bevilacqua, M., Braglia, M., Gabbrielli, R., 2000. Monte Carlo simulation approach for a modified FMECA in a power plant. Quality and Reliability Engineering International 16 (4), 313–324.

Blanco, A. M., Bandoni, J. A., 2003. Interaction between process design and process operability of chemical processes: An eigenvalue optimization approach. Computers & Chemical Engineering 27 (8-9), 1291–1301.

Bohm, M. C., Herzog, H. J., Parsons, J. E., Sekar, R. C., 2007. Capture-ready coal plants - Options, technologies and economics. International Journal of Greenhouse Gas Control 1 (1), 113–120.

Bonzani, F., Gobbo, P., 2007. Operating experience of high flexibility syngas burner for IGCC power plant. Vol. 2 of Proceedings of the ASME Turbo Expo. American Society of Mechanical Engineers, New York, pp. 65–71.

Chiesa, P., Lozza, G., Mazzocchi, L., 2005. Using hydrogen as gas turbine fuel. Journal of Engineering for Gas Turbines and Power 127 (1), 73–80.

Christensen, T. S., Christensen, P. S., Dybkjær, I., Hansen, J. H. B., Primdahl, I. I., 1998. Developments in autothermal reforming. Studies in Surface Science and Catalysis 119, 883–888.

Christensen, T. S., Primdahl, I. I., 1994. Improve syngas production using autothermal reforming. Hydrocarbon Processing 73 (3).

Descamps, C., Bouallou, C., Kanniche, M., 2008. Efficiency of an Integrated Gasification Combined Cycle (IGCC) power plant including CO_2 removal. Energy 33 (6), 874–881.

Dhillon, B. S., Rayapati, S. N., 1988. Chemical-system reliability: A review. IEEE Transactions on Reliability 37 (2), 199–208.

Dybkjær, I., 1995. Tubular reforming and autothermal reforming of natural gas - an overview of available processes. Fuel Processing Technology 42 (2-3), 85–107.

Eti, M. C., Ogaji, S. O. T., Probert, S. D., 2007. Integrating reliability, availability, maintainability and supportability with risk analysis for improved operation of the Afam thermal power-station. Applied Energy 84 (2), 202–221.

Grabke, H. J., Krajak, R., Müller-Lorenz, E. M., 1993. Metal dusting of high temperature alloys. Werkstoffe und Korrosion 44 (3), 89–97.

Grabke, H. J., Spiegel, M., 2003. Occurrence of metal dusting - referring to failure cases. Materials and Corrosion 54 (10), 799–804.

Grabke, H. J., Wolf, I., 1986. Carburization and oxidation. Materials Science and Engineering 87, 23–33.

Higman, C., DellaVilla, S., Steele, B., 2006. The reliability of integrated gasification combined cycle (IGCC) power generation units. In: Achema. Frankfurt, Germany.

IEC 60812, 2006. Analysis techniques for system reliability - Procedures for failure mode and effect analysis (FMEA). International Electrotechnical Commission, Geneva.

ISO 20815, 2008. Petroleum, petrochemical and natural gas industries – Production assurance and reliability management. International Organization for Standardization, Geneva.

Luyben, W. L., 1996. Design and control degrees of freedom. Ind. Eng. Chem. Res. 35 (7), 2204–2214.

Moulijn, J. A., Makkee, M., Diepen, A. v., 2007. Chemical Process Technology. Wiley, Hoboken, NJ.

Murthy, D. N. P., Rausand, M., Østerås, T., 2008. Product Reliability; Specification and Performanace. Springer, London.

Murthy Konda, N. V. S. N., Rangaiah, G. P., Krishnaswamy, P. R., 2006. A simple and effective procedure for control degrees of freedom. Chemical Engineering Science 61 (4), 1184–1194.

NERC, 2007. North American Electric Reliability Corporation: GADS report 2002-2006. http://www.nerc.com.

OREDA, 2002. Offshore Reliability Data Handbook, 4th Edition. DNV, Høvik, Norway.

Perrow, C., 1999. Normal Accidents; Living with High-Risk technologies. Princeton University Press, Princeton, NJ.

Rausand, M., Høyland, A., 2004. System Reliability Theory: Models, Statistical Methods, and Applications, 2nd Edition. Wiley, Hoboken, NJ.

Rostrup-Nielsen, J. R., 1997. Industrial relevance of coking. Catalysis Today 37 (3), 225–232.

Scottish Centre for Carbon Storage, 2009. School of Geosciences, University of Edinburgh, Scotland. http://www.geos.ed.ac.uk/ccsmap.

Shilling, N. Z., Jones, R. M., 2003. The impact of fuel flexible gas turbine control systems on integrated gasification combined cycle performance. Vol. 1 of Proceedings of the ASME Turbo Expo. American Society of Mechanical Engineers, New York, NY, USA, pp. 259–265.

Skogestad, S., 2004. Control structure design for complete chemical plants. Computers and Chemical Engineering 28 (1-2), 219–234.

Sperle, T., Chen, D., Lodeng, R., Holmen, A., 2005. Pre-reforming of natural gas on a Ni catalyst: Criteria for carbon free operation. Applied Catalysis A: General 282 (1-2), 195–204.

Teng, S.-H., Ho, S.-Y., 1996. Failure mode and effects analysis: An integrated approach for product design and process control. The International Journal of Quality & Reliability Management 13 (5), 8–26.

Teoh, P. C., Case, K., 2004. Failure modes and effects analysis through knowledge modelling. Journal of Materials Processing Technology 153-154 (1-3), 253–260.

Todd, D. M., Battista, Robert, A., 2000. Demonstrated applicability of hydrogen fuel for gas turbines. In: Gasification 4 the Future. Noordwijk, Netherlands.

van Loo, S., van Elk, E. P., Versteeg, G. F., 2007. The removal of carbon dioxide with activated solutions of methyl-diethanol-amine. Journal of Petroleum Science and Engineering 55 (1-2), 135–145.

Vannby, R., Winter Madsen, S. E. L., 1992. Adiabatic preforming. Ammonia Plant Safety (and Related Facilities) 32, 122–128.

Vasudevan, S., Murthy Konda, N. V. S. N., Rangaiah, G. P., 2008. Control degrees of freedom using the restraining number: Further evaluation. Asia-Pacific Journal of Chemical Engineering.