A Plantwide Control Procedure with Application to Control Structure Design for a Gas Power Plant*

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October 31, 2002

1 Introduction

A chemical plant may have thousands of measurements and control loops. In practice, the control system is usually divided into several layers, separated by time scale, including

- scheduling (weeks)
- site-wide optimization (day)
- local optimization (hour)
- supervisory (predictive, advanced) control (minutes)
- regulatory control (seconds)

The layers are linked by the controlled variables, whereby the setpoints are computed by the upper layer and implemented by the lower layer. An important issue is the selection of these variables.

Plantwide control deals with the structural decisions of the control system for an overall plant. Usually these decisions are taken based on pure experience and engineering insight. A recent review of the literature on plantwide control and a plantwide control procedure can be found in Larsson and Skogestad (2001). In this paper we extend the plantwide control procedure of Larsson and Skogestad (2001) and apply it to a gas power plant. A systematic approach to plantwide control starts by formulating the operational objectives. This is done by defining a cost function J that should be minimized with respect to the N_{opt} optimization degrees of freedom, subject to a given set of constraints.

*Presented at AiChe 2002 Annual Meeting, Indianapolis, USA, 3.-8. Nov. 2002
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2 Plantwide Control Structure Procedure

The plantwide control structure procedure is divided in two main parts:

I. Top-down analysis, including definition of operational objectives and consideration of degrees of freedom available to meet these. Steps:

1. Manipulated variables

2. Degrees of freedom analysis

3. Primary controlled variables (selected based on steady-state economics)

4. Production rate manipulator

II. Bottom-up design of the control system, starting with the stabilizing control layer. Steps:

5. Structure of regulatory control layer (including what to control)

6. Structure of supervisory control layer (including MPC applications)

7. Structure of optimization layer

The procedure is generally iterative and may require several loops through the steps, before converging at a proposed control structure.

A very important issue is to select what to control. First, we need to control the variables directly related to ensuring optimal economic operation (the primary controlled variables in step 3):

- Control active constraints
- Select unconstrained controlled variables so that with constant setpoints the process is kept close to its optimum in spite of disturbances and implementation errors.

Sub-steps:

Step 3.1 Determination of degrees of freedom for optimization

Step 3.2 Definition of optimal operation (cost and constraints)

Step 3.3 Identification of important disturbances

Step 3.4 Optimization (nominally and with disturbances)

Step 3.5 Identification of candidate controlled variables

Step 3.6 Evaluation of loss for alternative combinations of controlled variables (loss imposed by keeping constant setpoints when there are disturbances or implementation errors)

Step 3.7 Evaluation and selection (including controllability analysis)

In addition, we need to control variables in order to achieve satisfactory regulatory control (these are the secondary controlled variables, see step 5):

- Control unstable or integrating liquid levels
- Stabilize other unstable modes, e.g. for an exothermic reactor
- Control variables which would otherwise "drift away" due to large disturbance sensitivity. This involves controlling extra local measurements which can be used to local disturbance rejection.

3 Gas Power Plant Case Study

In this paper we apply the plantwide control structure procedure to a simple combined cycle power plant which is shown in figure 1. The process is developed from the work of Gronnaass (2001) and Saue (2002). The combined cycle power plant is composed of a gas cycle and a steam cycle. The gas cycle is fed by fuel (compressed natural gas and air which react in the combustor and give flue gas with high temperature). Electric power is produced through an expansion of the gas in the gas turbine. The steam cycle is supplied with the exhaust gases from the gas turbine and generates both electricity and steam for other industrial processes. In this paper we only consider production of electricity in the steam cycle. In the steam cycle heat is transfered from the flue gas to the steam in the pre-heater, economizer, evaporator and super-heater. Electric power is produced through the expansion of steam in the steam turbine. The pressure is increased by the low pressure (LP) pump and the high pressure (HP) pump. The deaerator is included to remove oxygen. A low pressure recycle around the pre-heater is included to avoid too low water inlet temperature to the pre-heater. A high pressure recycle from the super-heater outlet steam to the deaerator is included and gives a manipulated variable with fast and large effect on the deaerator pressure. Bypasses are included around all the heat exchangers (which make it possible to optimize the heat exchanger network operation).

Compared with most usual combined cycle power plants we have added a compressor on the gas feed flowrate and assumed that all compressors and turbines have different shafts.

3.1 Manipulated variables

The process has thirteen manipulated variables (gas compressor work, air compressor work, gas turbine work, super-heater bypass, evaporator bypass, economizer bypass, pre-heater bypass, LP pump work, HP pump work, LP recycle flowrate, HP recycle flowrate, steam turbine work and cooling water flowrate), see table 1.

3.2 Degrees of freedom analysis

There are eight steady-state degrees of freedom, see table 1. The equality constraint on the deaerator pressure consume one degree of freedom. The gas and steam turbine speeds are assumed given and consume two degrees of freedom. Two holdups which need to be control, have no steady-state effect and consume two degrees of freedom. When the gas feed flowrate or the net electricity production is given, there are only seven steady-state degrees of freedom.

3.3 Primary controlled variables

3.3.1 Degrees of freedom for optimization

The number of degrees of freedom for optimization is equal the number of steady-state degrees of freedom, seven or eight, see table 1.

3.3.2 Definition of optimal operation

Economic objective
The economic objective is to maximize the profit (the value of the products minus the costs of the utility and raw materials).

$$J = p_{el}(W_{t,g} + W_{t,s} - W_{c,g} - W_{c,a} - W_{p,LP} - W_{p,HP}) - p_g F_g - p_a F_a - p_{cw} F_{cw} \quad (1)$$

We here consider three different cases:

- Given net electricity production (case I)
- Given fuel gas flowrate (case II)
- Free net electricity production and free gas feed flowrate (case III)

If we assume free air and cooling water (acceptable with respect to Norwegian conditions), the objective for case I (given net electricity production) can be simplified to minimize the use of natural gas.

$$J = F_{gas} \quad (2)$$

If we assume free air and cooling water, the objective for case II can be simplified to maximize the net electricity production:

$$J = -W_{net} \quad (3)$$

If we assume free air and cooling water, the objective for case III when we assume p_{el}/p_{gas} equal 0.1, can be simplified to minimize the following function:

$$J = -0.1W_{net} + F_{gas} \quad (4)$$

The price ratio in Norway today is approximate 0.038 ($0.15 NOK/kWh$ and $0.7 NOK/Sm^3$) and gives larger operating costs than operating revenues. All three cases will probably occur during the plants life cycle, so we ought to find a control structure for each case and hopefully simple switches between them. Case III is especially relevant because of the liberation of the energy market. We will therefore mostly concentrate on case III.

Constraints

Gas turbine inlet temperature
According to Kehlhofer (1991) it is necessary to protect the last turbine stage from too high temperature. Kehlhofer (1991) switches from controlling the load to controlling the gas turbine inlet temperature.

The maximum temperature in the furnace is about 1500^oC. The gas turbine inlet temperature is wanted as high as possible to increase the efficiency. The optimal maximum turbine inlet temperature should ideally be computed by minimizing the sum of investment cost and operating cost over the life time.

Steam turbine inlet temperature
Usually no control of steam turbine inlet temperature is required because the flue gas temperature is quite low. However, with supplementary firing, control of the steam turbine inlet temperature may be crucial (Kehlhofer 1991).

Usually the steam turbine inlet temperature is 530-560^oC and the steam turbine inlet pressure is up to 250 bar. If we want to use higher temperature (which is wanted from a thermodynamic point of view), it requires another type of steal, which is more expensive and has other properties (Bolland 1989).

Wetness in the steam turbine exhaust gas
This is no longer a main problem, but the turbine efficiency is reduced with increasing wetness.

Dew-point of the exhaust gas
To keep low temperature corrosion within acceptable bounds, the feed water temperature must not, even in the lower load range, drop significantly below the acid or water dew-point (Kehlhofer 1991). The dew-point for flue gas from the gas turbine is about 40^oC. The flue gas temperature at the wall is approximately equal to the condensing water temperature (Bolland 1989).

Deaerator pressure
The deaerator is included to remove oxygen from the steam cycle. At 1 bar the separation of oxygen from water is acceptable.

Gas and steam turbine speed
The turbine speed is usually kept constant. It is however possible to operate the turbine with varying speeds, this requires changing the frequency to the alternating current. Norum (2002) assumed the efficiency equal 90%. The improved operation connected to varying the turbine speed must increase the turbine work with more than 10% to be profitable, and this seems to be rather tough to obtain.

Compressor surge
During operation compressor surge must be avoided in the gas and air compressor.

Flowrates

All flowrates should be at least zero. In addition we have upper limits, which are especially important for the low pressure recirculation flowrate and the cooling water flowrate.

The most important constraints are summarized below:

$$T_{comb} \leq 1500^oC \tag{5}$$

$$T_{w,pre,n} \geq 63^o \tag{6}$$

$$T_{w,super,1} \leq 550^oC \tag{7}$$

$$F_{cw} \leq F_{cw,max} \tag{8}$$

$$F_{LP,valve} \leq F_{LP,valve,max} \tag{9}$$

$$f_{surge,comp,gas}(P_{comb}/P_{gas}, N_{comp,gas}, F_{gas}) \leq 0 \tag{10}$$

$$f_{surge,comp,air}(P_{comb}/P_{air}, N_{comp,air}, F_{air}) \leq 0 \tag{11}$$

The temperature constraints are soft which means that limited violations of the temperature constraints in transients are acceptable. The other constraints are hard and cannot be violated.

3.3.3 Identification of important disturbances

Nominal values and expected variation for the disturbances are given in table 2. The most important disturbance for case I is the net electricity production (W_{net}) and for case II the gas feed flowrate (F_{gas}). For case III the disturbance with largest effect on the optimum is the air inlet temperature (T_{air}).

3.3.4 Steady-state optimization

We have done optimization for different gas feed flowrates. Gas feed flowrates with corresponding net electricity production, number of unconstrained degrees of freedom and active constraints are presented in table 3. Case I and case II have no unconstrained steady-state degrees of freedom, while case III has one unconstrained steady-state degree of freedom. At high flowrates the gas turbine inlet temperature is at maximum in optimum. The HP recirculating flowrate is at minimum. The cooling water flowrate and the LP recirculating flowrate are at maximum. The bypasses around the super-heater, the evaporator and the economizer are closed. The optimal active constraints are changed when the gas feed flowrate is less than 38.4267 kg/s.

The optimum for case III is obtained with a gas feed flowrate of 49.4784 kg/s and a net electricity production of 1195.3742 MW. The objective function J is then 70.0590.

The maximum electricity production is achieved with a gas feed flowrate of 50.6697 kg/s. The net electricity production is then 1202.2403 MW.

The maximum gas feed flowrate is 52.1408 kg/s. The net electricity production is then 1042.8165 MW.

The maximum efficiency (W_{net}/F_{gas}) of 25.4387 is achieved with a gas feed flowrate equal 33.1756 kg/s.The optimal active constraints have changed; the maximum steam turbine inlet temperature is active instead of the minimum high pressure recirculating flowrate.

We have also done optimization for case III with respected to different disturbances. The optimal active constraints are not changed. Expected disturbances in the air inlet temperature have the largest effect on the optimal operation.

3.3.5 Identification of candidate controlled variable sets

The candidate controlled variables (c) consist of the manipulated variables (u), measurements (y_m) and flowratios (FR). Possible measurements of pressure (P), flowrates (F), temperatures (T), duty (Q), work (W) and compressor speed (N) are included. The candidate controlled variables are summarized below.

$$c^T = [u\ P\ F\ T\ Q\ W\ N\ FR] \tag{12}$$

The implementation errors are initially assumed as $\pm$ 10% for flowrates, $\pm$ 2.5% for pressures, $\pm$ 1 ^oC for temperatures, $\pm$ 30% for duties, $\pm$ 30% for work and $\pm$ 10% for compressor speed. An exception is combustor temperature / gas turbine inlet temperature (T_{comb}) where the temperature is very high and the implementation error is expected equal $\pm$ 10 ^oC.

For case I and case II there are no unconstrained degrees of freedom at optimum. Because we expect a so high gas feed flowrate (more than 38.4287 kg/s) that the optimal active constraints are not changing, we select to control the variables lying at the optimal active constraints (see table 3): Maximum gas turbine inlet temperature, maximum cooling water flowrate, minimum high-pressure recirculating flowrate, maximum low-pressure recirculating flowrate, no super-heater bypass, no evaporator bypass and no economizer bypass.

For case III we select to control the active constraints (see table 3). We have still one steady-state degree of freedom and need to select one more controlled variable. The number of candidate controlled variables is large. Initial screening is performed by maximizing the steady-state gain ($|G(0)|$) where $G(0)$ is obtained with the active constraints kept constant. The candidate controlled variables are scaled with respect to variation in optimal values (maximum distance between the nominal optimal value and optimal value for some expected disturbances) and implementation errors. The super-heater gas inlet temperature ($T_{g,super,1}$) seems to be the most promising variable, see table 7. Candidate controlled variables with zero gain have no steady-state effect or are dependent on other selected controlled variables. Infeasible alternatives have often small steady-state gain, but not necessarily.

3.3.6 Loss evaluation with nominal optimal setpoints

For the alternative candidate controlled variable sets we evaluate the economic loss imposed by using constant setpoints instead of re-optimization (with no implementation errors) for case III. Average loss (and setpoints) when using nominal optimal setpoints are shown in table 8 which contains only the feasible alternatives. In addition we have evaluated the loss connected to using perfect on-line optimization.

With **perfect on-line optimization** we assume that the disturbances are perfectly tracked by the optimization, so that the loss with respect to disturbances is zero. Introducing on-line optimization do not affect the loss with respect to the implementation errors. We assume that the optimal solution is implemented by controlling the active constraints and some unknown unconstrained controlled variables with zero implementation errors. The loss by using perfect on-line optimization is then connected to the implementation errors of the variables at the active constraints and is found by doing optimization where simple back-off from the optimal active constraints are included.

In table 8 we have evaluated the loss for different candidate controlled variables with expected disturbances in inlet air temperature and implementation errors in the combustor temperature and the selected candidate controlled variable when we keep constant nominal optimal setpoints for case III. Some general trends from table 8: For feasible alternatives the loss connected to the disturbances (d) is rather small. The loss connected to the implementation error in the optimal active constraint (e_1), combustor temperature, is significant, but rather independent of selected alternative. There is much money to earn by reducing the implementation error connected to the combustor temperature (T_{comb}). The difference in loss between the alternatives is due to the implementation error in the unconstrained controlled variable. Temperatures and pressures are significant less loss-makers than the other unconstrained controlled variables.

Controlling the super-heater gas inlet temperature ($T_{g,super,1}$) gives the smallest loss. In this case the reduction of loss by using on-line optimization is limited to 0.16%. The super-heater gas inlet temperature ($T_{g,super,1}$) which is the pinch temperature in the steam cycle heat-exchanger network, is least sensitive to implementation error. Physically, controlling the super-heater inlet temperature simplifies the operation of the plant by decoupling the gas and steam cycle operation.

The through-put is limited by the required deaerator pressure and the bypass structure. The alternatives which are infeasible, have problems with limited through-put for a increase in the air inlet temperature (disturbance) and / or implementation error connected to the unconstrained controlled variable.

Since the loss when controlling the super-heater gas inlet temperature is not much larger than loss with on-line optimization, the possible improvement by finding other combinations of variables to control not so large.

3.3.7 Loss evaluation with robust optimal setpoints

Optimal back-off (or robust optimization) is used to improve the operation point by achieving feasibility or by just keeping the operation closer to optimum. The possible reduction in loss by using optimal back-off is small for feasible alternatives, because the loss with nominal optimal setpoints is not much larger than the loss by using perfect on-line optimization. Optimal back-off is here used for some alternatives which are infeasible with nominal optimal setpoint. We will here look at controlling the gas feed flowrate (in addition to the active constraints) which is infeasible with nominal optimal setpoints. Optimal back-off when controlling the gas feed flowrate is obtained by reducing the gas feed flowrate from 48.72 kg/s to 46.12 kg/s. Controlling the gas feed flowrate is now feasible, but gives a large loss (9.44%), see table 4.

An alternative way to achieve feasibility is to use flexible back-off instead of optimal back-off. **Flexible back-off** is based on the nominal optimal setpoints, and feasibility is achieved by solving the following problem on-line:

$$min_{u,x,c_{s,flex}}(c_s - c_{s,flex})^T Q(c_s - c_{s,flex}) \tag{13}$$

$$f(u,x,d) = 0 \tag{14}$$

$$g(u,x,d) \leq 0 \tag{15}$$

$$c(u,x,d) = c_{s,flex} + e \tag{16}$$

$$d \epsilon D \tag{17}$$

$$e \epsilon E \tag{18}$$

Flexible back-off is often included in the steady-state optimization part of a model predictive controller (MPC) to achieve feasibility. Flexible back-off gives an measure of the steady-state effect of the MPC. We here give the controlled variables at optimal active constraints higher priority than the unconstrained controlled variable and only allow flexible back-off in the unconstrained controlled variable. Using flexible back-off controlling the gas feed flowrate is now feasible, but gives large loss (9.23%), see table 5. Flexible back-off gives somewhat less loss than optimal back-off.

For case I and II small gas feed flowrates or net electricity production can occur and the optimal active constraints often change. The feasibility region is smaller. Backoff may be necessary to achieve feasibility and to improve the operation. Flexible back-off is usually best to achieve feasibility when the feasibility region is small. (Flexible back-off is also good when we have a small economic region and a large feasibility region.)

We select the super-heater gas inlet temperature in addition to the gas turbine inlet temperature and deaerator pressure as primary controlled variables, because this alternative gives the smallest loss.

3.4 Production rate manipulator

Bottlenecks

Two extensive variables are at constraints at optimum: The cooling water flowrate (F_{cw}) and the low pressure recirculating flowrate ($F_{LP,valve}$). Increasing the maximum cooling water flowrate ($F_{cw,max} \Delta F_{gas}/\Delta F_{cw}$=0.007) has a relatively larger effect than increasing the maximum low pressure recirculating flowrate ($F_{LP,valve,max} \Delta F_{gas}/\Delta F_{LP,valve}$=0.0044), and the cooling water flowrate is therefore the main bottleneck. When we reoptimize with a fixed gas feed flowrate smaller than the nominal optimal gas feed flowrate, the nominal optimal active constraints are still active. The bottlenecks are steady-state (design) bottlenecks, not dynamic (control) bottlenecks.

We are here dealing with a gas phase system and the production manipulator is not so clear.

3.5 Structure of regulatory control layer

Stabilization
The levels in the evaporator drum and the condenser drum need to be stabilized. The deaerator drum is much larger than the other drums and controlling the deaerator drum level is not needed. At high flowrates the steam turbine flowrate is given, independent of the steam turbine speed. Since the steam turbine flowrate is the inlet flowrate to the condenser drum, we select to stabilize the condenser drum level by manipulating the outlet flowrate, i.e. the low-pressure pump flowrate. Since the steam turbine flowrate gives the outlet flowrate to the evaporator drum level, we select to stabilize the evaporator drum level by manipulating the inlet flowrate, i.e. the high-pressure flowrate pump.

Local disturbance rejection
Local disturbance rejection is done by controlling flowrates by manipulating the valve openings, compressor speeds and pump effects, respectively. These loops are not included in our study.

3.6 Structure of supervisory control layer

Decentralized control
We want to control the gas turbine inlet temperature (T_{comb}), the super-heater gas inlet temperature ($T_{g,super,1}$) and the deaerator pressure ($P_{w,deaerator}$):

$$y^T = [T_{comb}\ T_{g,super,1}\ P_{w,deaerator}] \tag{19}$$

Available manipulated variables are the gas feed flowrate (F_{gas}), the steam feed flowrate (F_{air}) and the pre-heater bypass flowrate ($F_{g,pre,bypass}$).

$$u^T = [F_{gas}\ F_{air}\ F_{g,pre,bypass}] \tag{20}$$

The frequency dependent relative gain array (RGA) is shown in figure 2. Pairing the deaerator pressure with the pre-heater bypass flowrate gives no two-way interactions with the other loops. The other two pairings are not so clear. Pairing on the diagonal elements give the smallest RGA-number, see figure **??**. We avoid pairing on the negative elements in the steady-state RGA:

$$\Lambda(0) = \begin{bmatrix} 1.8674 & -0.8674 & 0.0000 \\ -0.8674 & 1.8674 & 0.0000 \\ -0.0000 & 0.0000 & 1.0000 \end{bmatrix} \tag{21}$$

Multivariable control
A multivariable controller including the two temperature control loops could be evaluated, since the interactions seem to be rather strong.

Model predictive control (MPC)
Because the selected alternative has no feasibility problems (problems with violation of constraints), MPC is not needed.

3.7 Structure of optimization layer

The loss reduction by introducing on-line optimization when we control the inlet gas temperature, is small (less than 0.16%), see table 8. On-line optimization is therefore not needed.

3.8 Validation of selected control structure

Figure 4 shows the selected control structure for case III. Stabilization is performed by controlling the evaporator drum level and condenser drum level by respectively manipulating the flowrate through the HP pump and the LP pump. The inlet gas turbine temperature is controlled by manipulating the gas feed flowrate, and the deaerator temperature is controlled by manipulating the pre-heater bypass flowrate. The super-heater gas inlet temperature is controlled by manipulating the air feed flowrate. When the net electricity production is given (case I), the feed gas flowrate is used to control the net electricity production instead of the super-heater gas inlet temperature, see Figure 5. The gas turbine inlet temperature is now controlled by the air feed flowrate. When the gas feed flowrate is given (case II), the super-heater gas inlet temperature is no longer controlled, see figure 6. The gas turbine inlet temperature is controlled by manipulating the air feed flowrate.

We will now validate the proposed control structure for case III by doing nonlinear simulation. The controllers are tuned by using simple IMC-tuning rules, see (Skogestad 2001). The wanted closed-loop time constants (τ_c) are selected equal 0.01 s for the inlet gas turbine temperature controller and the super-heater gas inlet temperature, 120 s for the deaerator pressure controller and 10 s for the level controllers. Figures 7 - 12 show the responses in controlled variables (the gas turbine inlet temperature, the super-heater gas inlet temperature and the deaerator pressure) and the corresponding manipulated variables (gas feed flowrate, air feed flowrate and pre-heater bypass flowrate) for a step in the feed air temperature (increased with 10^oC). The other disturbances have only minor effect on the operation. The initial control errors in the gas turbine inlet temperature and the super-heater gas inlet temperature are large, respectively $+40^oC$ and $+14^oC$. The temperature constraints are soft and the constraint violations during fast transients are acceptable.

4 Conclusion

A systematic procedure for selecting plantwide control structure are demonstrated on a simple combined cycle power plant process. Selection of the primary controlled variables is the most important step in procedure.

The process has one unconstrained steady-state degree at optimal operation. Controlling the super-heater gas inlet temperature in addition to the variables at active constraints at optimum gives the smallest loss (0.16% compared to on-line optimization). Controlling the super-heater inlet temperature partly decouples the operation of the gas turbine and the steam turbine cycle.

The disturbances have rather small effect on the optimal operation of the process. The implementation errors connected to the controlled variables have significantly larger effect. The main difference in loss between the alternatives comes from the implementation of the unconstrained controlled variables. Temperatures and pressures give significant less loss than the other candidate controlled variables.

The through-put limits of the process make some of the alternatives infeasible, e.g. controlling the gas feed flowrate. Feasibility can be achieved by doing back-off. Both optimal back-off (robust optimization offline) or flexible back-off (on-line) give feasibility when controlling the gas feed flowrate, but significantly larger loss than controlling the super-heater gas inlet temperature with nominal optimal setpoints.

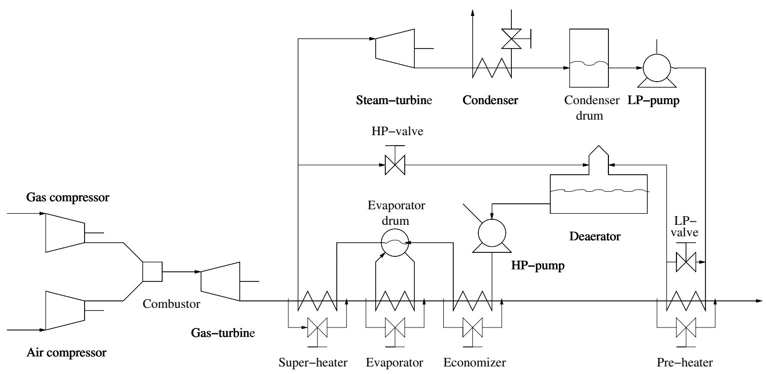

Figure 1: A simple combined cycle power plant process

Table 1: Manipulated variables and degrees of freedom

	Manipulated variables		13
	Gas compressor work	$W_{c,g}$	
	Air compressor work	$W_{c,a}$	
	Gas turbine work	$W_{t,g}$	
	Super-heater bypass	F_{super}	
	Evaporator bypass	F_{evap}	
	Economizer bypass	F_{econ}	
	Pre-heater bypass	F_{pre}	
	LP pump work	$W_{p,LP}$	
	LP valve opening	z_{LP}	
	HP valve opening	z_{HP}	
	HP pump work	$W_{p,HP}$	
	Steam turbine work	$W_{t,s}$	
	Cooling water flowrate	F_{cw}	
−	Levels with no steady-state effect		2
	Condenser drum holdup	M_{cond}	
	Evaporator drum holdup	M_{evap}	
−	Other		4/3
	Deaerator pressure	P_d	
	Gas turbine speed	$\omega_{t,g}$	
	Steam turbine speed	$\omega_{t,s}$	
	Net electricity load(case I)	$W_{net,s}$	
	Gas feed flowrate (case II)	F_{gas}	
=	Number of steady-state degrees of freedom		7/8

Table 2: Disturbances with expected variations

Disturbance	Nominal value	Expected variation	Unit
Gas feed pressure (P_g)	20	4	bar
Gas feed temperature (T_g)	10	2	^oC
Air feed pressure (P_a)	1	0.01	bar
Air feed temperature (T_a)	10	10	^oC
Cooling water temperature (T_{cw})	10	5	^oC
Flue gas outlet pressure (P_f)	1	0.01	bar
Net electricity production(W_{net})(case I)	1500	500	MW
Gas feed flowrate (F_{gas})(case II)	25	5	kg/s

Table 3: Optimization for different gas feed flowrates (F_{gas}) with corresponding net electricity production (W_{net}), number of unconstrained degrees of freedom (DOF (case I, case II, case III)) and active constraints (1: Max gas turbine inlet temperature, 2: Min pre-heater inlet water temperature, 3: Max cooling water flowrate, 7: Min HP recycle steam flowrate, 11: Min super-heater bypass flowrate, 12: Min evaporator bypass flowrate, 13: Min economizer bypass flowrate, 14: Min pre-heater bypass flowrate, 17: Max LP recycle flowrate, 18: Max steam turbine inlet temperature)

F_{gas} $[kg/s]$	W_{net} $[MW]$	DOF (I,II,III)	Active constraints	Comment
40	1010.9215	0, 0, 1	1,3,7,11,12,13,17	
45	1121.1309	0, 0, 1	1,3,7,11,12,13,17	
49.4784	1195.3742	0, 0, 1	1,3,7,11,12,13,17	Optimum for case III
50	1199.7787	0, 0, 1	1,3,7,11,12,13,17	
50.6697	1202.2403	0, 0, 1	1,3,7,11,12,13,17	Maximum net electricity production (W_{net})
52.1408	1042.8165	0, 0, 1	1,3,7,11,12,13,17	Maximum gas feed flowrate (F_{gas})

Table 4: Average objective value and loss for different candidate controlled variables (c) when keeping robust optimal setpoints (c_s) for case III.

Rank	c_1	c_2	$c_{s,1}$	$c_{s,2}$	$J(d)$	$J(e_1)$	$J(e_2)$	$J_{average}$	Loss (%)
Optimal	–	–	–	–	73.0582	72.6367	72.6367	72.6644	–
On-line opt.	–	–	–	–	70.9782	70.8776	70.8755	70.9054	2.42
–	T_{comb}	F_{gas}	1490	46.13	65.9879	67.0398	63.7524	65.8056	9.44
–	T_{comb}	F_{gas}/F_{air}	1490	0.02410	65.5781	65.9966	40.7408	58.6695	19.26

Table 5: Average objective value and loss for different candidate controlled variables (c) when doing flexible back-off with nominal optimal setpoints (c_s) for case III.

Rank	c_1	c_2	$c_{s,1}$	$c_{s,2}$	$J(d)$	$J(e_1)$	$J(e_2)$	$J_{average}$	Loss (%)
Optimal	–	–	–	–	73.0582	72.6367	72.6367	72.6644	–
On-line opt.	–	–	–	–	70.9782	70.8776	70.8755	70.9054	2.42
–	T_{comb}	F_{gas}	1490	48.72	63.5387	68.2093	64.9286	65.9553	9.23

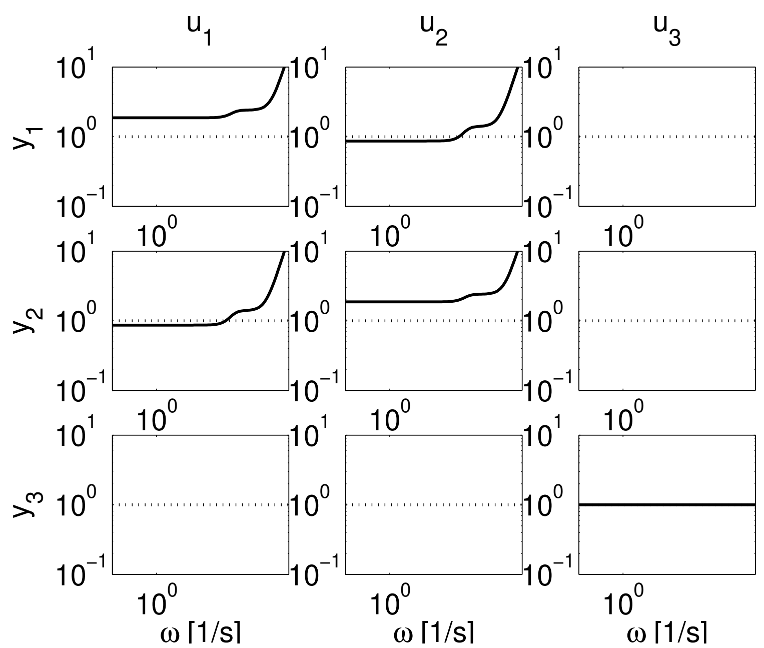

Figure 2: Frequency-dependent RGA ($\Lambda(j\omega)$)

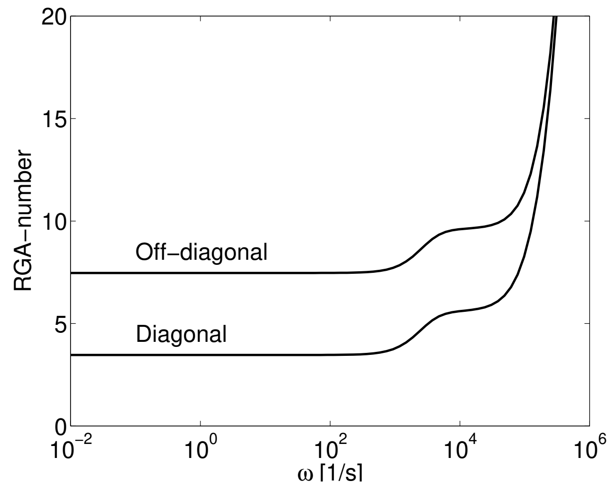

Figure 3: Frequency-dependent RGA-number ($||\Lambda(j\omega) - I||_{sum}$)

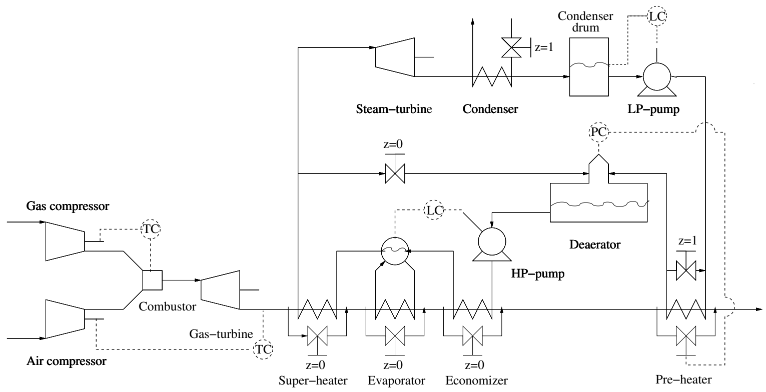

Figure 4: Combined cycle power plant process with proposed control structure for case III

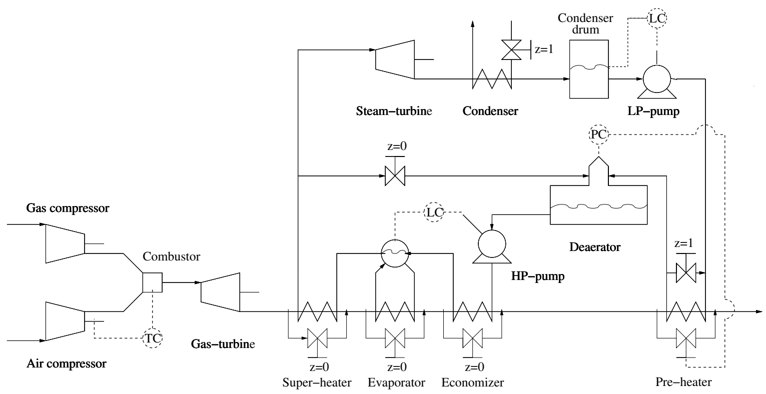

Figure 5: Combined cycle power plant process with proposed control structure for case I

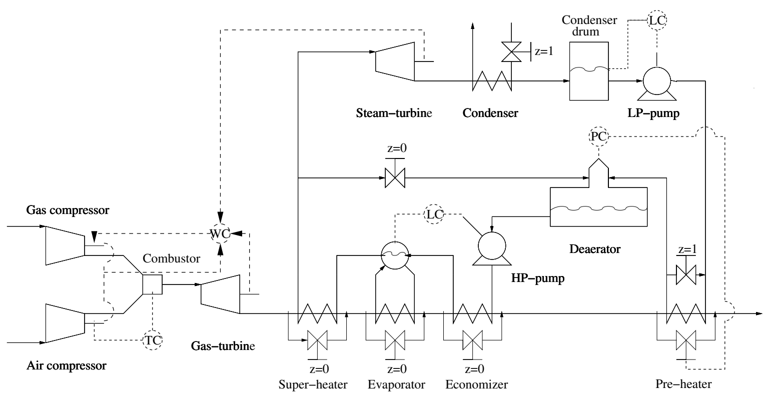

Figure 6: Combined cycle power plant process with proposed control structure for case II

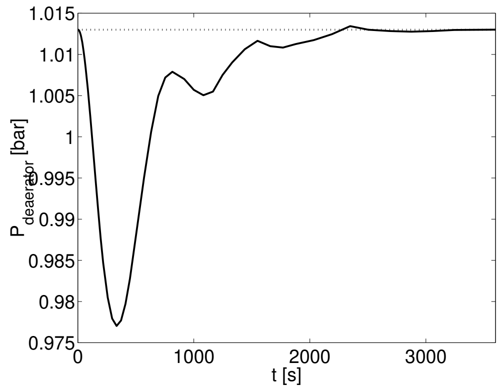

Figure 7: The deaerator pressure response when increasing the feed air temperature with 10^oC

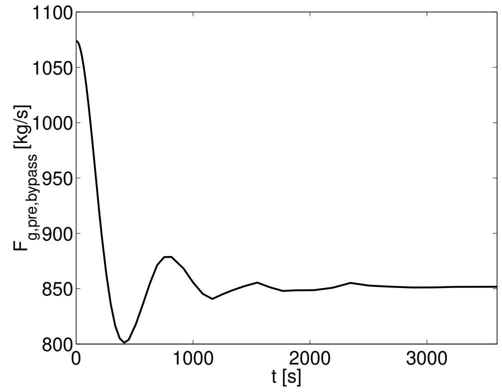

Figure 8: The pre-heater bypass flowrate response when increasing the feed air temperature with 10^oC

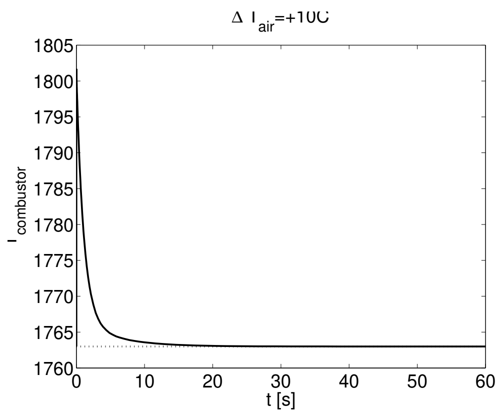

Figure 9: The combustor temperature response when increasing the feed air temperature with 10^oC

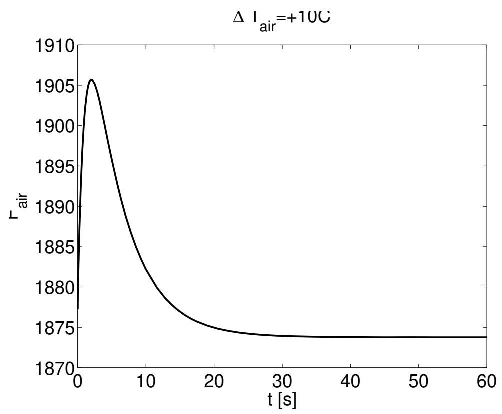

Figure 10: The feed air flowrate response when increasing the feed air temperature with 10^oC

References

Bolland, O. (1989). Termiske kraftstasjoner. dampturbiner og kjeler.

Gronnaass, L (2001). Dynamisk modellering og reguleringsstruktur av gasskraftverk. Technical report. NTNU.

Kehlhofer, R (1991). *Combined-Cycle Gas and Steam Turbine Power Plants.* Chap. 8. The Fairmont Press Inc.

Larsson, T and S Skogestad (2001). Plantwide control: A review and a new design procedure. *Modeling, Identification and Control* **22**(1), 1–32.

Norum, L. (2002). Personal communication.

Table 6: Candidate controlled variables (c_2) ranked by the (scaled) steady-state gain ($|G(0)|$).

Rank	c_2	$\|G(0)\|$
1	$F_{g,pre}$	-1.5391
2	$F_{g,pre,bypass}$	3.8911
3	$F_{g,econ}$	2.2931
4	$F_{g,econ,bypass}$	0.0000
5	$F_{g,evap}$	2.2931
6	$F_{g,evap,bypass}$	0.0000
7	$F_{g,super}$	2.2931
8	$F_{g,super,bypass}$	0.0000
9	$F_{g,gasturbin}$	2.2931
10	$F_{g,air}$	2.3358
11	$F_{g,gas}$	0.9587
12	$F_{w,pump,LP}$	-1.2288
13	$F_{w,valve,LP}$	0.0000
14	$F_{w,pre}$	-0.0207
15	$F_{w,valve,HP}$	0.0000
16	$F_{w,pump,HP}$	-1.2288
17	$F_{w,pump,HP}$	-1.2288
18	$F_{w,evap}$	-1.2288
19	$F_{w,super}$	-1.2288
20	$F_{w,steamturbine}$	-1.2288
21	$P_{g,comb}$	5.4713
22	$P_{w,cond}$	-3.6352
24	$P_{w,evap}$	-4.6744
26	$T_{g,super,1}$	-7.6936
27	$T_{g,super,2}$	-7.6898
28	$T_{g,super,3}$	-7.6839
29	$T_{g,super,4}$	-7.6748
30	$T_{g,super,5}$	-7.6613
31	$T_{g,super,6}$	-7.6417
32	$T_{g,super,7}$	-7.6148
33	$T_{g,evap,1}$	-7.5419
34	$T_{g,evap,2}$	-7.4616
35	$T_{g,evap,3}$	-7.3748
36	$T_{g,evap,4}$	-7.2828
37	$T_{g,evap,5}$	-7.1868
38	$T_{g,evap,6}$	-7.0886
39	$T_{g,evap,7}$	-6.9897
40	$T_{g,econ,1}$	-6.8679
41	$T_{g,econ,2}$	-6.6881
42	$T_{g,econ,3}$	-6.4145
43	$T_{g,econ,4}$	-5.9528
44	$T_{g,econ,5}$	-5.0688
45	$T_{g,econ,6}$	-3.5759
46	$T_{g,econ,7}$	-0.6635
47	$T_{g,pre,1}$	-1.0101
48	$T_{g,pre,2}$	-1.3670
49	$T_{g,pre,3}$	-1.7340
50	$T_{g,pre,4}$	-2.1113
51	$T_{g,pre,5}$	-2.4641
52	$T_{g,pre,6}$	-2.6817
53	$T_{g,pre,7}$	-2.8786
54	$T_{w,cond}$	-1.5724
55	$T_{w,pre,1}$	0.0000
56	$T_{w,pre,2}$	0.0187
57	$T_{w,pre,3}$	0.0373
58	$T_{w,pre,4}$	0.0558
59	$T_{w,pre,5}$	0.0741
60	$T_{w,pre,6}$	0.0922

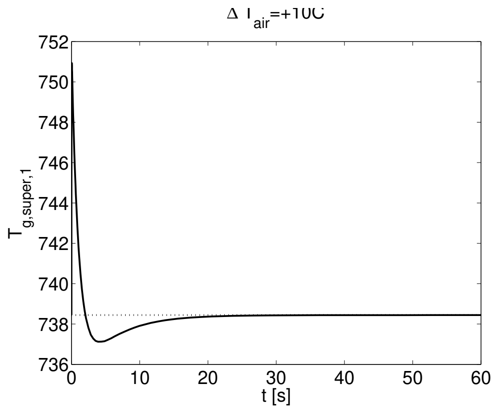

Figure 11: The super-heater gas inlet temperature response when increasing the feed air temperature with 10^oC

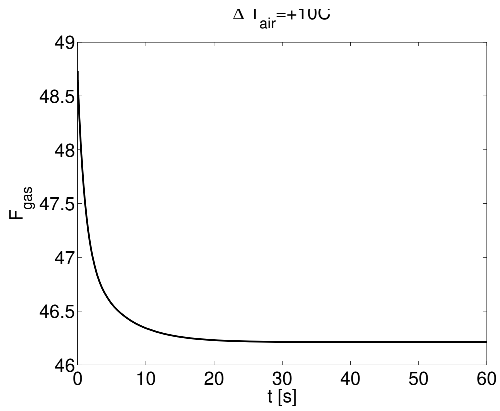

Figure 12: The feed air flowrate response when increasing the feed air temperature with 10^oC

Table 7: Candidate controlled variables (c_2) ranked by the (scaled) steady-state gain ($|G(0)|$).

Rank	c_2	$\|G(0)\|$
61	$T_{w,pre,6}$	0.1102
62	$T_{w,deaerator}$	0.0000
63	$T_{w,econ,1}$	-2.0919
64	$T_{w,econ,2}$	2.2232
65	$T_{w,econ,3}$	4.2523
66	$T_{w,econ,4}$	5.3397
67	$T_{w,econ,5}$	5.9329
68	$T_{w,econ,6}$	6.1383
69	$T_{w,econ,7}$	5.6884
70	$T_{w,evap}$	-5.9777
71	$T_{w,super,1}$	-7.6405
72	$T_{w,super,2}$	-7.6035
73	$T_{w,super,3}$	-7.5437
74	$T_{w,super,4}$	-7.4464
75	$T_{w,super,5}$	-7.2858
76	$T_{w,super,6}$	-7.0168
77	$T_{w,super,7}$	-6.5726
78	Q_{cond}	-0.4525
79	Q_{pre}	-0.3704
80	Q_{econ}	-0.4638
81	Q_{evap}	-0.3620
82	Q_{super}	-1.3846
83	$W_{comp,gas}$	1.4496
84	$W_{comp,air}$	1.4687
85	$W_{pump,LP}$	-0.4347
86	$W_{pump,HP}$	-0.9494
87	$W_{gas,turbine}$	1.0653
88	$W_{steam,turbine}$	-0.6568
89	$N_{gas,dim}$	0.9587
90	$N_{air,dim}$	2.6644
91	$F_{g,gas}/F_{g,air}$	-0.6211

Table 8: Average objective value and loss for different candidate controlled variables (c) when keeping nominal optimal setpoints (c_s) for case III.

Rank	c_1	c_2	$c_{s,1}$	$c_{s,2}$	$J(d)$	$J(e_1)$	$J(e_2)$	$J_{average}$	Loss (%)
opt	–	–	–	–	70.1595	70.0590	70.0590	70.0877	–
online	–	–	–	–	68.3165	68.5945	68.3339	68.4034	−2.40
1	$T_{g,comb}$	$F_{g,pre}$	1490	844.2354	68.2568	68.1975	63.2473	66.8196	−4.66
2	$T_{g,comb}$	$F_{g,pre,bypass}$	1490	1081.7586	68.2186	68.3349	68.0055	68.2074	−2.68
3	$T_{g,comb}$	$F_{g,econ}$	1490	1925.9941	68.2082	68.3351	66.6735	67.8239	−3.23
5	$T_{g,comb}$	$F_{g,evap}$	1490	1925.9941	68.2082	68.3351	66.6735	67.8239	−3.23
7	$T_{g,comb}$	$F_{g,super}$	1490	1925.9941	68.2082	68.3351	66.6735	67.8239	−3.23
9	$T_{g,comb}$	$F_{g,gasturbin}$	1490	1925.9941	68.2082	68.3351	66.6735	67.8239	−3.23
10	$T_{g,comb}$	$F_{g,air}$	1490	1877.2604	68.2237	68.3354	66.7168	67.8408	−3.21
12	$T_{g,comb}$	$F_{w,pumpLP}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
16	$T_{g,comb}$	$F_{w,pumpHP}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
17	$T_{g,comb}$	$F_{w,pumpHP}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
18	$T_{g,comb}$	$F_{w,evap}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
19	$T_{g,comb}$	$F_{w,super}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
20	$T_{g,comb}$	$F_{w,steamturbin}$	1490	148.3785	68.2875	68.0414	59.4146	65.6887	−6.28
21	$T_{g,comb}$	$P_{g,comb}$	1490	4289495.9378	68.2073	68.3307	68.2309	68.2674	−2.60
22	$T_{g,comb}$	$P_{w,cond}$	1490	2758.3129	68.2853	68.0591	67.5059	68.0049	−2.97
24	$T_{g,comb}$	$P_{w,evap}$	1490	18443043.2560	68.2795	68.1027	68.0009	68.1572	−2.75
26	$T_{g,comb}$	$T_{g,super,1}$	1490	738.4425	68.2449	68.2798	68.3286	68.2915	−2.56
27	$T_{g,comb}$	$T_{g,super,2}$	1490	738.1387	68.2450	68.2795	68.3285	68.2914	−2.56
28	$T_{g,comb}$	$T_{g,super,3}$	1490	737.6457	68.2451	68.2791	68.3284	68.2913	−2.56
29	$T_{g,comb}$	$T_{g,super,4}$	1490	736.8455	68.2453	68.2785	68.3282	68.2911	−2.56
30	$T_{g,comb}$	$T_{g,super,5}$	1490	735.5469	68.2456	68.2775	68.3280	68.2909	−2.56
31	$T_{g,comb}$	$T_{g,super,6}$	1490	733.4393	68.2462	68.2758	68.3276	68.2904	−2.56
32	$T_{g,comb}$	$T_{g,super,7}$	1490	730.0188	68.2471	68.2732	68.3270	68.2898	−2.57
33	$T_{g,comb}$	$T_{g,evap,1}$	1490	718.6409	68.2504	68.2636	68.3252	68.2875	−2.57
34	$T_{g,comb}$	$T_{g,evap,2}$	1490	708.5846	68.2537	68.2529	68.3231	68.2848	−2.57
35	$T_{g,comb}$	$T_{g,evap,3}$	1490	699.6965	68.2571	68.2413	68.3205	68.2817	−2.58
36	$T_{g,comb}$	$T_{g,evap,4}$	1490	691.8407	68.2604	68.2287	68.3176	68.2782	−2.58
37	$T_{g,comb}$	$T_{g,evap,5}$	1490	684.8974	68.2637	68.2153	68.3143	68.2743	−2.59
38	$T_{g,comb}$	$T_{g,evap,6}$	1490	678.7606	68.2668	68.2014	68.3106	68.2702	−2.59
39	$T_{g,comb}$	$T_{g,evap,7}$	1490	673.3367	68.2697	68.1871	68.3066	68.2658	−2.60
40	$T_{g,comb}$	$T_{g,econ,1}$	1490	671.0885	68.2729	68.1712	68.3010	68.2606	−2.61
41	$T_{g,comb}$	$T_{g,econ,2}$	1490	667.9811	68.2777	68.1454	68.2915	68.2519	−2.62
42	$T_{g,comb}$	$T_{g,econ,3}$	1490	663.6861	68.2852	68.1001	68.2740	68.2361	−2.64
43	$T_{g,comb}$	$T_{g,econ,4}$	1490	657.7496	68.2982	68.0078	68.2357	68.2025	−2.69
44	$T_{g,comb}$	$T_{g,econ,5}$	1490	649.5443	68.3234	67.7602	68.1263	68.1077	−2.83
65	$T_{g,comb}$	$T_{w,econ,}$	1490	619.4084	68.1701	68.2920	68.1545	68.2239	−2.66
66	$T_{g,comb}$	$T_{w,econ,}$	1490	596.5496	68.2020	68.3360	68.2698	68.2785	−2.58
67	$T_{g,comb}$	$T_{w,econ,}$	1490	564.9546	68.2148	68.3350	68.2961	68.2894	−2.57
68	$T_{g,comb}$	$T_{w,econ,}$	1490	521.2845	68.2213	68.3302	68.3007	68.2912	−2.56
69	$T_{g,comb}$	$T_{w,econ,}$	1490	460.9243	68.2246	68.3266	68.2757	68.2840	−2.57
70	$T_{g,comb}$	$T_{w,evap}$	1490	632.0651	68.2795	68.1027	68.2567	68.2303	−2.65
71	$T_{g,comb}$	$T_{w,super}$	1490	735.3269	68.2453	68.2779	68.3277	68.2908	−2.56
72	$T_{g,comb}$	$T_{w,super}$	1490	733.0823	68.2456	68.2765	68.3271	68.2903	−2.56
73	$T_{g,comb}$	$T_{w,super}$	1490	729.4393	68.2462	68.2742	68.3261	68.2896	−2.57
74	$T_{g,comb}$	$T_{w,super}$	1490	723.5270	68.2472	68.2702	68.3244	68.2882	−2.57
75	$T_{g,comb}$	$T_{w,super}$	1490	713.9315	68.2491	68.2628	68.3211	68.2857	−2.57
76	$T_{g,comb}$	$T_{w,super}$	1490	698.3585	68.2528	68.2471	68.3143	68.2803	−2.58
77	$T_{g,comb}$	$T_{w,super}$	1490	673.0842	68.2611	68.2085	68.2975	68.2669	−2.60
82	$T_{g,comb}$	Q_{super}	1490	28559883.6519	68.2478	68.2695	60.8696	66.1582	−5.61
83	$T_{g,comb}$	$W_{compgas}$	1490	7647527.3701	67.7918	68.3146	64.0736	66.9563	−4.47
84	$T_{g,comb}$	$W_{compair}$	1490	2399731637.6085	68.4136	68.3343	61.7553	66.4772	−5.15
86	$T_{g,comb}$	W_{pumpHP}	1490	2721519.7459	68.2831	68.0764	50.0754	63.0291	−10.07
87	$T_{g,comb}$	$W_{gasturbin}$	1490	3397557688.3306	68.2031	68.3246	56.7562	64.9860	−7.28
90	$T_{g,comb}$	$N_{air,dim}$	1490	759.1992	68.3823	68.3354	66.7168	67.8861	−3.14

Saue, N.E. (2002). Selvoptimaliserende regulering av gasskraftverk. Master's thesis. NTNU.

Skogestad, S. (2001). Probably the best simple pid tuning rules in the world.