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Multivariable Control of a Pneumatic Conveying System

Wolfgang Birk

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Multivariable Control of a Pneumatic Conveying System

by

Wolfgang Birk

Control Engineering Group
Department of Computer Science and Electrical Engineering
Luleå University of Technology
S-971 87 Luleå, Sweden

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Supervisor
Professor Alexander Medvedev,
Luleå University of Technology, Sweden

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To Ulrika and
my parents

Abstract

This licentiate thesis deals with a multivariable controller of a pneumatic conveying system. The industrial process used to illustrate design, analysis and implementation of a multivariable controller is the injection of fine coal into a blast furnace.

In the blast furnace process, coke is replaced by fine coal for economical and environmental reasons. The coal mass flow to the blast furnace becomes a crucial parameter for its operation and hence a control system should be in place to maintain a constant flow.

An optimal multivariable linear-quadratic controller has been designed in order to replace the conventional PI-controllers, which were not able to achieve the desired control objectives. The design is analyzed by deriving the sensitivity functions of the closed loop system to noise, disturbances and uncertainties.

The multivariable controller is then validated through experiments and test operation at the coal injection plant at SSAB Tunnplåt AB in Luleå, Sweden. The reached performance improvements compared with the conventional PI-controllers can be up to 80%.

Finally, the controller is combined with a gas-leakage detection system to make up the commercially available product SafePCI, which is now installed and in permanent operation at SSAB Tunnplåt AB in Luleå, Sweden.

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Preface

In writing this thesis, the author's aim has been to summarize research work that has been conducted at the Control Engineering Group at Luleå University of Technology in Sweden, under the supervision of Professor Alexander Medvedev.

The presented results are application oriented and the work has been carried out in cooperation with SSAB Tunnplåt AB in Luleå, Sweden. The motivation for the research is twofold. On the one hand our industrial partner's need for improved control of the processes and, on the other hand the need to gain more insights into multivariable control of industrial processes, *i.e.* pneumatic conveying of solids. Hence, unique conditions have been created. Researchers have the opportunity to perform experiments at industrial plants with willing support of the personnel at the plant as well as the possibility of exchanging practical know-how from the industry with theoretical knowledge from the university.

Multivariable control of a pneumatic conveying system has been embedded as a sub-project of the project *Reliable Process Control* which is conducted by the Center for Process and System Automation (ProSA), at Luleå University of Technology. The project was financed by grants from Norrbottens Research Council. *Reliable Process Control* has not only dealt with control aspects but also with fault detection and isolation. Since the application in both cases was one and the same, a cooperation between Andreas Johansson, responsible for the fault detection and isolation part, and the author was a natural way to perform experiments at the industrial plant. Furthermore, control and fault detection could be combined to create a more advanced control strategy, which pursues control, fault diagnosis, monitoring and supervision at the same time.

The thesis is organized in two parts, an introduction and published papers.

Briefly described, the introduction will give an overview of linear multivariable control of industrial processes and pneumatic conveying. Furthermore, it links both areas together by discussing the necessity of automatic control in pneumatic conveying, especially multivariable control. The second part consists of the published papers, of which the following are included:

Paper 1. W. Birk and A. Medvedev, "Pressure and Flow Control of a Pulverized Coal Injection Vessel," in *Proc. of the 1997 IEEE International Conference on Control Applications in Hartford, Connecticut USA, October 5-7*, pp. 127–132, 1997.

Paper 2. W. Birk, A. Johansson, and A. Medvedev, "Model-based Control for a Fine Coal Injection Plant," *IEEE Control Systems Magazine*, vol. 19, pp. 33–43, February 1999.

Paper 3. W. Birk and A. Medvedev, "Sensitivity Analysis of an LQ Optimal Multivariable Controller for a Fine Coal Injection Vessel," presented at the *1999 IEEE Industrial Applications Society 34th Annual Meeting in Phoenix, Arizona USA, October 3-7*, 1999.

The following two published papers are not included:

Paper 4. W. Birk, A. Johansson and A. Medvedev, "Control and Gas Leakage Detection in a Fine Coal Injection Plant: Design and Experiments," in *Proc. of 9th IFAC Symposium on Automation in Mining, Mineral and Metal Processes in Cologne, Germany, September 1-3*, pp. 271–276, 1998.

Paper 5. A. Johansson, W. Birk and A. Medvedev, "Model-based Gas Leakage Detection and Isolation in a Pressurized System via Laguerre Spectrum Analysis," in *Proc. of the 1998 IEEE International Conference on Control Applications in Trieste, Italy, September 1-4*, pp. 212–216 , 1998.

The author's contribution to the fault detection parts in the papers is restricted to the application of the work and not the theoretical part, which has been completely conducted by the co-authors.

WOLFGANG BIRK
Luleå, August 1999

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Many people have influenced this work, directly or indirectly. First of all, I want to thank my supervisor Prof. Alexander Medvedev for his skilful guidance, enthusiasm and valuable criticism of my work over the last few years, not to mention his vast amount of knowledge. I would also like to thank Andreas Johansson with whom I have been working together in the research project *Reliable process control* and who was frequently a sounding board for me. Thanks to all the members of the Control Engineering Group for creating an inspiring working atmosphere, as well as for their help, especially Britta Fischer for all the pleasant discussions.

I want to express my gratitude to Robert Johansson, Bert Paavola, Egon Hilberg and Mikael Ullenius from SSAB Tunnplåt AB in Luleå, for all their help and support during the project. Furthermore, and very sadly, I bid farewell to Ingemar Lundström, who passed away shortly after finishing the project. He will be remembered for the time we have spent together during the experiments and tests at the plant and for his help.

Finally, a special thank to my fiancee Ulrika for her never ending support and encouragement.

Introduction

In this introduction, the basic concepts of pneumatic conveying of solids are described and an overview of multivariable control is given. Furthermore, the use of multivariable control in pulverized coal injection, a special case of pneumatic conveying of solids, is motivated. Finally, further references for the interested reader are given.

1 Pneumatic Conveying

Transportation of solids, mostly powdered or granular, by means of a gas stream is called pneumatic conveying. In [KMRL97] and [Wöh86] an overview and thorough description of this technology can be found. The realisation of such a transportation mechanism is complex and power consuming. Furthermore, one might think that a conveyor belt could do the same job. However, there are circumstances where conveying with a traditional conveyor belt is not applicable, *e.g.* to provide dust-free transport or transport of flammable material.

For a solids transportation system to qualify as a pneumatic conveying system, four zones need to be present:

- Prime mover zone. The pressure difference needed for a reliable transportation of the solids is created here. It is usually the source of the transport gas flow.

- Feeding, mixing and acceleration zone. The solids have to be introduced into the transport gas flow and a two phase mixture has to be created.

- Conveying zone. The solids are moved together with the transport gas as a two phase flow from the source to the destination.

- Gas-solids separation zone. Here, the two phases are separated by e.g. a filter.

The above specifications can be applied to a wide range of transportation systems, why a more precise classification is needed. The following classification factors appear to be reasonable:

Solid type. In general, the material can be lump, granular or powdered, that means the grain size differs.

Material. Each material can have different conveying properties, e.g. coal powder and metal oxide fines do not need to have the same properties.

Conveying pressure. The solid transport is achieved by a pressure difference and the conveying source needs a higher pressure level than the conveying destination. The pressure level can be high or low and also the pressure at the destination can be negative or positive.

Conveying phase. The solids are conveyed through a pipe. The involved two phase flow can have different characteristics, mainly in respect to the mass flow ratio, solid mass to gas mass. If the ratio exceeds 15, the flow is called a dense phase flow, and a dilute phase flow otherwise.

A very important aspect in pneumatic conveying is fluidisation, an often misunderstood term. Fluidisation of solids means a reduction of the angle of repose and not that solid material will get fluid properties, [Wil83]. Everybody is aware of the fact that a hovercraft is only able to move because of the low friction between the air-cushion and the ground/craft. In respect to solid particles, the presence of the fluidisation gas has the same effect as the above mentioned air-cushion. The friction between each solid particle and its surrounding is reduced, which yields an increased mobility.

2 Pulverized Coal Injection

In the blast furnace process coke is replaced by fine coal for economical and environmental reasons, [SCC$^+$93]. The fine coal is injected at tuyere level into the blast furnace, making the process very sensitive to coal flow outages. One advantage of pulverized coal injection is that it yields a shorter reaction time in the blast furnace process. Other effects of pulverized injection and a projection of its future use are discussed in [Ame98].

Pulverized coal injection is a form of pneumatic injection which is a special case of pneumatic conveying, see [Sno81]. Instead of separating the solids from the gas in a filter and finally collecting the particles in a silo, the solids are injected in the reactive zone of a process. Usually, the solids are conveyed in the dense phase and the used systems are positive pressure systems with a very high pressure.

As the name of the section already reveals, the conveyed solid is coal in pulverized form. The pulverized coal is conveyed from the injection vessel through an injection pipe. Later, the flow is distributed to several smaller injection lines. And finally the injection lines end up in the tuyeres of a blast furnace where the pulverized coal is injected in the raceway of a blast furnace. The gas-solid separation is achieved through a property of gas-solids jets, as the two phases will separate automatically [SS78].

As can be easily imagined, the dynamical behaviour of the flow from the injection vessel to the blast furnace is very complex. First, there is a two phase flow (gas/solids) through a long and not straight pipe. Then, there is the distributor, where the two-phase flow should be evenly divided. And finally, the injection, which is dynamically described by the behaviour of a two-phase jet. Consequently, very little can be found in the literature about the dynamics of these systems.

The continuity of the pulverized coal mass flow to the blast furnace is very important for blast furnace operation, when pulverized coal injection is applied, [ER95]. Using high coal injection rates the stability of a blast furnace is considerably dependent on the coal mass flow. As a result, the coal mass flow should vary very little and furthermore should be reliable. Thus, there is a high demand for the coal mass flow control. In the sequel it is shown that the use of a multivariable control scheme yields a much better control of the coal mass flow to the blast furnace.

3 Multivariable Industrial Control

For a long time controllers were used in the industry to automate processes and to improve product quality and reduce production costs. Probably one of the most known and most used controller types is the PID controller. Usually it is used to control sub-processes of industrial processes in so-called single loops, where the individual controller reacts with individual actuator movements on deviations from a given set-point or trajectory. This traditional control concept and its design are well described in the literature, see *e.g.*

[FPEN94], [ÅW90], [FPW98], [DFT92] or [DB98], and it is a tool in every control engineer's tool-box.

Since each control loop only deals with one process variable, such a decentralized structure has its pros and cons. One of many advantages is that the PID controller seems to be easy to maintain. But there is also a major drawback. Many of these individual loops disturb each other, preventing the affected loop to achieve its objective. Furthermore, the structure offers only one degree of freedom.

An obvious alternative is a more centralized structure with multiple inputs and multiple outputs (MIMO) which can handle several process variables at a time, a so-called multivariable controller. A direct result is the increase in the number of degrees of freedom. The controller can use multiple actuators to achieve one control objective. This, of course, depends on the structure of the industrial process, as several actuators and sensors have to be present. Consequently, the design of multivariable controllers depends even more on the characteristics of the plant than in the scalar case, [SP96]. Moreover, the design procedure is more complex which is reflected into the maintenance of a multivariable control system. Thus, multivariable control concepts are not found that often in industry.

Still some examples of multivariable control in industry can be found: *e.g.* vibration suppression [DK96], control of plate-like structures [LEL96] and control of vapor compression [HALI98]. In each of the three examples a multivariable linear quadratic gaussian (LQG) controller is used.

3.1 Industrial Control Systems

Processes in industry are usually controlled by a so called programmable logic controller (PLC). PLCs provide the user with basic function blocks for data processing, storage, exchange as well as decision and control blocks. Mostly, the function block for control is a PID controller block, which can be configured into different modes: P, PI, PD and PID. Often the available function blocks are not sufficient for implementation of a multivariable controller and in addition older PLCs are mostly running at their computational capacity limits.

But over the years PLCs have become more and more sophisticated as processor power has increased radically. Thus, more process control loops with short cycle times and more complex computations can be handled. All manufacturers have their own concepts of providing the user with advanced features. Two examples of commonly available systems are given here: Honeywell-

Measurex Totalplant AlCont and Allen Bradley's PLC-5.

Honeywell-Measurex Totalplant AlCont provides the engineer with a standard function block set, from which new blocks with more functionality can be created. Additionally, function blocks can be directly generated from Pascal-code with freely definable input/output interface. These so-called Pascal-blocks are treated in the same way as the standard function blocks by the PLC. The engineer can simulate and test the blocks in an off-line environment and later simply download the newly created blocks to the PLC. Finally, signal information in all blocks of the application can be monitored on-line from a Windows NT environment.

Allen Bradley's PLC-5 can be expanded with a control coprocessor. The coprocessor programs can be developed in BASIC, C or assembler languages and will later run in the coprocessor as an asynchronous process. In contrast to the above mentioned system, the developed algorithms are not integrated in the PLC standard function blocks and synchronization with the main processor has to be realized by the programmer. An integrated development environment (IDE) is used to create the program, debug it and finally download it to the coprocessor.

The availability of such advanced features facilitates the design, implementation and maintenance of multivariable controllers for industrial processes. Thereby, the time period from controller design to industrial product shrinks and the product costs are reduced. As a result the control and gas-leakage detection system *SafePCI* has been successfully installed in Honeywell-Measurex Totalplant AlCont.

3.2 Design Strategies

In the literature one can find a wide range of design strategies for multivariable controllers, see e.g. [SP96], [Zho96] or [Var91]. Practically all depend on the existence of a parametrical linear multivariable process model, which can be given in the transfer function form by

$$y(s) = G(s)u(s)$$

where $y(s)$ is an $l \times 1$ output vector, $u(s)$ is an $m \times 1$ input vector and $G(s)$ is an $l \times m$ transfer function matrix.

Based on the process model, different methodologies are pursued to design a multivariable controller, four are discussed in more detail:

- Decentralized control
- Decoupling control
- Optimal control
- Robust control

Generally, these methods can be divided in two groups. The first two methods reduce the multivariable control problem to several scalar control problems, where traditional control theory can be applied. Consequently, two steps are involved: problem reconfiguration and controller design. When using optimal or robust design method for the controller synthesis, the multivariable controller is directly obtained from the multivariable control problem.

In decentralized control, it is assumed that the multivariable process has a diagonal structure ($G(s)$ diagonal). Thus, input u_i only interacts with output y_i. Consequently, the multivariable process is a collection of several independent scalar processes, for which scalar controllers can be designed. In the reconfiguration step, the input/output pairs for the controller design are selected. If the assumption of a diagonal structure is not fulfilled, the off-diagonal elements in the process have to be considered.

The concept of decoupling control is based on the idea of minimizing the influence of the off-diagonal elements in $G(s)$, a so-called decoupling. The result is a modified multivariable process $G'(s)$ that is a collection of several independent scalar processes. To achieve the decoupling, a network around $G(s)$ is designed, which minimizes the influences of the off-diagonal elements. In the simplest structure case, the decoupling network is a transfer function matrix of dimension $l \times l$ or $m \times m$, when connected to the input or the output of $G(s)$, respectively. Then, for each scalar process in $G'(s)$ a scalar controller is designed. Consequently, the control performance depends on the success of the decoupling step. If the process model is not exact, which can be a result of modelling uncertainties or non-linearities, the decoupling can only be achieved under certain circumstances or not at all. Hence, design strategies that can handle multivariable processes have to be chosen.

In optimal and robust control, the multivariable nature of the process can be taken into consideration. The main issue of optimal control is disturbance rejection where the problem is recast into the minimization of a cost or loss function. The controller is then obtained from an optimization of a quadratic criterion. In the linear quadratic (LQ) and LQG problem the issue is to minimize the variance for a signal in the loop, *e.g.* the control error. Even here the controller synthesis is based on a process model, and thus the control performance depends on the accuracy of the model. Furthermore, optimal controller designs often create sensitivity peaks, which can lead to loss of stability under plant perturbation. In order to solve this problem, uncertainties have to be taken into account, and this approach leads to robust control.

Instead of focusing on control performance alone, in robust control the control performance and loop stability under plant perturbation is the main issue. The involved uncertainties are modelled by hard upper bounds for system norms, which are chosen conservatively. Consequently, the performance of robust designs is usually poor compared with optimal designs, as it is a result of a trade-off between performance and guaranteed stability margins.

In this thesis the third approach is used for the controller synthesis, the so-called LQG design. The controller is an optimal state feedback, where the state information is provided by an optimal state estimator. Because of the above mentioned difficulties with the approach, the sensitivity of the resulting closed loop has been borne in mind.

4 Contribution

Up to now the control of the coal mass flow from the injection vessels to the blast furnace has been a scalar control. A problem with scalar control is the always present high variations in the coal mass flow, resulting in a limitation of the maximum secure injection rates. The coal mass flow is not only intended to be an additional cheap fuel source but also a control parameter with a small reaction time in the blast furnace process. But, because of the variations in the coal mass flow, this parameter loses a lot of its potential effect.

The introduction of multivariable control for the pulverized coal injection vessels, used at SSAB Tunnplåt AB in Luleå, led to a significant reduction of the variations in the coal mass flow to the blast furnace. The designed controller is now installed at the plant and in permanent operation.

The contribution of the author is design and analysis of multivariable controllers for a class of pneumatic conveying systems. The LQG design methodology is applied. Analysis is first performed in a more traditional manner with simulation studies and later in the form of sensitivity analysis in the multivariable case. The work involved with implementation of the multivariable controller in an industrial environment for experiment and test operation is only briefly taken up by the included papers, primarily the safety aspects are derived and presented.

The modelling of a pneumatic conveying system is mainly based on results of the project group, both from system identification and modelling based on the laws of physics.

5 Future Research

- **Limitations in multivariable control system**
  As already described in [SBG97] and [Che95], there exist limitations in multivariable control system. The limiting influence on multivariable control in industrial processes should be investigated, and it should also be compared with the effect of these limitations on scalar control.

- **Comparison of different design strategies**
  It should be analysed how effective different design strategies are at producing reliable multivariable controllers and to what extent the design process can be automated. Nowadays, the design of multivariable controllers is an extensive process and thus, more handy tools for the design are needed. Even maintenance of multivariable controllers has to be taken into consideration. It should be studied which design strategies comply best.

- **Parameter dependent multivariable control**
  Most industrial processes are non-linear and can only be approximated by linear multivariable models at certain equilibrium points. Using parameter dependent linear multivariable models, these approximations can be valid for wider ranges around these equilibrium points. The effects on controller design and their application to industrial processes should be investigated.

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Papers

Paper 1

Pressure and Flow Control of a Pulverized Coal Injection Vessel

Published as

W. Birk and A. Medvedev, “Pressure and Flow Control of a Pulverized Coal Injection Vessel,” in *Proc. of the 1997 IEEE International Conference on Control Applications in Hartford, Connecticut USA, October 5-7*, pp. 127–132, 1997.

Pressure and Flow Control of a Pulverized Coal Injection Vessel

Wolfgang Birk and Alexander Medvedev
Control Engineering Group
Luleå University of Technology
97187 Luleå, SWEDEN
wolfgang.birk@sm.luth.se

Abstract

This paper deals with model-based pressure and flow control of a fine coal injection vessel for the use of the blast furnace injection process.

By means of system modeling and identification, the structure and behavior of the coal injection vessel are analyzed. It is shown that by use of model-based design, the control goals can be reached and the control performance can be significantly improved compared to the conventional PI-controllers. Several dynamic models of the plant are developed. A number of control strategies are presented and compared by means of practical tests. The LQG design method is used to design the controllers. All the controllers are validated through experiments on the coal injection plant at SSAB Tunnplåt in Luleå, Sweden.

1 Introduction and Background

Iron is a mass product, produced in blast furnaces and later refined to steel. This could lead to the conclusion that the process is efficient and optimized, which is not completely true. It has become more usual to bring down the cost factors in the iron production, by reducing the share of undesired by-products or the costs of energy supply.

Although coke is one of the most expensive energy carriers, it is common to use it in iron production. In Luleå , SSAB Tunnplåt reduces the production costs by partly substituting coke and using pulverized coal instead. Pulverized coal is about 40% cheaper than coke, which fact makes it very attractive.

Since pulverized coal, in its pure form, is highly inflammable even under normal conditions, it is difficult to supply it to the process. Therefore, it is important to keep the pulverized coal isolated from the air, which can be done by using a pneumatic conveying device. The used coal injection plant is planed, designed and constructed by BMH (Babcock Materials Handling in Hamburg, Germany), where two injection vessels are used alternatively to maintain a continuous injection.

Injecting coal powder in the blast furnace at a high rate of about $190kg/thm$ makes the blast furnace process very sensitive. According to [1], the process outage becomes critical, due to the fact that large amounts of coal are not delivered to the furnace. The prime concern in fine coal injection is therefore a constant coal mass flow to the blast furnace.

Since the injection vessels operate under high pressure, pressure stabilization during the injection phase is also a control concern, but somewhat of lesser importance than mass flow control. In the experiments, it appears that a constant pressure in the injection vessel makes flow stabilization easier.

2 Coal injection plant

The coal injection plant is a highly automated plant, where incoming raw coal is stored, grinded, dried and finally injected into the blast furnace. During operation, human interaction is only needed for set point adjustments.

Fig. 1 shows the structure of the plant, where the different sections are marked and referred to by the capital letters in the marked area. The sections A, B, C are common for the two fine coal silos. Sections D, E and F, G belong to Blast Furnace 1 and Blast Furnace 2, respectively. This article deals only with the coal injection of Blast Furnace 2. Therefore, the emphasize is placed on section E.

A closer description of each section and its functionality can be found in [2].

2.1 Injection process

Principally, the injection process can be divided into two separate phases: a high-pressure and a low-pressure phase, where, as the names already imply, the pressure in the injection vessel is high or low, respectively.

One vessel is depressurized, charged and pressurized while the other vessel is injecting coal powder. To facilitate process identification and control, the high pressure and low pressure phases are sub-divided into more specific phases.

Fig. 2 shows one process cycle of injection vessel S21. A represents the low pressure phase and B to E belong to the high pressure phase. In Table 1, the nomenclature used in the sequel to refer to process phases is summarize.

In [2] the process phases are described more in detail. This paper only deals with the injection phase. Some particular properties, from which the primary control goals can be derived, are:

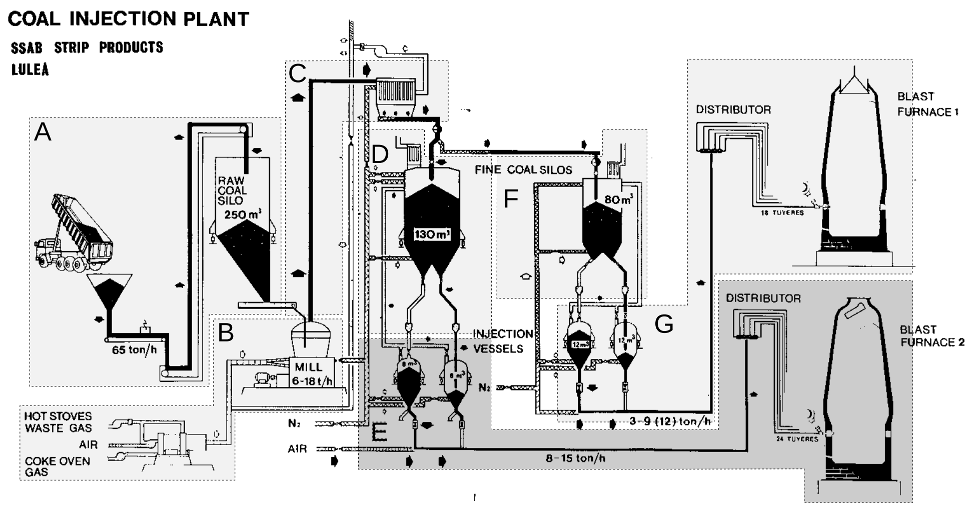

Fig. 1: Coal injection plant

Phase	Name	Description
A	Charging	The pressureless vessel is filled with coal powder
B	Pressurizing	The injection vessel is set under pressure
C	Pressure holding	Standby until the other vessel has finished injection
D	Injection	The coal powder is injected into the blast furnace
E	Ventilation	Depressurizing and ventilation of the vessel

Table 1: Process phases

- The pressure in the injection vessel is constant.
- The coal mass flow to the blast furnace is constant
- The mass of the injection vessel follows a ramp, where the slope of the ramp depends on the coal mass flow set-point.
- The injection phase ends when a minimum weight is reached.

3 Existing control unit

The control unit of the fine coal injection plant consists of two independent loops. One loop is for injection vessel pressure stabilization, and the other is for controlling the coal mass flow to the blast furnace. The process working cycle duration depends on the actual coal mass flow set point.

The pressure control loop is implemented by means of a PI controller, to achieve a zero steady state control error. The controller performance is exhibited in Fig. 3a. The coal mass flow controller is also a PI controller. However, since the coal mass flow is not directly measured in this particular installation, the coal mass flow is evaluated from the vessel's weighing system readings. The computations involve a differentiation, which naturally results in amplification of the measurement noise. Thus, low-pass filtering is applied and a set point compensation is employed to correct measurement errors.

As seen in Fig. 3b, the mass flow controller fails to drive the control error to zero during the injection phase. The shortcomings of the existing control strategy become even more prominent when the deviation of the mass loss in the injection vessel from the ideal mass loss is analyzed (Fig. 3c).

To recapitulate, under the existing control laws, the pressure p in the injection vessel is oscillative and the coal mass flow is not held constant. In the sequel, alternative ways of controlling the injection vessels are discussed.

4 Modeling

Basically, an injection vessel is a pressurized tank process (Fig. 4). There, two output signals are available:

- Pressure p in the vessel

- Coal mass m_C, which is the integral of the coal mass flow to the blast furnace. The coal mass flow is not measurable, as no flow meter is installed.

Natural control signals are:

- Valve opening u_N of the pressure control valve (PCV)
- Valve opening u_C of the flow control valve (FCV)

Since not all signals are suitable for process control, some assumptions are made, for the sake of model simplification:

1. Variations of the pressure in the nitrogen net are small and p_N is assumed to be constant.
2. The pressure p_T at the injection point in the transportation pipe is constant.
3. Temperature variations are small.
4. No nitrogen leakage from the injection vessel during the injection phase ($u_I = u_V = 0$, and these valves are tight)
5. The flow due to the weight of the coal is negligible.

A non-linear behavior of the injection vessel has been observed in [3], and a dynamic model based on the physical contiguities has been suggested.

In an earlier study [4], a black-box model of the injection vessel has been developed and successfully used for a pressure controller design. Following the recommendation worked out in the study, SSAB Tunnplåt in Luleå carried out constructive changes in the coal injection plant hardware. These changes in the equipment crave model validation to be performed anew.

According to the assumptions made, two input signals can be manipulated and two output signals measured, the opening of the valves u_N and u_C, the pressure p and the mass m. These signals are used to identify and to control the process.

4.1 Identification

When it comes to coal powder flow control, two additional models are needed to describe the process dynamic properties. One model whose output is the coal mass m_C and the input is the opening of the FCV u_C. The second model is a multiple input, multiple output (MIMO) model with p, m_C as outputs and u_N, u_C as inputs. The latter can be used to design a MIMO controller for the process.

Identification and validation data sets are logged on the process with a sampling time of $1s$. Exerting the least-squares method, a SISO model for the mass and a MIMO model for the mass and pressure are identified. The resulting models are presented in the ARX-form. Equation 1 gives the SISO model.

$$m(t) = \frac{b_{m1}q^{-1}}{1 - a_{m1}q^{-1}} \cdot u_C(t) \tag{1}$$

The MIMO model is defined by the polynomial matrices

$$\mathbf{A}(q^{-1}) \begin{bmatrix} p(t) \\ m(t) \end{bmatrix} = \mathbf{B}(q^{-1}) \begin{bmatrix} u_N(t) \\ u_C(t) \end{bmatrix} \tag{2}$$

with

$$\mathbf{A}(q^{-1}) = \mathbf{I} + \mathbf{A}_1 q^{-1} + \mathbf{A}_2 q^{-2} \tag{3}$$
$$\mathbf{B}(q^{-1}) = \mathbf{B}_1 q^{-1} + \mathbf{B}_2 q^{-2} \tag{4}$$

where $\mathbf{A}_1, \mathbf{A}_2, \mathbf{B}_1, \mathbf{B}_2$ are 2×2 matrices containing the polynom coefficients and $\mathbf{I}$ is the identity matrix.

Following the suggestions of [5, chapter 11], two validation tests are applied:

1. Common sense test
2. Statistical test

To spare place only the common sense validation plots are presented. Figure 7a,b shows the validation plot for the two SISO models and Fig. 7c,d for the MIMO model.

For controller design, the acquired matrix transfer functions are converted into state space models. Furthermore, the model order of the MIMO model has been reduced. The resulting state space equations for the SISO and MIMO models are presented below.

$$m(k+1) = \Phi_m m(k) - \Gamma_m u_C(k) \tag{5}$$

$$\begin{bmatrix} p(k+1) \\ m(k+1) \end{bmatrix} = \mathbf{\Phi} \begin{bmatrix} p(k) \\ m(k) \end{bmatrix} + \mathbf{\Gamma} \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} \tag{6}$$

In both cases the state vectors are transformed so that the states coincide with the outputs and, therefore, have a physical meaning.

5 Control strategies

As mentioned before, the control objectives are to maintain a constant pressure in the injection vessel and to guarantee a constant coal mass flow from the injection vessel to the blast furnace. Though, a steady coal mass flow has a higher priority than pressure stabilization. This can be achieved by pursuing different strategies. Here are two of them:

1. Two separate control loops for the pressure and the mass.
   One unit controls the vessel pressure and the other controls the coal powder mass, where the mass has to follow a pre-defined ramp. The slope of the ramp is the coal mass flow set point. For the controllers design, two SISO models for the plant are used.

2. A MIMO design controlling the pressure and mass.

Strategy 1 and 2 both have their own advantages and disadvantages.

Strategy 1 is based on SISO models which makes the controller design and implementation easy to handle. The main disadvantage is that the couplings in the process are not taken in account. To achieve a better control performance, the effects of the coupling between the injection vessel pressure and the fine coal flow have to be modeled as disturbances.

Strategy 2 takes the couplings into consideration. Furthermore, this strategy has one more degree of freedom, as the two actuators are used together to achieve the control objectives. Another advantage is that the controller can eventually be tuned so that the two loops work separately. In this case, a controller similar to that of Strategy 1 is obtained, and yet the couplings in the plant are accounted for through the model. A relative disadvantage is that the design process appears to be more complicated.

An optimal design method is used to design the controllers. The separate loops and the MIMO design are based on the so-called linear quadratic gaussian (LQG) theory, as described in [6]. It is also a major design tool for multivariable linear systems.

Normally, assuming state feedback, a proportional-integral control law is chosen to eliminate the steady state error. However, controlling the fine coal mass, one of the closed-loop system outputs has to follow a ramp. Then, the use of a single integrator leads to a constant steady state response error. Thus, a double integration is used to drive the error in reference signal following to zero. In order to recast this control law into the framework of LQG design, the process models have to be augmented with a double integrator.

For the resulting system, a steady state Kalman filter is designed, in order to obtain filtered versions of the measured signals. The steady state Kalman filter seems to be sufficiently fast, as the process itself is very slow.

Furthermore, to reduce the controller settling time and retain a smoother set point change response, a feedforward signal from the desired coal mass flow to the valves is introduced. The design ideology presented in [7, chapter 6] is adopted. The feedforward signal can be introduced both in the MIMO controller and the separate pressure and flow control loops.

Fig. 6 depicts two separate control loops with additional feedforward path and Fig. 5 shows a similar solution for the MIMO design.

6 Practical tests

In this section, two different tests are described. Firstly, the MIMO controller is tested and, secondly, a pressure control is run, with the existing controller for the mass flow. The latter test is important, as it demonstrates that a better pressure control leads to less variations in the mass flow, although the controller for the mass flow has not been changed. Furthermore, the couplings in the plant dynamics can be illustrated by this test too, since there would be no change in the control of the mass flow otherwise.

In the following four controllers are compared to each other :

1. Existing controller (PI). This controller is referred to as *old PI controller*.

2. Existing controller (PI, better tuned). In the course of project, the tuning of the PI controller has been improved. Naturally, this one is denoted as *new PI controller* in the comparative study.

	old PI	new PI	Combined	MIMO
mass $[kg]$	51.8	11.3	6.8	1.5
pressure $[kPa]$	5.1	5.1	1.0	1.0
mass flow $\left[\frac{t}{h}\right]$	1.8	1.6	0.9	1.0

Table 2: Standard deviations

	Improvement
mass	79%
pressure	86%
mass flow	35%

Table 3: Control performance improvements

3. Combined controller (LQG-design). The model-based controller from the preliminary study [4] is used to control the vessel pressure and the existing controller for the mass flow is used. All curves of this control solution are entitled *combined controller.*

4. MIMO controller (LQG-design). As this controller is completely model-based the curves are entitled *model-based controller.*

Fig. 8 depicts performance for each controller. As expected, the MIMO controller yields the best result. This can also be seen by comparing the pressure deviation, (Fig. 9).

From the experiments on the plant, it becomes clear that in the process there exist strong dynamic couplings that can not be neglected in the controller design. Although the mass flow control has not been changed, the overall controller performance becomes better because of a tighter pressure control loop.

Table 2 shows the standard deviations achieved by the corresponding controllers. Once again, it can be seen that the MIMO controller produces the best results.

Performance improvements achieved by the MIMO controller compared to the new PI controller are given in Table 3.

7 Conclusions

Identification and control of the coal injection process are discussed. It is shown that by use of model-based designs, the flow and pressure control of the coal injection vessel could be significantly improved. With the new control, the coal mass flow can be used as a control parameter for the blast furnace. High injection rates can be used and more coke substituted.

8 Acknowledgment

The authors want to thank SSAB Tunnplåt in Luleå, for making the coal injection plant and maintainance personnel available to us. Particularly, a special thank to Ingemar Lundström and Robert Johansson who helped with the experiments. Information and assistance provided during the project by Andreas Weber, BMH are also gratefully acknowledged.

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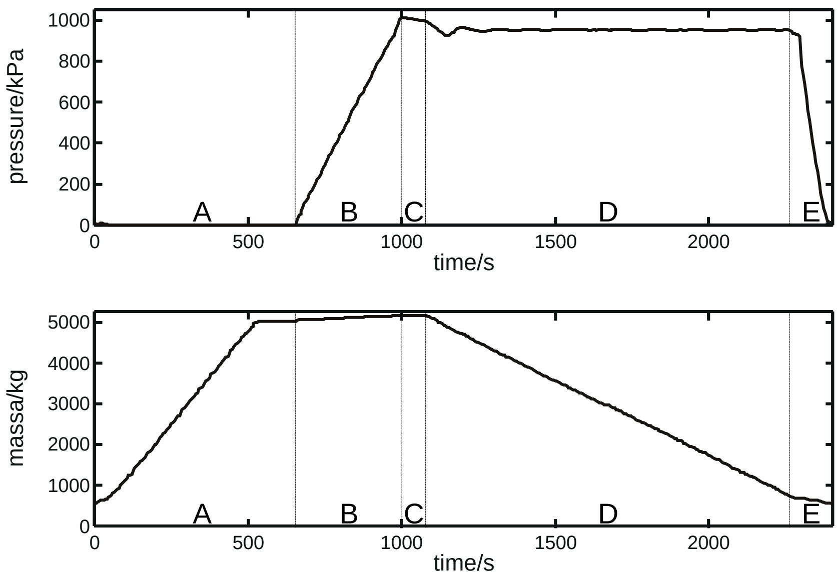

Fig. 2: Process working cycle of injection vessel S21

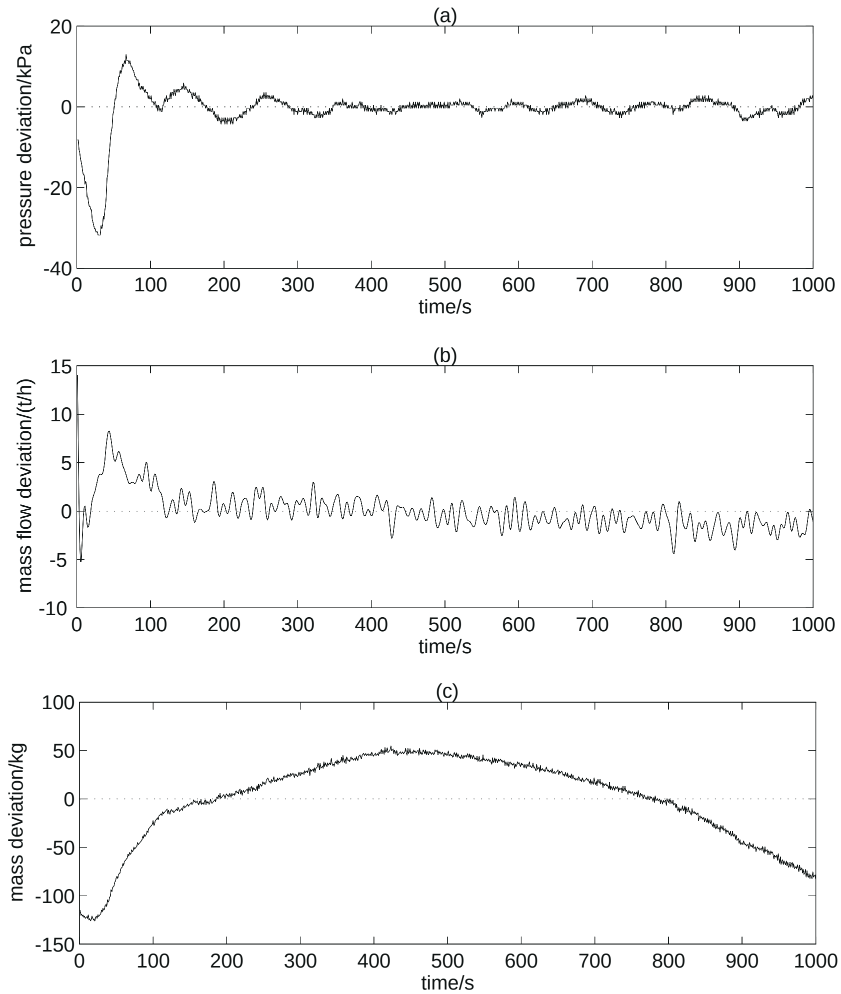

Fig. 3: Performance of the existing control unit

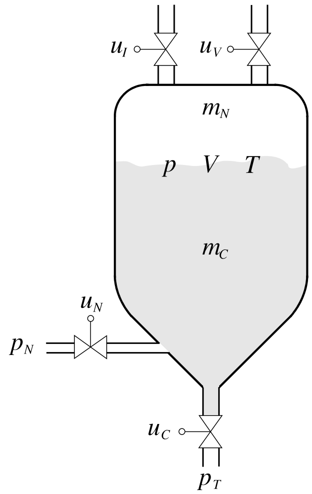

Fig. 4: Schematic drawing of an injection vessel

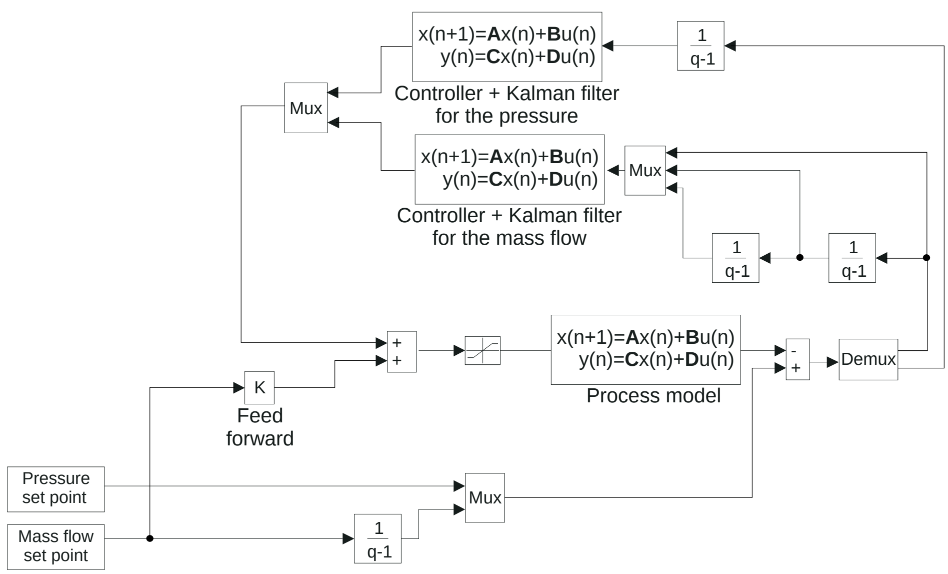

Fig. 5: Block diagram of the two separate control loops with implemented feed forward

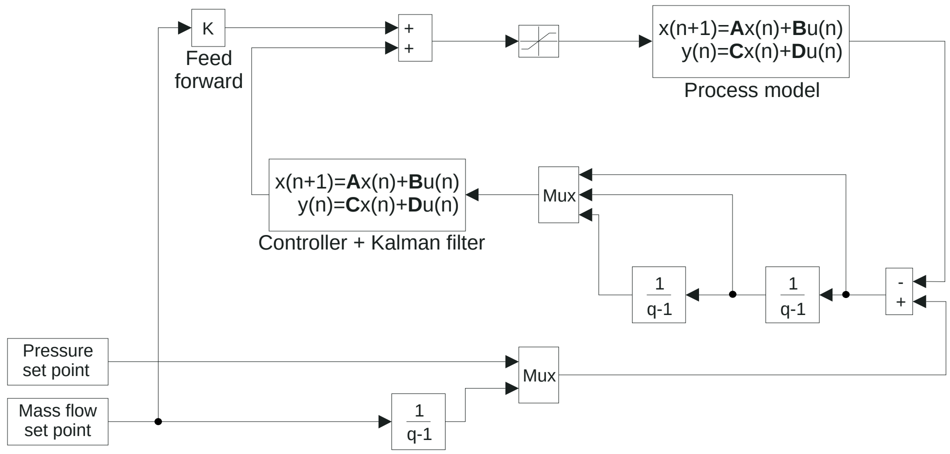

Fig. 6: Block diagram of the MIMO controller with implemented feed forward

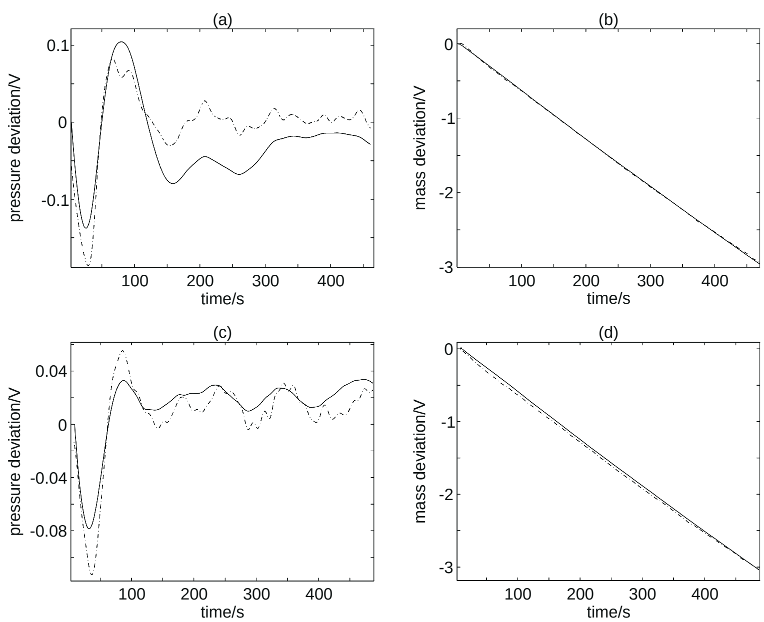

Fig. 7: Common sense tests. Measured output (dashed-dotted) and simulated output (solid).

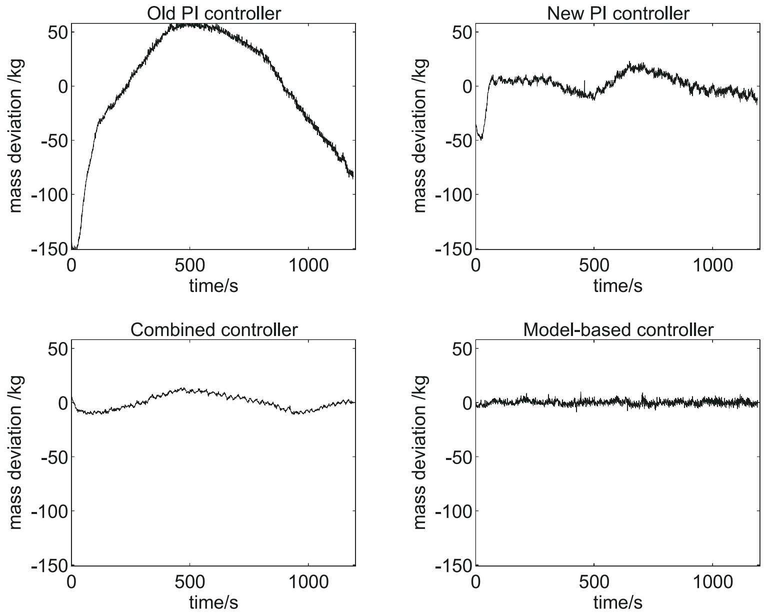

Fig. 8: Comparison of the mass deviation of the four controllers.

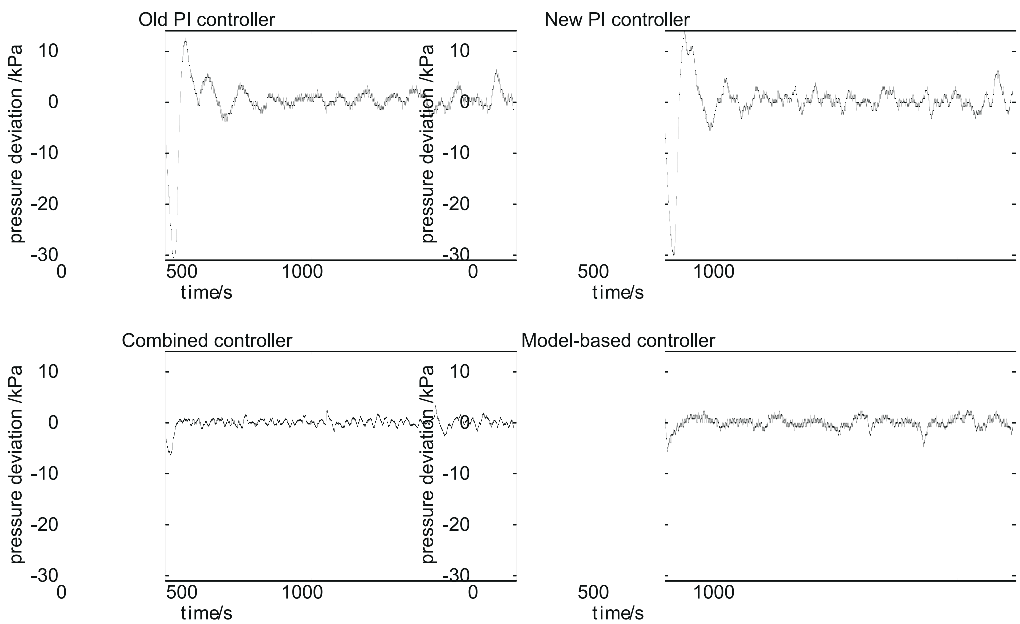

Fig. 9: Comparison of the pressure deviation of the four controllers.

Paper 2

Model-Based Control for a Fine Coal Injection Plant

Published as

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Model-Based Control for a Fine Coal Injection Plant

Wolfgang Birk, Andreas Johansson, and Alexander Medvedev [1]

Abstract

This paper deals with modeling, identification, model-based control, and fault detection of a fine coal injection vessel for the use of the blast furnace injection process. A nonlinear dynamic model of the plant is developed in order to describe the plant behavior under normal and abnormal conditions. An observer is employed to estimate the leakage flow. Then, by means of the Generalized Likelihood Ratio method, the most feasible cause of the leakage is singled out from a given set of hypotheses. A linear dynamic model is identified in the closed-loop mode, without disrupting normal process operation, and utilized for controller performance analysis and re-design. The controller and leakage detection algorithms are implemented in a PC-based control system called SafePCI. SafePCI has successfully been tested during a two-weeks period at No.2 Blast Furnace at SSAB Tunnplåt, Luleå.

1 Introduction

The high coal injection rate target (over 200 kg/tHM) generally set by steelmakers means larger grinding mills, injection vessels, control valves and, at the same time, demands higher system availability. Eichinger and Rafi describe in [1] how high coal injection rates influence blast furnace operation. Introducing closer process control and monitoring is a reasonable measure to achieve a higher equipment performance and reliability.

Fine coal injection vessels constitute a crucial part of a coal injection system for the blast furnace providing control of the fine coal flow being blown into the furnace. One way of achieving a steady and controllable flow is to stabilize the pressure in the sending vessel. Conventionally, a standard PI-controller

[1] Wolfgang Birk, Andreas Johansson, and Alexander Medvedev are with the Control Engineering Group at Luleå University of Technology, SE - 971 87 Luleå, Sweden. e-Mail: Wolfgang.Birk@sm.luth.se
The authors want to thank SSAB Tunnplåt in Luleå, for making the coal injection plant and maintenance personnel available to us. Particularly, special thanks to Ingemar Lundström and Robert Johansson who helped with the experiments. Financial support of the Center for Process and System Automation (*ProSA*) at Luleå University of Technology provided by *Norrbottens Forskningsråd* is also gratefully acknowledged.

is used to keep the vessel pressure constant in spite of the losses through the valve regulating coal powder discharge. If the discharge valve had a constant opening, the integral term in the PI-controller would probably accommodate for the corresponding outgoing flow. Since there is an independent flow control loop comprising the discharge valve, the varying valve opening acts as a fast varying disturbance preventing the pressure in the sending vessel from coming to a steady state value. Therefore, a more sophisticated control system should be in place to guarantee reliable pressure stabilization.

The prime drawback of coal powder is its flammability. The ability to self-ignite in contact with air makes it inconvenient to handle. It has to be stored and transported under inert conditions and injected under pressure directly into the reaction core. For these reasons it is of interest to have a reliable coal powder injection plant. A leakage in a valve could, for example, allow air to enter the injection vessel with possibly catastrophic consequences. In fact, one of the prime motivations for this work was an ember fire in one of the injection vessels at SSAB Tunnplåt in Luleå.

Model-based fault detection methods [2] appear to provide adequate tools for the solving the leakage detection and isolation problem. State estimation by observers is often used and a number of different linear techniques exist, for example Unknown Input Observers, Dedicated Observers, Parity Space and Kalman Filter-based Methods. Faults are usually divided in two groups: abrupt and incipient faults. A comprehensive study on the detection of abrupt changes in stochastic processes is the book by Basseville and Nikiforov [3]. Incipient faults are, in most cases, handled by the methods of System Identification.

In this paper, a project aimed at bringing the control system of fine coal injection vessel up to date with the increasing demands for quality process control is described. The paper's outline is as follows. First, a brief description of the coal injection plant is given. Then, a non-linear physical and a linear black-box model is considered for the plant description. Taking advantage of the identified dynamic model of the injection vessel, alternatives in the model-based controller design are analyzed. To achieve a tighter control, a multi-input multi-output controller is then suggested. Further, an observer-based gas leakage detection technique is developed to evaluate residuals sensitive to particular types of faults in worn-out valves. Implementation issues pertaining to the integrated fault-detection and control system are discussed. Finally, the results of experimental system exploitation at SSAB Tunnplåt, Luleå during a two-weeks period are reported.

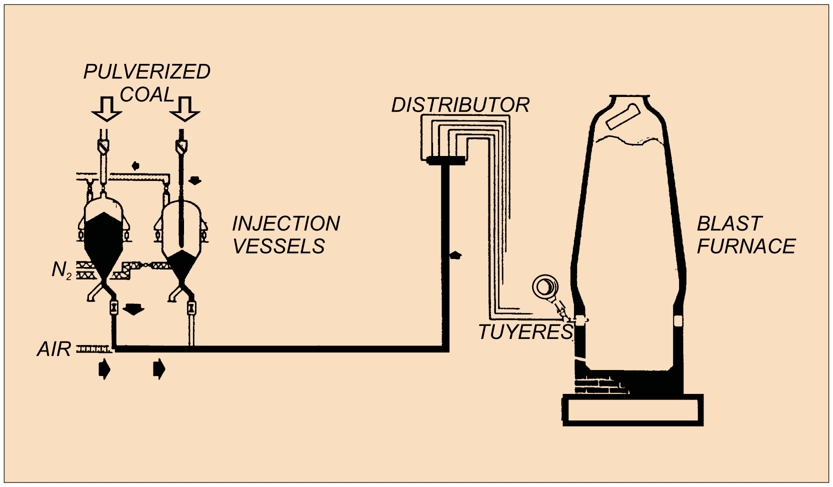

Fig. 1: Coal injection plant (injection vessels, distributor and blast furnace).

2 Process description

A coal injection plant is a highly automated plant, where incoming raw coal is stored, ground, dried and finally injected into the blast furnace. During operation, human interaction is only needed for set point adjustments.

Fig. 1 shows the structure of the plant, where only the injection vessels, distributor and the blast furnace are depicted.

2.1 Injection process

Principally, the injection process can be divided into two distinct phases: a high-pressure and a low-pressure phase, where, as the names already imply, the pressure in the injection vessel is high or low, respectively.

One vessel is depressurized, charged and pressurized while the other vessel is injecting coal powder, to provide continuous injection. To facilitate process modeling, identification and control, the high pressure and low pressure phases are sub-divided into more specific phases.

Fig. 2 shows one process cycle of an injection vessel. Part A represents the low pressure phase and the parts from B to E belong to the high pressure phase. In Table 1, the nomenclature used in the sequel to refer to different process phases is summarized.

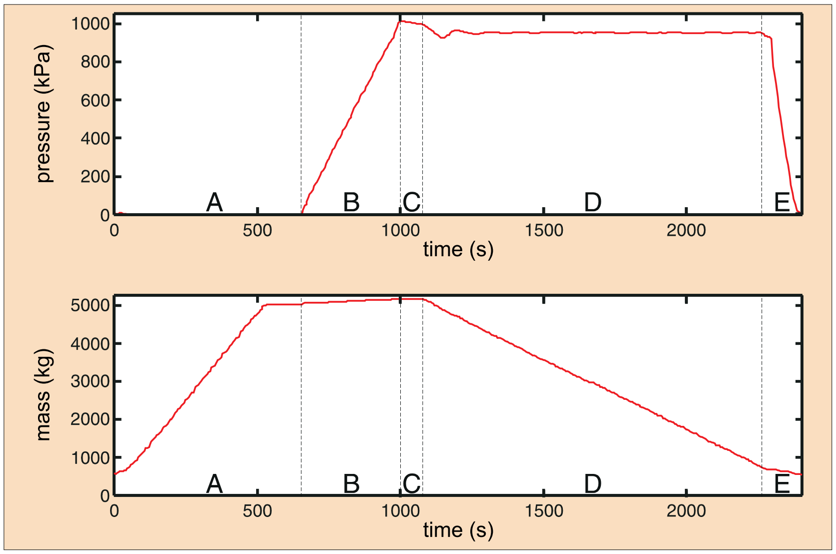

Fig. 2: Injection vessel process cycle.

Phase	Name	Description
A	Charging	The pressureless vessel is filled with coal powder
B	Pressurizing	The injection vessel is set under pressure
C	Pressure holding	Standby until the other vessel has finished injection
D	Injection	The coal powder is injected into the blast furnace
E	Ventilation	Depressurizing and ventilation of the vessel

Table 1: Process phases

Since only the pressure holding and injection phases require closed-loop control, the desired system performance at the individual phases can be summarized as follows.

1. Pressure holding phase:
   - The pressure in the injection vessel is held constant.
   - The mass of the injection vessel is constant (no discharge).
2. Injection phase:
   - The pressure in the injection vessel is held around a working point of 950 kPa.
   - The mass of the injection vessel follows a ramp, where the desired slope of the ramp is given by the coal mass flow set-point.
   - The injection phase ends when a minimum weight of fine coal in the vessel is reached.

The flammability of coal powder and relatively high price of nitrogen motivates the development of a gas leakage detection system for the injection vessels [4]. The specifications of the leakage detection system include:

- Leakage detection during normal plant operation.
- Isolation of the leakage to a certain valve or place.

3 Modeling

Basically, the injection vessels (see Fig. 3) are modeled as pressurized tanks. Both the pressurized tank model and the valves have non-linear behavior. Furthermore, additional non-linearities arise in this case due to the two-phase nature of the gas and coal powder mixture transported in the injection plant. In the next section, two dynamic process models, one a non-linear model based on physical conservation laws, and another, a linear model estimated from measurement data, are presented.

3.1 First principle model

The notation used in the following development is summarized in the Appendix.

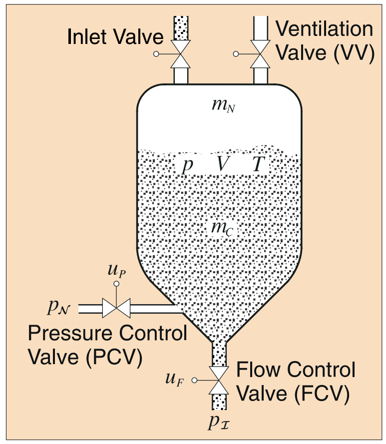

Fig. 3: Schematic drawing of an injection vessel.

3.1.1 Flows through Valves

Expressions for mass flow of fluids through restrictions can be derived from Bernoulli's law. For an incompressible liquid the flow is

$$q\left(p_1, p_2\right) = a f_{\text{liq}}\left(p_1, p_2\right) \triangleq a\sqrt{2\rho\left(p_1 - p_2\right)} \tag{1}$$

where p_1 is the pressure on the inlet side of the restriction and p_2 is the pressure on the opposite side. The constant ρ denotes the density of the liquid and a is the minimum cross-section area of the flow. In this paper, the equation above is used to describe the flow of coal powder.

When modeling a gas flow, the compressibility has to be taken into consideration. If the dynamic effects due to this property are neglected and the expansion process in the restriction is assumed to be adiabatic, then the following expression holds for the mass flow of an ideal gas.

$$\begin{aligned} q(p_1, p_2) &= a f_{\text{gas}}(p_1, p_2) \\ &\triangleq a p_1 \begin{cases} \left(\frac{p_2}{p_1}\right)^{\frac{1}{\gamma}} \sqrt{\frac{2\gamma}{RT_1(\gamma-1)}\left[1 - \left(\frac{p_2}{p_1}\right)^{\frac{\gamma-1}{\gamma}}\right]} & \frac{p_2}{p_1} < \left(\frac{2}{\gamma+1}\right)^{\frac{\gamma}{\gamma-1}} \\ \left(\frac{2}{\gamma+1}\right)^{\frac{1}{\gamma-1}} \sqrt{\frac{2\gamma}{RT_1(\gamma+1)}} & \text{Otherwise} \end{cases} \end{aligned} \tag{2}$$

The constant R and γ are the molecular mass and the compressibility factor of the gas, respectively, while T_1 represents the temperature on the inlet side.

For a control valve, the area a in (1) and (2) is a function of the input signal u, *i.e.* $a = kg(u(t))$ where k is a scaling factor and $g(u)$ is called the characteristic function of the valve. In order to make the definition above unambiguous it is also stated that $g(1) = 1$.

3.1.2 Output transformation

The pressure and total mass (p and m) of the vessel can be calculated from the masses of coal and nitrogen (m_C and m_N) using basic physical principles including the ideal gas law. With the definitions $y \triangleq \begin{bmatrix} m & p \end{bmatrix}^T$ and $x \triangleq \begin{bmatrix} m_C & m_N \end{bmatrix}^T$ the transformation can be expressed as

$$y = h(x) \triangleq \begin{bmatrix} m_C + m_N \\ m_N \frac{R_N T \rho_C}{V\rho_C - m_C} \end{bmatrix}$$

Index C in the equation above refers to pure coal and not coal powder. The reason for this is that the nitrogen is assumed to fill out the space between the coal particles. Since $h(x)$ is invertible, the coal and nitrogen masses can be considered measurable. The inverse transformation is given below.

$$x = h^{-1}(y) = \begin{bmatrix} \frac{mR_N T\rho_C - pV\rho_C}{R_N T\rho_C - p} \\ p\frac{V\rho_C - m}{R_N T\rho_C - p} \end{bmatrix}$$

Upon entering the vessel, the nitrogen passes through the coal powder. Since nitrogen has much lower heat capacity than coal powder, it is assumed to be instantaneously heated to the temperature of the coal powder ($60 - 70°C$). Therefore T is given the constant value of 338 K.

3.1.3 Entire system

The equations in this subsection are based on the mass conservation of nitrogen and coal. The change in coal mass and nitrogen mass in the vessel is the sum of the material flows into the vessel.

During the pressurization and injection phases, the material transport takes place through the PCV and the FCV. Therefore, the change in the coal and nitrogen masses of the vessel can be expressed as

$$\dot{m}_C = -q_{C,F} \tag{3}$$

$$\dot{m}_N = -q_{N,F} + q_{N,P} \tag{4}$$

where the q's denote the mass flows of coal (index C) and nitrogen (index N) through the FCV (index F) and the PCV (index P).

With the following definition

$$u \triangleq \begin{bmatrix} u_{C,F} \\ u_{N,F} \\ u_{N,P} \end{bmatrix} \triangleq \begin{bmatrix} f_{\text{liq}}(p, p_{\mathcal{I}})\, g_F(u_F) \\ f_{\text{gas}}(p, p_{\mathcal{I}})\, g_F(u_F) \\ f_{\text{gas}}(p_{\mathcal{N}}, p)\, g_P(u_P) \end{bmatrix}$$

where g_F and g_P are the characteristic functions of the FCV and PCV, respectively, the system (3) and (4) can be written as

$$\dot{x} = Ax + Bu \tag{5}$$

where

$$A \triangleq \begin{bmatrix} a_C & 0 \\ 0 & a_N \end{bmatrix}$$

$$B \triangleq \begin{bmatrix} k_{C,F} & 0 & 0 \\ 0 & -k_{N,F} & k_{N,P} \end{bmatrix}$$

$$u \triangleq \begin{bmatrix} u_{C,F} & u_{N,F} & u_{N,P} \end{bmatrix}^T$$

For an ideal plant, the constants a_C and a_N should be equal to zero, but to obtain extra degrees of freedom, they are considered unknown. When identifying, the parameter a_N takes a small negative value, which suggests inherent leakage in the vessel.

Fig. 4 shows a simulation of the pressure output of the non-linear system with input signals u_F and u_P.

During the ventilation phase, the only material flow is the nitrogen flow through the ventilation valve (VV). Thus

$$\begin{aligned} \dot{m}_C &= 0 \\ \dot{m}_N &= -q_{N,V} \end{aligned} \tag{6}$$

The nitrogen flow can be accurately described by a first order polynomial in p. Since $p = c(m_C)m_N$ where $c(m_C) \triangleq R_N T \rho_C / (V\rho_C - m_C)$, (6) can be written as

$$\dot{m}_N = -k_0 - k_1 c(m_C)\, m_N$$

The coefficients k_0 and k_1 have been estimated by least squares identification. Fig. 5 shows a simulation of the pressure during ventilation in comparison with logged data.

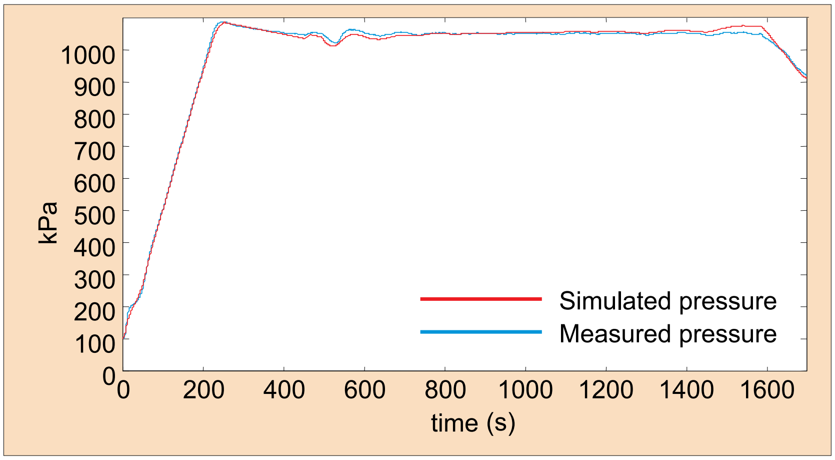

Fig. 4: Simulation of the pressure in the vessel during pressurization and injection.

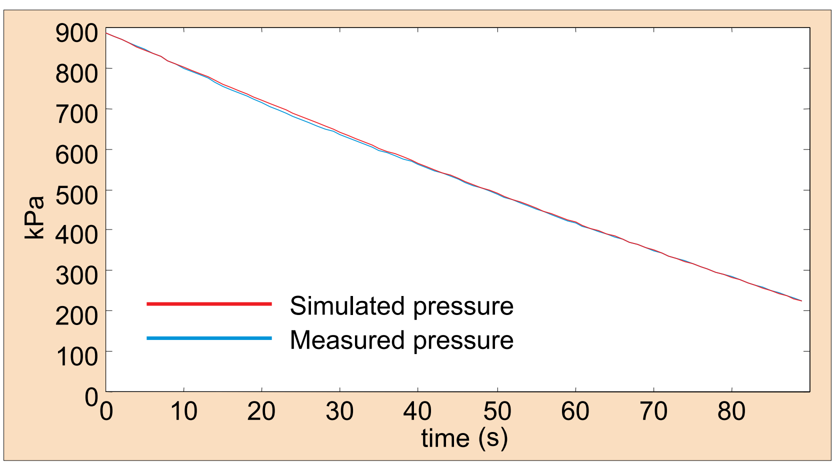

Fig. 5: Simulation of the pressure in the vessel during ventilation.

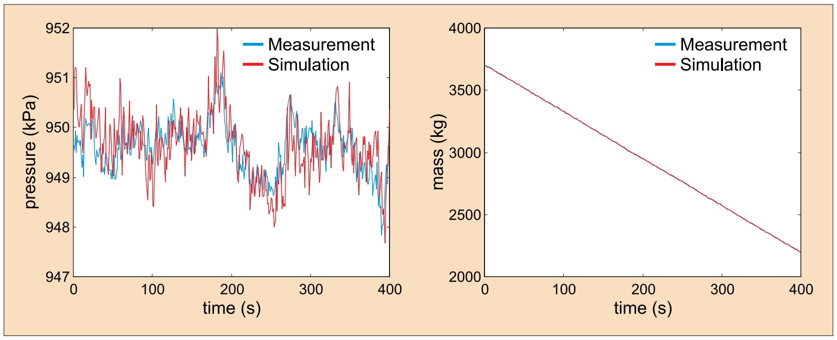

Fig. 6: Common sense tests, measured output (dashed-dotted) and simulated output (solid).

3.2 Identification

To obtain a more accurate model of the injection phase, process identification from measured plant data from is applied, assuming that the non-linear plant dynamics can be linearized around a working point. There are two candidate model structures to start with: two single input, single output models (SISO) or one multiple-input, multiple-output (MIMO) model. However, SISO models would not take into account the couplings between pressure and mass which are obvious from the nonlinear model 5

The MIMO model describes the process dynamics for p, m as outputs and u_P, u_F as inputs. The latter can be used to design a MIMO controller for the process.

Identification and validation data sets are logged on the process with a sampling time of 1 s. The identification method is discussed in [5], where the subspace identification method *n4sid* is applied to the Laguerre spectra of the input/output data. As shown in [5], *n4sid* appears to perform better in the Laguerre domain compared to the time domain, when disturbances are mostly caused by unmodelled dynamics and nonlinearities. The subspace identification method is chosen due to the MIMO nature of the process.

Fig. 6 shows the simulated output of the MIMO model and the corresponding measured output. The mismatch of simulated and measured output results from measurement noise and process uncertainty. It has to be remarked that the logged data sets are acquired in closed loop, since the plant is continuously in operation.

4 Controller design

As mentioned before, the control objective is to guarantee a constant coal mass flow from the injection vessel to the blast furnace. This can be achieved by pursuing different strategies. Here are two of them:

1. Two separate control loops for the pressure and the coal powder discharge.
   One unit controls the vessel pressure and the other controls the coal powder discharge, where the vessel mass has to follow a pre-defined ramp. The slope of the ramp is the coal mass flow set point. For the controllers design, two SISO models for the plant are used.

2. A MIMO design controlling the pressure and the coal powder discharge.

From the physical MIMO nature of the process follows that the pressure and mass variables are coupled. That means changes in the pressure influence coal mass discharge. By designing two individual SISO control loops based on SISO models for the pressure and the mass, these couplings cannot be used advantageously. In fact, neglecting the couplings can have negative effects on the control performance. Naturally, a MIMO controller based on a MIMO model of the process can be designed to take advantage of the couplings in the process. Then, the pressure can be varied to counter-act the flow disturbances in coal powder discharge, providing an additional control action in comparison to the case of two SISO systems. Notice that a set point for the pressure is needed for both SISO and MIMO design, since the pressure has to be hold in a certain working range. However, in the MIMO design, the variations in the sending vessel pressure have a stabilizing effect on the flow, whilst they are highly undesirable in the SISO control strategy. Basing on these considerations, the MIMO controller is chosen for implementation.

An optimal design method is used to design the controller. The MIMO design is based on the so-called linear quadratic gaussian (LQG) theory, as described in [6]. It is also a major design tool for multi-variable linear systems. To obtain the above described behavior of the controller, the weighting matrices in the design procedure are chosen so that the pressure variations contribution to the quadratic loss function is small compared to that of the flow variations. Therefore, the controller keeps the coal powder flow constant by manipulating the flow control valve as well as by properly varying the pressure in the vessel.

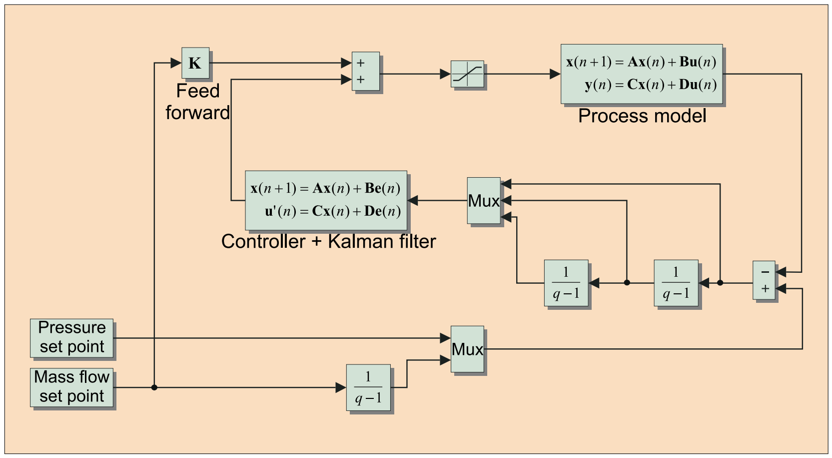

Fig. 7: MIMO controller with feedforward.

Normally, assuming state feedback, a proportional-integral control law is chosen to eliminate the steady state error. However, controlling the fine coal mass, one of the closed-loop system outputs has to follow a ramp. Then, the use of a single integrator leads to a constant steady state response error. Thus, a double integration is used to drive the error in reference signal following to zero. To recast this control law in the framework of LQG design, the process models have to be augmented with a double integrator.

For the resulting system, a steady state Kalman filter is designed, to obtain filtered versions of the measured signals. The steady-state Kalman filter seems to be sufficiently fast, as the process itself is very slow.

Furthermore, to reduce the controller settling time and retain a smooth set point change response, a feedforward signal from the desired coal mass flow to the valves is introduced. The design method presented in [7, chapter 6] is adopted.

Practical tests with both controllers revealed that the MIMO controller yields the better results. In [8] it has been shown that performance improvements up to 86% compared to the conventional controller can be reached, where the standard deviations from the set-points are used as a measure. Fig. 7 shows a block diagram for the MIMO controller design, including the process model. Since the MIMO controller is superior to the two SISO controllers, only the MIMO controller is used during the full-scale experiments.

Leakage	Possible consequence	Notation
To the atmosphere	Loss of nitrogen	$\mathcal{A}$
From the nitrogen network	Overpressurized vessel	$\mathcal{N}$
To/from the injection pipe	Fire	$\mathcal{I}$
No Leakage	-	$\emptyset$

Table 2: Leakages

5 Fault detection and isolation

5.1 Leakage model

Three different leakages are considered (Table 2). The set of leakages is denoted $\mathcal{L} \triangleq \{\mathcal{A}, \mathcal{N}, \mathcal{I}, \emptyset\}$. A leakage can be interpreted as the flow through a valve with an unknown control signal. The nitrogen leakage flow can thus be represented by

$$q_\ell = k_\ell f_\ell(\cdot) \quad \ell \in \mathcal{L} \tag{7}$$

where k_ℓ is an unknown time-varying factor and $f_\ell(\cdot)$ is a function of the pressures on each side of the leakage. The trivial leakage function for the event of 'No Leakage' is $f_\emptyset = 0$. The other leakage functions ($f_\mathcal{A}$, $f_\mathcal{N}$ and $f_\mathcal{I}$) are developed from (2).

5.2 Fault detection

A leakage flow is modeled as an extra term added to the right hand side of (4) and (6). The purpose of the fault detection algorithm is thus to calculate these terms. However, the presence of noise makes it necessary to utilize filtering. The nonlinear models developed in the modeling section are of the Hammerstein type (linear dynamic system with static input or output nonlinearities) and linear observers can be used.

5.2.1 Pressurization and injection

When leakage is taken into account, equation (4) has to be extended with the term q_L that represents the net leakage of nitrogen into the vessel, *i.e.* $\dot{m}_N = -q_{N,F} + q_{N,P} + q_L$. A linear observer for the system (5) is given by $\dot{\hat{x}} = A\hat{x} + Bu + K\epsilon$ where

$$\epsilon \triangleq \begin{bmatrix} \epsilon_C \\ \epsilon_N \end{bmatrix} \triangleq \begin{bmatrix} m_C - \hat{m}_C \\ m_N - \hat{m}_N \end{bmatrix}$$

Since in this system the state variables are decoupled (*i.e.* A is diagonal), it is assumed that a diagonal observer gain matrix is sufficient.

$$K \triangleq \begin{bmatrix} K_C & 0 \\ 0 & K_N \end{bmatrix}$$

With the definitions above, it can be shown that the residual ϵ_N is the net leakage q_L, filtered through a first order filter. This filter is given by

$$\epsilon_N(t) = \frac{1}{\mathbf{p} - a_N + K_N} q_L(t) = H_{\mathfrak{P}}(\mathbf{p})\, q_L(t)$$

where $\mathbf{p}$ is the differentiation operator. Because the net leakage is assumed to be slowly varying in time, the residual divided by the static gain of the filter above is a good approximation of the net leakage whose measurement is corrupted by white noise, thus $\hat{q}_L(t) = \epsilon_N(t)/H_{\mathfrak{P}}(0)$.

5.2.2 Ventilation

The net leakage q_L is introduced in the ventilation phase model in a manner similar to the leakage in the model of pressurization and injection phase, thus $\dot{m}_N = -q_{N,V} + q_L$. An observer for the ventilation phase is given by $\dot{\hat{m}}_N = -k_0 - k_1 c\hat{m}_N + K_N \epsilon_N$. The residual ϵ_N is then the net leakage filtered through a first order filter.

$$\epsilon_N(t) = \frac{1}{\mathbf{p} + k_1 c + K_N} q_L(t) = H_{\mathfrak{V}}(\mathbf{p})\, q_L(t)$$

As before, the residual is divided by the static gain of the filter, *i.e.* $\hat{q}_L(t) = \epsilon_N(t)/H_{\mathfrak{V}}(0)$.

5.3 Isolation of leakages

Since the nonlinear models are of the Hammerstein type, it is a straightforward task to obtain an approximate discretization of the observers. In this section, all signals are assumed to be discrete, which is indicated by the new time variable n.

The calculated leakage is assumed to be the sum of a scaled leakage function and a disturbance, i.e. $\hat{q}_L(n) = k_\ell f_\ell(n) + e(n)$ where $\ell \in \mathcal{L}$ and the term $e(n)$ is stationary zero-mean white gaussian noise with variance σ^2, i.e. $e(n) \in \mathcal{N}(0, \sigma)$

Actually, $e(n)$ is not gaussian (probably due to unmodeled non-linearities), but this fact, as appears, does not have any major influence. When the transients are excluded from the residual, it is fairly near normal distribution, but the results are virtually the same.

The factor k_ℓ in (7) is a measure of the size of the hole through which the leakage takes place. This means that k_ℓ varies slowly in time when describing incipient leakages. If it is assumed to be constant during a reasonably long period of time (for example a process cycle), it can be estimated using the Generalized Likelihood Ratio.

5.3.1 Leakage hypothesis testing

Isolate the leakage flow $\hat{q}_L(t)$ to a certain valve or place can be done automatically. It is also desirable to provide the user with an estimate of the physical size of the leakage and the confidence level of its detection and isolation. This motivates the use of the Generalized Likelihood Ratio (GLR) for fault isolation.

Four hypotheses ($\mathcal{H}_\emptyset$, $\mathcal{H}_\mathcal{A}$, $\mathcal{H}_\mathcal{N}$ and $\mathcal{H}_\mathcal{I}$) are formed in agreement with the leakage events. The three leakage hypotheses are tested one by one against $\mathcal{H}_\emptyset$ using the Generalized Likelihood Ratio (GLR). If $\mathcal{H}_\emptyset$ is rejected in more than one of these tests, the hypothesis with the highest GLR is accepted. The likelihood functions for the hypotheses can be expressed as

$$P_\ell(\hat{q}_L) = \prod_{n=1}^{N} \frac{1}{\sigma}\varphi\left(\frac{\hat{q}_L(n) - k_\ell f_\ell(n)}{\sigma}\right)$$

where φ is the density function of the normal distribution. The GLR for each leakage hypothesis is

$$\Lambda_\ell(\hat{q}_L) = \frac{\sup\limits_{k_\ell>0} P_\ell(\hat{q}_L)}{P_\emptyset(\hat{q}_L)}$$

The restriction on k_ℓ comes from the fact that a negative k_ℓ would imply a leakage flow from a lower pressure to a higher. It can be shown that under the conditions above, the leakage likelihood function, $P_\ell(\hat{q}_L)$, is maximized by

$$\hat{k}_\ell = \arg\sup_{k_\ell>0} P_\ell(\hat{q}_L) = \begin{cases} \frac{C_\ell}{\sum_{n=1}^{N} f_\ell^2(n)} & C_\ell > 0 \\ 0 & \text{Otherwise} \end{cases}$$

where $C_\ell \triangleq \sum_{n=1}^{N} \hat{q}_L(n) f_\ell(n)$.

The logarithmic GLR can then be expressed as

$$\ln\left(\Lambda_\ell\left(\hat{q}_L\right)\right) = \begin{cases} \frac{C_\ell^2}{2\sigma^2 \sum_{n=1}^{N} f_\ell^2(n)} & C_\ell > 0 \\ 0 & \text{Otherwise} \end{cases}$$

A threshold value, λ, for Λ_ℓ must be set. If Λ_ℓ exceeds this value then the null hypothesis, $\mathcal{H}_\emptyset$, is rejected. The threshold is generally calculated using the probability of rejecting the null hypothesis when it is true, i.e. $\mathbb{P}[\Lambda_\ell(\hat{q}_L) > \lambda] \triangleq \alpha$, where $\mathbb{P}$ is the probability operator. The probability α is called the level of the test and is usually set to a value in the range of $0.1\% \leqslant \alpha \leqslant 5\%$. In the case of the pressurization and injection system, however, the severity of the modeling errors causes the stationarity and zero-mean conditions on the disturbance $e(n)$ to fall. This makes it necessary to use very low test levels to prevent a high rate of false alarms. For this reason, no fault probability is chosen but instead a suitable value of the threshold λ is evaluated on the basis of experimental data (see the Experiments section).

6 Implementation

Since the currently used control systems at SSAB are not suitable for model-based control and on-line gas leakage detection, additional hardware becomes necessary. To reduce the cost factors for the experiments and achieve a high degree of flexibility, a PC-based implementation is chosen.

The needed software, called *SafePCI*, has been specially designed to maximize the performance. In the first version of *SafePCI*, a DOS-based control system is developed, because of stability reasons and the short design times for this operating system. Later versions are planned to take advantage of real-time operating systems like *real-time Linux* or *QNX*. All components are standard and commercially available at relatively low cost. Fig. 8 shows the complete system hardware and its general arrangement.

Since model-based control and leakage detection result in rather complex mathematical computations, the control and detection algorithms should be implemented in separate computers. An obvious advantage of this solution is an increased reliability. Furthermore, all data can be backed up on two independent computers. A disadvantage is that the sampled signals have to be transferred between the computers.

For the control part, the program RegSim [9] is used, which has been modified so that communication between the two computers is possible. The

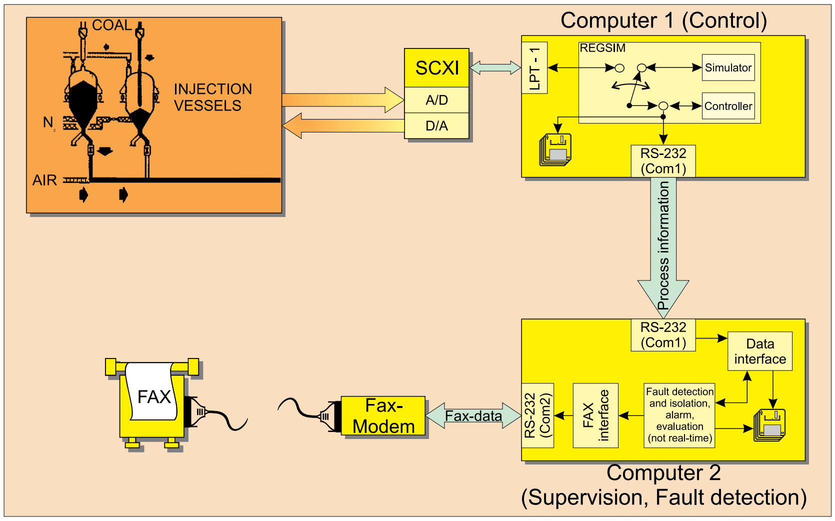

Fig. 8: Hardware structure, data flow and used components.

leakage detection and messaging are implemented in the second computer using an in-house software.

6.1 User interface of *SafePCI*

Both computers are equipped with a graphical user interface (GUI) enabling plotting curves and directly accessing information on the plant's state. All logged information is stored, what makes it possible to review process history, as for example the detection of a fault or the standard deviations of the controlled variables at the injection phase. The user has the possibility to change the appearance of information on-line. The GUI can be accessed in a MATLAB-like way, with specific commands that are issued to the system via the command line.

Control and detection parameters can also be adjusted on-line via the command line, which feature necessitates a multi-user structure, where each user has his/her own account to access system commands or parameters. A set of macro-commands is pre-defined to facilitate screen adjustments, and to enable fast switching between different views.

Fig. 9 shows the GUI screen run on Computer 2. The encircled letters denote different significant areas on the screen. Area A is the command line,

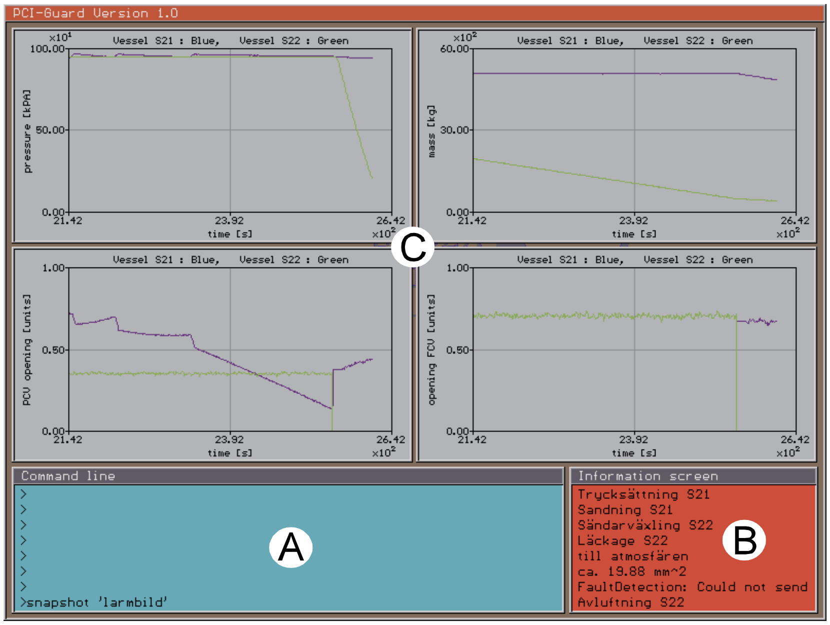

Fig. 9: Graphical user interface on Computer 2, a leakage is detected.

from which the user can interact with the program. Area B is the information window, where the program displays messages from the tasks being executed. The leakage detection task displays all necessary information and toggles the background color from green to red in case a leakage is detected.

The GUI toggles the background color back to green, only after all leakage detection reports are read by the user. Area C is the plot area and can be freely defined by any registered user. The user can define plot windows and assign measurement and evaluation data to these windows. Once the assignments are made, the plot windows are updated automatically. Finally, an on-line help function for all implemented system commands is provided.

The GUI implemented on Computer 1 is simpler. On-line user interaction is only supported in a very restricted manner. The information display is fixed and cannot be adjusted as on Computer 2.

6.2 Simulation and Real-time Execution

As indicated by the switch in Fig. 8, the model-based control and gas leakage detection system can be run in either simulation mode or real-time execution. Since the simulation model supports leakage, the controllers can be tested in advance regarding their reliability and performance under different types of leakages. The leakage detection algorithm implemented on the second computer can therefore be tested in the closed-loop mode, too.

6.3 Safety aspects

The add-on character of the implementation has however a drawback. Communication between the rest of the control unit hardware and the PCs becomes the crucial part of the implementation. In case of a malfunction in either Computer 1 or the communication link, the control unit has to respond to it and switch back to the simple PI-controllers implemented in the conventional system. For safety reasons, these malfunctions have to be detected and an appropriate reaction has to be initiated.

The following scenarios are considered and appropriate safety measures are suggested.

- Computer 1 (Control) crashes.
  In this case the output signals of the D/A converter 'freeze' at the last received values, and the controller has no access to the actuators. Furthermore, no data are transferred to Computer 2.
  Two safety measures are suggested here to handle this kind of problem. On the one hand, Computer 1 sends an "alive-signal" to the conventional control unit in the form of a square wave. In case of a malfunction, the square wave disappears and the control unit will switch to its back-up controllers.

  On the other hand, a time-out function is implemented on Computer 2, monitoring the time slot between two consequent data transfers between the computers. In case the time-out value is exceeded, an emergency flag is set in the control unit which causes the control unit to switch to the back-up controllers. The computer has to be restarted.

- Computer 2 (Supervision, fault detection) crashes.
  Computer 2 ceases to send regular evaluation facsimile transmissions.
  Since the reaction time to a fault message is not crucial and normally measured in hours, Computer 2 has simply to be restarted.

- Malfunction of the controller.
  Computer 2 is provided with supervision and evaluation algorithms to detect this malfunction.
  Computer 2 sets an emergency flag in the control unit and causes the control unit to switch back to the back-up controllers. A malfunction message is generated and sent to the staff.

- Cable fault.
  The same effects as a malfunction of the controller.

7 Experiments

The experiments have been performed at SSAB Tunnplåt, Luleå, during a two week period. This enables studying the system performance under set-point changes and in abnormal situations.

The system hardware (Fig. 8) is connected to the coal injection plant via buffer amplifiers. Altogether, 40 signals are sampled and four actuators are manipulated. Eight of the sampled signals are needed for control and the additional signals are for supervision and fault detection purposes.

7.1 Control

During the experiment, the control performance is analyzed by comparing standard and maximum deviations from set-points when running the conventional control system. To prevent integrator wind-up, it is checked whether the control signals to the actuators are within working range.

Fig. 10 shows the deviations from the set-point for the pressure and mass during the injection phase of a process cycle. The deviations are very low compared to the deviations produced by the conventional control unit. A maximum deviation of less than 4 kPa for the pressure and 10 kg for the mass has been reached, that is less than 1% of the corresponding reference values.

Since the controller has kept this control performance during a significant number of working cycles for different set-points, it can be considered for permanent installation.

7.2 Gas leakage detection

Different leakages were artificially created during a period of six process cycles.

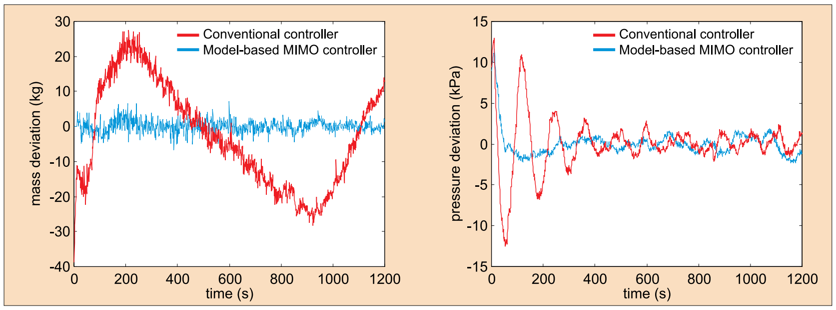

Fig. 10: Control performance (example).

The logarithmic GLR, $\ln(\Lambda_\ell(\hat{q}_L))$, for each leakage type and each cycle is shown in Fig. 11 and Fig. 12. The dashed line in Fig. 12 shows the threshold for GLR when $\alpha = 1\%$ while the two dashed lines in Fig. 11 mark an interval for the threshold. Note that the hypothesis of leakage from the nitrogen supply system is absent in Fig. 11. The reason for this is that the pressure control valve, which connects the nitrogen network with the vessel, is open during pressurization and injection. There can, for obvious reasons, not be a leakage through an open valve.

Cycle	Pressurization and injection: Conclusion	Pressurization and injection: Real Leakage	Ventilation: Conclusion	Ventilation: Real Leakage
1	Atmosphere	Atmosphere	Atmosphere	Atmosphere
2	Atmosphere	Atmosphere	No leakage	Atmosphere and nitrogen network
3	No leakage	No leakage	Nitrogen network	Nitrogen network
4	No leakage	No leakage	Nitrogen network	Nitrogen network
5	No leakage	No leakage	No leakage	No leakage
6	No leakage	No leakage	No leakage	No leakage

Table 3: The detected leakages

Table 3 shows the conclusions drawn for the pressurization and injection phases if the threshold for the GLR is placed anywhere between the dashed lines of Fig. 11 and the conclusions for the ventilation phase with level 1%. Included in the table are also the real leakages. It can be seen that the leakages are correctly isolated in all cases but one. In this case there were two simultaneous leakages (Atmosphere and nitrogen network) that nearly canceled each other.

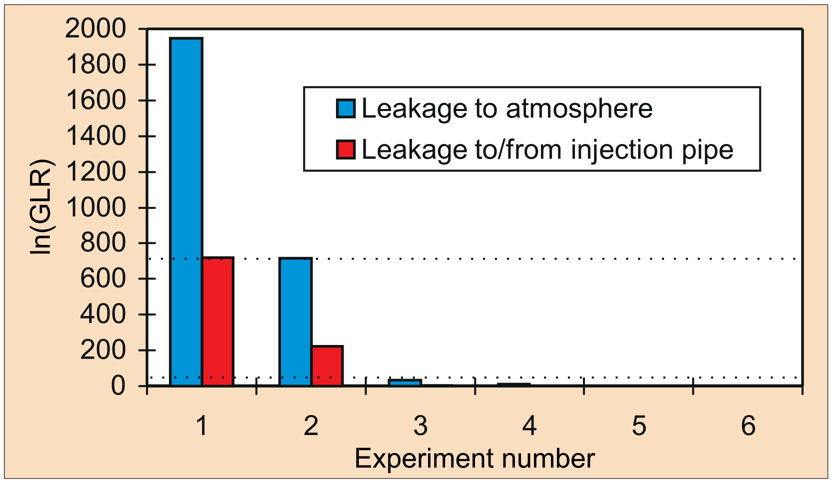

Fig. 11: The logarithm of the GLR for each leakage type and each experiment during pressurization.

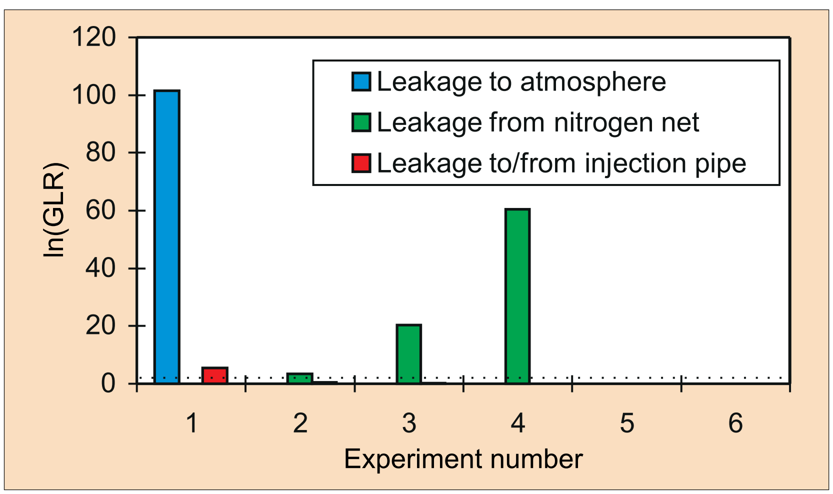

Fig. 12: The logarithm of the GLR for each leakage type and each experiment during ventilation.

8 Conclusions

Modeling, control of and gas leakage detection in the coal injection process are discussed. It is shown that by use of model-based methods, the flow and pressure of the coal injection vessel are reliably controlled. With the new control law, the coal mass flow can be used as a control parameter for the blast furnace. High injection rates can be used and more coke substituted. This is expected to yield a cost reduction in the iron production.

An experimental comparison of the conventional control unit with the one suggested in this paper shows that an improvement of the process efficiency can be reached by other means than increasing the capacity of the plant.

The experiments have also proved the model-based on-line gas leakage detection to be a promising way of maintaining high availability and safety in the plant.

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APPENDIX

A Nomenclature

Variable	Meaning
a_C, a_N	Parameters of the pressurization and injection process
k_0, k_1	Parameters of the ventilation process
$k_{C,F}$, $k_{N,F}$, $k_{N,P}$	Parameters of the pressurization and injection process
$k_{\mathcal{A}}$, $k_{\mathcal{I}}$, $k_{\mathcal{N}}$	Leakage parameters
m	Total mass in the vessel
m_C, m_N	Masses of coal and nitrogen in the vessel
p	Pressure in the vessel
$p_{\mathcal{A}}$, $p_{\mathcal{I}}$, $p_{\mathcal{N}}$	Pressure of the atmosphere, injection pipe and nitrogen network
$q_{C,F}$	Mass flow of coal through the FCV
q_L, $\hat{q}_L$	Real and approximated leakage flow
$q_{N,F}$, $q_{N,P}$, $q_{N,V}$	Mass flow of nitrogen through the FCV, PCV and VV
R_N	Gas constant for nitrogen
ρ_C	Density of coal
T	Temperature in the vessel
u_F, u_P	Control signal for the FCV and the PCV
V	Volume of the vessel

B Leakage Detection Algorithm

1. Leakage estimation

Pressurization and injection: Calculate the residual of the discrete time observer

$$\hat{x}(n+1) = \Phi\hat{x}(n) + \Gamma u(n) + K\epsilon(n)$$

where $\Phi = e^{A\Delta}$ and $\Gamma = \int_0^\Delta e^{A\tau} d\tau B$. The sampling interval is represented by Δ and the residual is defined by

$$\epsilon \triangleq \begin{bmatrix} \epsilon_C \\ \epsilon_N \end{bmatrix} \triangleq \begin{bmatrix} m_C - \hat{m}_C \\ m_N - \hat{m}_N \end{bmatrix}$$

The observer feedback gain is

$$K \triangleq \begin{bmatrix} K_C & 0 \\ 0 & K_N \end{bmatrix}$$

Depressurization: Calculate the residual of the discrete time observer

$$\hat{m}_N(n+1) = \phi \hat{m}_N(n) - k + K_N \epsilon_N(n)$$

where $\phi = e^{-k_1 c(m_C)\Delta}$ and $k = \frac{k_0}{k_1 c(m_C)}(1-\phi)$.

The estimated leakage flow is then obtained from

$$\epsilon_N(n) \approx \frac{1}{K_N} \hat{q}_L(n)$$

2. The logarithmic GLR for each leakage hypothesis is calculated using

$$\ln\left(\Lambda_\ell\left(\hat{q}_L\right)\right) = \begin{cases} \frac{C_\ell^2}{2\sigma^2 \sum_{n=1}^{N} f_\ell^2(n)} & C_\ell > 0 \\ 0 & \text{Otherwise} \end{cases}$$

where $C_\ell \triangleq \sum_{n=1}^{N} \hat{q}_L(n) f_\ell(n)$. An estimate of the size of the leakage hole for each hypothesis is determined from

$$\hat{k}_\ell = \arg \sup_{k_\ell > 0} P_\ell\left(\hat{q}_L\right) = \begin{cases} \frac{C_\ell}{\sum_{n=1}^{N} f_\ell^2(n)} & C_\ell > 0 \\ 0 & \text{Otherwise} \end{cases}$$

3. The logarithmic GLR for the leakage hypotheses are compared with each other. The hypothesis with the highest value is chosen if it exceeds a predetermined threshold, otherwise the hypothesis 'No Leakage' is assumed true.

Paper 3

Sensitivity Analysis of an LQ Optimal Multivariable Controller for a Fine Coal Injection Vessel

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Sensitivity Analysis of an LQ Optimal Multivariable Controller for a Fine Coal Injection Vessel

Wolfgang Birk and Alexander Medvedev
Control Engineering Group
Luleå University of Technology
SE-971 87 Luleå, Sweden
Wolfgang.Birk@sm.luth.se

Abstract

This paper deals with a sensitivity analysis of an LQ optimal multivariable controller for a fine coal injection vessel used in the blast furnace process. The multivariable controller from a previous work is briefly presented and the closed loop system is studied by means of a sensitivity analysis. Effects of disturbances and uncertainty on the closed loop system are studied basing on analysis of the singular values of the sensitivity and the complementary sensitivity functions, the relative gain array and the minimized condition numbers. Finally, the sensitivity analysis is validated by the use of logged data from test operation at the coal injection plant at SSAB Tunnplåt Luleå, Sweden.

1 Introduction

Nowadays, iron producers are reducing production costs by replacing the expensive energy carrier coke by other cheaper alternatives. In Luleå, SSAB Tunnplåt AB is partly substituting coke by fine coal, which is 40% cheaper, in their iron production. The economical benefits of pulverized coal injection (PCI) are discussed in [1].

Since fine coal, in its pure form, is highly inflammable even under normal conditions, it is difficult to supply it to the process. Therefore, it is important to keep the fine coal isolated from the air, which can be done by using a pneumatic conveying device (see [2],[3]), where the transportation gas is nitrogen or at least has a higher rate of nitrogen compared to that of the air. The fine coal injection vessel is a part of a coal injection plant for a blast furnace, where fine coal is pneumatically conveyed to the blast furnace and finally injected at tuyere level. The coal injection plant at SSAB Tunnplåt AB in Luleå (Fig. 1), which has been used for the experiments, is described in more detail in [4].

A drawback of substituting coke by fine coal is that it can result in blast furnace instabilities if coal flow outages appear [5]. Hence, a tight and reliable control of the fine coal flow from the injection vessel to the blast furnace becomes necessary.

A combined model-based control and leakage detection system for a fine coal injection plant has been developed in [6]. The designed controller is a linear quadratic (LQ) optimal multivariable controller for the control of the fine coal flow out of the injection vessel. Since the controller is exposed to plant dynamics alternations because of components wear-out and replacement, as well as noise, it is necessary to analyse the sensitivity towards these effects. Typically, sensitivity analysis is used to obtain the necessary information. In [7] some schemes for sensitivity analysis in the multivariable case are given and used for the subsequent sensitivity analysis.

Following up results of experiments and test operation is an essential part of an industrial project, since controller performance specifications have to be checked. From a theoretical point of view such an analysis helps to improve controller designs and pinpoints possible shortcomings in the control strategy. It can also motivate further research in the area. This paper discusses such a follow-up in order to validate theoretical results of a sensitivity analysis.

The paper is organized as follows. In Section 2 the controller design is presented. The succeeding Section 3 discusses the sensitivity analysis and creates a framework for the follow-up. Finally, in Section 4, the acquired data from the tests is used to validate the anteceding analysis.

2 LQ optimal multivariable controller

The multivariable controller is a part of the loop structure depicted in Fig. 2 and is a result of a previous study, [6].

Besides the state vector feedback controller, a Kalman filter, a feedforward controller and an actuator saturation are present in the closed-loop system. Both Kalman filter and multivariable controller design are based on an identified multiple input, multiple output (MIMO) model of the process dynamics.

The structure of an injection vessel can be described by Fig. 3, and is principally a pressurized tank process. During the injection of coal the valves u_I and u_V are closed. Consequently, the two actuators u_N and u_C can be used to control the vessel. Measured outputs of the vessel are the net mass m of the vessel. The latter is identical to the sum of the nitrogen mass m_N and the coal mass m_C, and the pressure p in the vessel. Since the injection vessel is injecting coal at a certain flow rate, the net mass of the vessel has to follow a trajectory. Using direct identification, a model describing the process

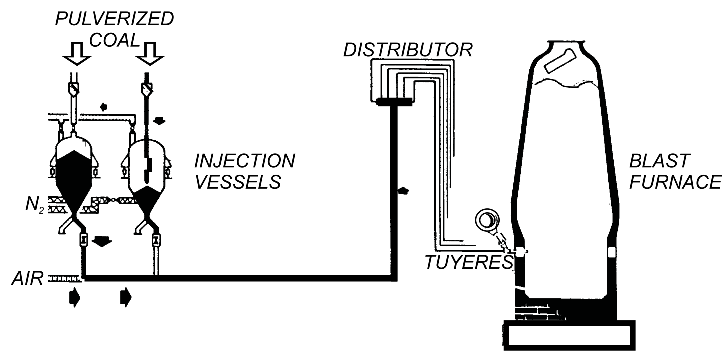

Fig. 1: Coal injection plant (injection vessels, distributor and blast furnace).

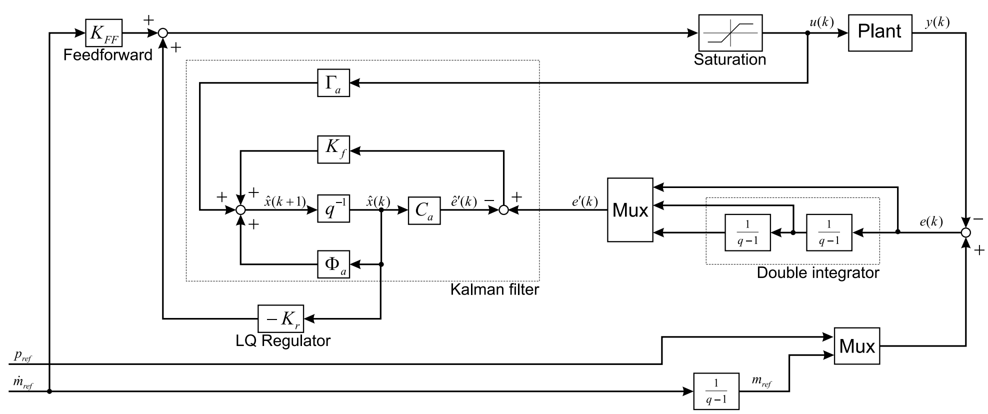

Fig. 2: Block diagram of the closed-loop structure

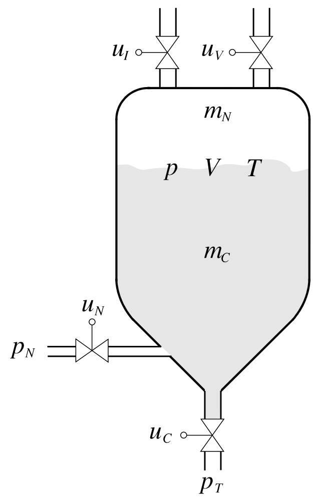

Fig. 3: Simplified injection vessel structure

dynamics can be obtained. The identification method and its application to the coal injection plant are discussed in [8], where the subspace identification method *n4sid* is applied to the Laguerre spectra of the input/output data. There, it has been shown that *n4sid* performs better in the Laguerre domain compared to the time domain, when it is used for identification of the injection vessels. The obtained MIMO model is of order two and given by

$$x(k+1) = \Phi x(k) + \Gamma \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} \tag{1a}$$

$$\begin{bmatrix} p(k) \\ m(k) \end{bmatrix} = Cx(k) + D \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix}, \tag{1b}$$

Since C is invertible, the similarity transformation $x(k) = C^{-1}x'(k)$ can be applied which yields

$$x'(k+1) = C\Phi C^{-1}x'(k) + C \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} \tag{2a}$$

$$\begin{bmatrix} p(k) \\ m(k) \end{bmatrix} = I_2 x'(k) + D \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} \tag{2b}$$

Hence, the states of the transformed dynamic system coincide with the outputs.

As mentioned above, the net mass of the injection vessel has to follow a trajectory, which is usually a ramp. Hence, the open-loop system (controller and plant) needs to include at least two integrators, in order to drive the steady state error to zero. For Kalman filter and controller design, the identified model is augmented with a double-integrator for each of the outputs. The resulting state space system is given by

$$\begin{aligned} x'_a(k+1) &= \underbrace{\begin{bmatrix} C\Phi C^{-1} & 0 & 0 \\ I_2 & I_2 & 0 \\ 0 & I_2 & I_2 \end{bmatrix}}_{\Phi_a} x'_a(k) + \\ & \underbrace{\begin{bmatrix} C\Gamma \\ 0 \\ 0 \end{bmatrix}}_{\Gamma_a} \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} \end{aligned} \tag{3a}$$

$$\begin{aligned} y(k) &= \underbrace{\begin{bmatrix} I_2 & 0 & 0 \\ 0 & I_2 & 0 \\ 0 & 0 & I_2 \end{bmatrix}}_{C_a} x(k) + \\ & \begin{bmatrix} D \\ 0 \\ 0 \end{bmatrix} \begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix}. \end{aligned} \tag{3b}$$

Using the standard LQG design procedure, a MIMO LQG controller with a stationary Kalman filter is obtained (see [4]). The optimal multivariable controller can be written in the form:

$$x_c(k+1) = \begin{bmatrix} I_2 & 0 \\ I_2 & I_2 \end{bmatrix} x_c(k) + \begin{bmatrix} I_2 \\ 0 \end{bmatrix} e(k) \tag{4a}$$

$$e'(k) = \begin{bmatrix} 0 & 0 \\ I_2 & 0 \\ 0 & I_2 \end{bmatrix} x_c(k) + \begin{bmatrix} I_2 \\ 0 \\ 0 \end{bmatrix} e(k) \tag{4b}$$

$$\begin{bmatrix} u_N(k) \\ u_C(k) \end{bmatrix} = -K_r \cdot e'(k), \tag{4c}$$

where $e(k) = r(k) - y(k) = \begin{bmatrix} p_{ref}(k) - p(k) \\ m_{ref}(k) - m(k) \end{bmatrix}$. The measurement signal, vector $y(k)$, contains the pressure $p(k)$ and the net mass $m(k)$, whereas the

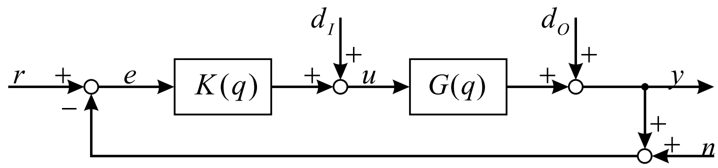

Fig. 4: Closed loop system for sensitivity analysis

reference signal vector contains the pressure set-point $p_{ref}(k)$ and the net mass trajectory $m_{ref}(k)$.

According to the *separation principle*, the Kalman filter and LQ controller dynamics are not coupled, which allows separate study of their dynamical behaviour. As the feedforward controller is designed for steady state and driven by an external signal (coal flow set point, $\dot{m}_{ref}(k)$), the influence of this controller on the closed loop dynamics can be neglected. Furthermore, it is assumed that the control signals are not saturated.

The resulting closed-loop structure for the controller and the plant is given in Fig. 4, where $K(q)$ denotes the controller in (4), $G(q)$ is the process model in (2) and q is the forward-shift operator. Additional to the loop structure in Fig. 2, disturbance and noise inputs are considered.

3 Sensitivity Analysis

Three sensitivity functions are considered:

- Complementary sensitivity function T, which is the transfer matrix from reference input r to the output y.
- Input sensitivity function S_I describes the transfer matrix from the disturbance input d_I to the control error e.
- Output sensitivity function S_O is similar to S_I but for the disturbance input d_O.

For the sake of simplicity the operator s is dropped.

Analysing the block structure in Fig. 4 the following sensitivity and complementary sensitivity functions are obtained:

$$T = GK(I+GK)^{-1} \tag{5}$$
$$S_I = -G(I+GK)^{-1} \tag{6}$$
$$S_O = -(I+GK)^{-1} \tag{7}$$

It can be mentioned that the sensitivity function for the noise input n to the control error e is identical to S_O, and the sensitivity function from the reference input r to the control error is equal to $-S_O$. Obviously, an analysis of these two more functions would not contribute with more information.

While the sensitivity and the complementary sensitivity functions are directly analysed in the scalar case, the quantities of interest in the multivariable case are the singular values of these functions. However, the role of the sensitivity and the complementary sensitivity functions in both cases are similar, since their magnitudes are usually used to measure stability robustness with respect to modelling uncertainties.

Of great interest in the sensitivity analysis is the supremum of the singular values of the sensitivity functions. High peaks can lead to instability of the closed loop system under perturbation and should be avoided. In the controller design, the characteristics of the singular values can be used to achieve a closed-loop system with minimized peaks in the singular values. Although the magnitudes of the sensitivity and complementary sensitivity functions are not a good measure for the gain of the MIMO system, information on the character of the cross-couplings in the MIMO system can be obtained, and exploited in the design process.

Finally, the sensitivity of the plant towards element-by-element uncertainty and input uncertainty is analysed using the relative gain array (RGA) and the minimized condition numbers for the plant and controller.

3.1 Input sensitivity function

Fig. 5 shows the singular values of the input sensitivity function. Obviously, the magnitude is very small compared with the other sensitivity or complementary sensitivity functions (Fig. 6, Fig. 7 respectively). Since disturbances of magnitude larger than one decade are unlikely to occur, the sensitivity to disturbances at the plant inputs can be neglected.

3.2 Output sensitivity function

Information on the bandwidth of the system with respect to output disturbance attenuation can be obtained from the singular value plot of the output sensitivity function (Fig. 6). Since the bandwidth for a multivariable system depends on the input/output directions, a fixed bandwidth value cannot be given. Therefore, the lowest bandwidth value for output disturbance attenuation should be chosen. To determine the bandwidth of a system the definition

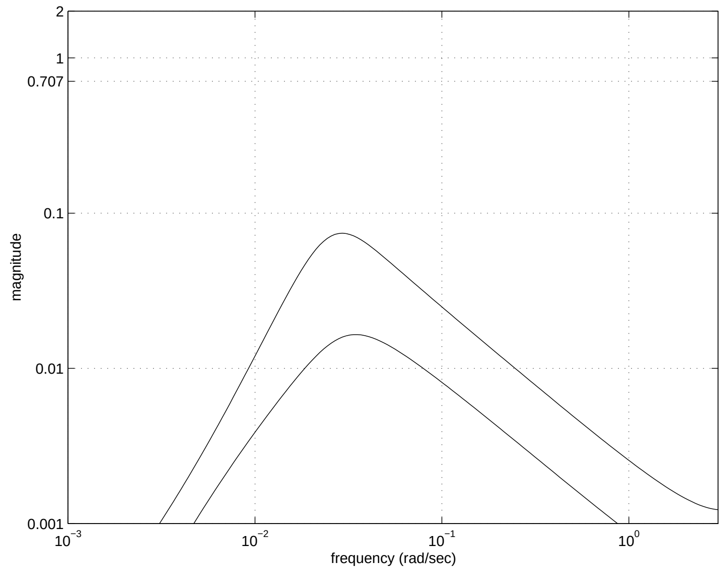

Fig. 5: Singular values of the input sensitivity function S_I.

in [9] is used. According to Fig. 6, disturbances up to 0.02 rad/sec can be attenuated.

Furthermore, the peak value of the sensitivity function (1.35) is a quite small value which indicates that output disturbances can lead to a slight overshoot in the transient behaviour.

3.3 Complementary sensitivity function

Similarly, the bandwidth for the reference tracking is defined, but here the higher value is chosen. Another important factor is the roll-off, *i.e.* a high negative slope above the cross-over frequency.

Fig. 7 shows the singular values for T. The bandwidth for reference tracking is about 0.095 rad/sec and the closed-loop system has a high negative slope of -2 above the cross-over frequency. The peak value of T is also rather small and is 1.85. According to [7], recommended peak values are less than two.

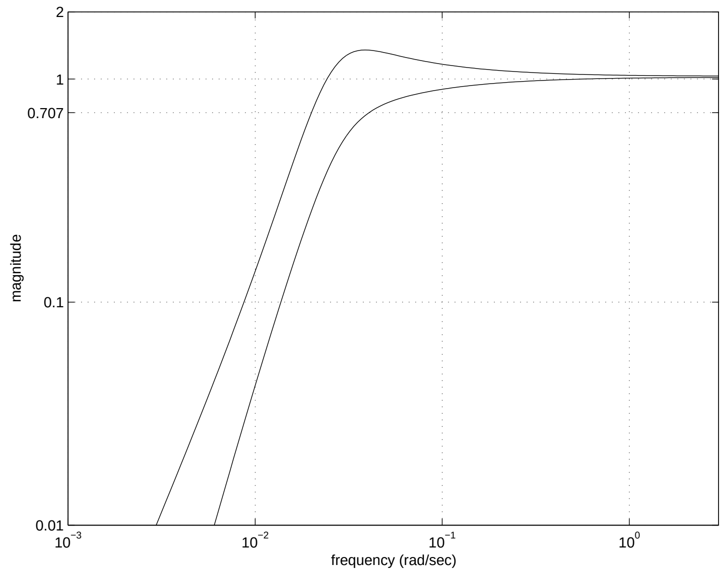

Fig. 6: Singular values of the output sensitivity function S_O.

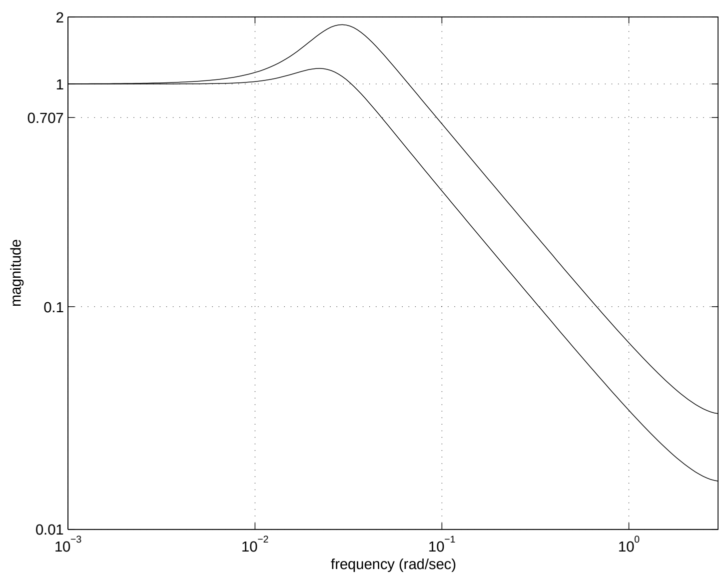

Fig. 7: Singular values of the complementary sensitivity function T.

3.4 Sensitivity to uncertainty

In [7] the RGA and the minimized condition numbers are used to evaluate sensitivity to uncertainty. In this case the MIMO model for design is obtained from direct identification of the plant dynamics. Because of non-linearities as well as disregarded dynamics, the model only approximates the plant behaviour. Hence, uncertainty has to be taken into consideration.

The RGA is given by

$$RGA(G) = G \times (G^{-1})^T,$$

where $\times$ is the Schur product. The RGA is computed at discrete frequency points. Here, the RGA is a symmetric 2×2 matrix of the form

$$\begin{bmatrix} a & b \\ b & a \end{bmatrix}.$$

Large values in the RGA indicate that G will become singular if the element g_{ij} in G is multiplied by a factor $(1 - \frac{1}{\lambda_{ij}})$, where λ_{ij} is the respective entry in the RGA. Thus, the RGA should contain small values.

In Fig. 8 the entries of the RGA are displayed. Because of the above given form of the RGA, only two values are plotted. The values are small and therefore the closed loop system should not be sensitive to element-by-element changes. An important property of the RGA is that sign changes of entries over the frequency axis indicate the presence of right half plane (RHP) zeroes in G or at least in one subsystem of G. As pointed out in [10] and [11], such non-minimum phase zeroes lead to fundamental performance limitations of the closed-loop system. Since one RGA entry is changing sign (see Fig. 8) at least one subsystem of G has a RHP zero and thus, performance limitations of the closed-loop system exist.

The minimized condition numbers $\gamma_I^*(G)$ and $\gamma_O^*(K)$ are a measure of robust performance to diagonal input uncertainty. According to [7], these condition numbers can be derived as follows:

$$\gamma_I^*(G) = \min_{D_I} \gamma(GD_I)$$

$$\gamma_O^*(K) = \min_{D_O} \gamma(D_O K),$$

where D_I and D_O are scaling matrices. The minimized condition number is the result of a minimization of the condition number over all possible scales.

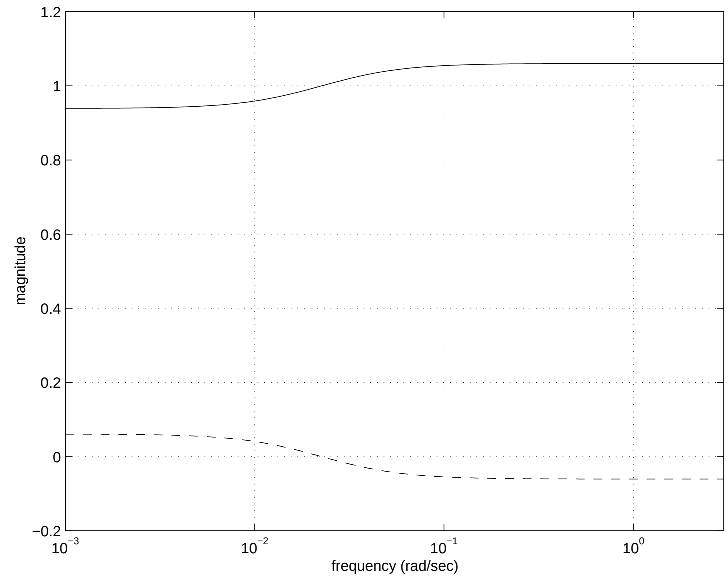

Fig. 8: RGA of G.

For the closed-loop system to be insensitive to input uncertainty, the values of $\gamma_I^*(G)$ and $\gamma_O^*(K)$ should be around 2 or smaller in the cross-over region. Fig. 9 shows both condition numbers. In the cross-over region (between 0.024 rad/sec and 0.065 rad/sec), the minimized condition numbers are obviously around 2 (between 1.94 and 2.76), which yields an relative insensitivity of the closed-loop system to input uncertainty.

4 Follow-up

The LQ optimal multivariable controller has been tested during a period of two weeks at the coal injection plant. During this period data has been logged. Since two injection vessels are injecting coal alternatingly, the time for one injection phase is limited and depends on the coal flow set-point. Thus, the frequency range is not only bounded from above with half the sampling frequency (π rad/sec) but also from below. The minimum recordable frequency is approximately 0.002 rad/sec.

Several injection series are randomly selected, and the power spectral density is estimated. Then, the expectation value for every frequency point is

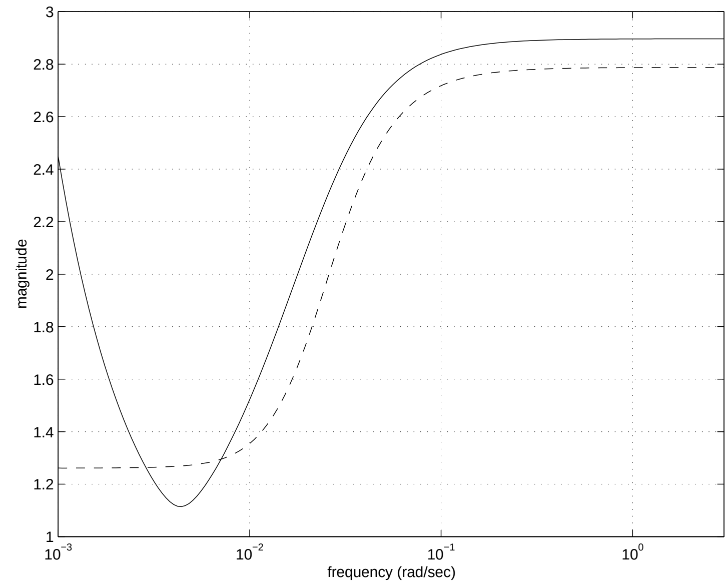

Fig. 9: Minimized condition numbers: $\gamma_I^*(G)$ solid, $\gamma_O^*(K)$ dashed.

estimated and plotted versus the maximum singular value of the output sensitivity and complementary sensitivity function (Fig. 10).

Obviously, the peaks in the maximum singular values of S_O and T are in the same frequency range as the ones in the power spectral density of the mass and the pressure deviation. Accordingly, the sensitivity analysis for S_O and T is validated by these results.

Since the input sensitivity function S_I contributes only with a comparably small magnitude, pure data analysis is not sufficient to validate the analysis. Therefore, a disturbance at the plant inputs has to be introduced in an experiment.

In practice, disturbances at the plant inputs are introduced by adding an extra signal to the input of the actuators, which can be achieved by sending a falsified position signal to the valve's position controller. Here, the position sensor signal is simply saturated before reaching its usual upper bound. Therefore, the transmitted position signal is incorrect if the valve has opened more than this artificially introduced upper bound. As a result the valve's position controller is reacting on a non-existing position error and thus, is opening the valve completely as soon as the physical position of the valve

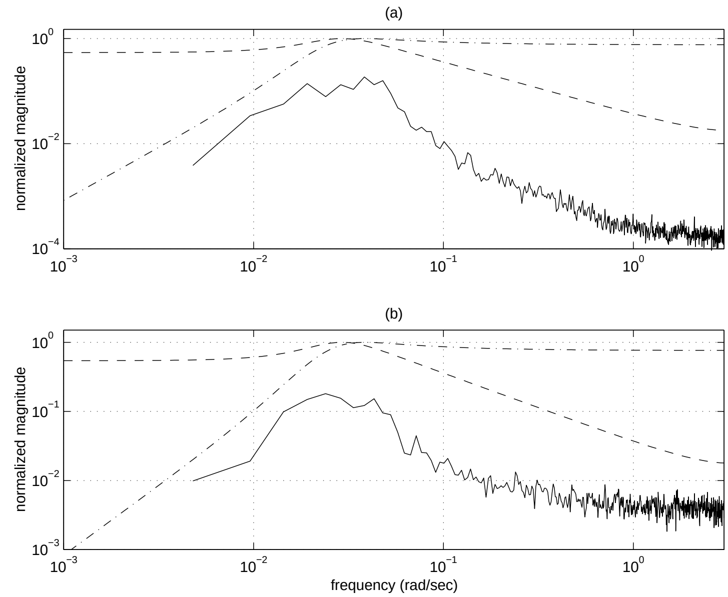

Fig. 10: Power spectral density (solid) versus maximum singular value of T (dashed) and S_O (dashed-dotted). (a) power spectral density of the pressure error, (b) power spectral density of the mass error.

exceeds the artificial bound.

Such a situation can occur when the valves position sensor malfunctions or a transmitting buffer amplifier saturates.

During the experiment, the closed-loop system did not loose its stability, but performance losses could be observed (see Fig. 11). In Fig. 12 the position signal from the valve and the control signal to the valves position controller are displayed.

Since these rather extreme input disturbances have not invalidated the analysis of the input sensitivity function, it can be concluded that the analysis is reliable.

In order to test the closed-loop sensitivity for input uncertainty and element-by-element changes in practice, the design is based on a model for one injection vessel but also run on the second injection vessel, which is equipped with a larger sized pressure control valve (u_N). During test-operation no performance losses due to this fact could be recognized.

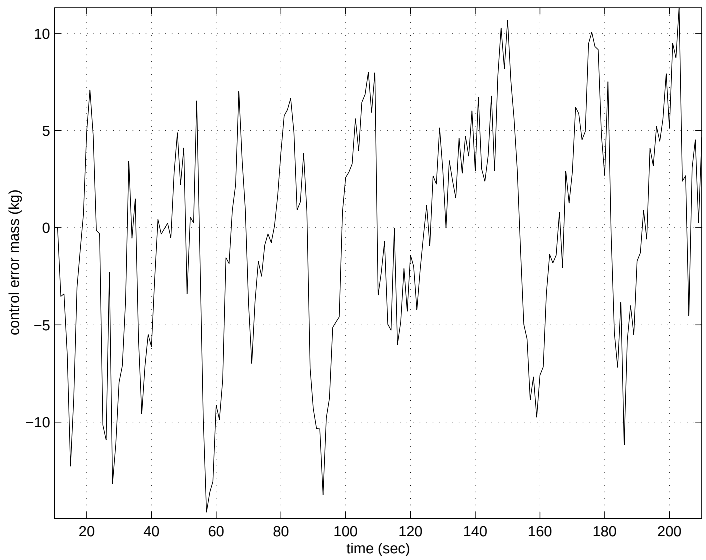

Fig. 11: Control error in the mass for the malfunctioning position control of the valve

5 Conclusions

In this paper an LQ optimal multivariable controller for a fine coal injection vessel is analyzed. The indications from the theoretical sensitivity analysis are validated through evaluation of data acquired during test operation of the controller at the coal injection plant at SSAB Tunnplåt Luleå, Sweden. It could be shown that the behaviour of the closed-loop system during operation is according to the results from the sensitivity analysis. Therewith, the controller design is validated and reliable enough to be considered for permanent installation. The commercially available control and leakage detection system *SafePCI* is now equipped with the above controller.

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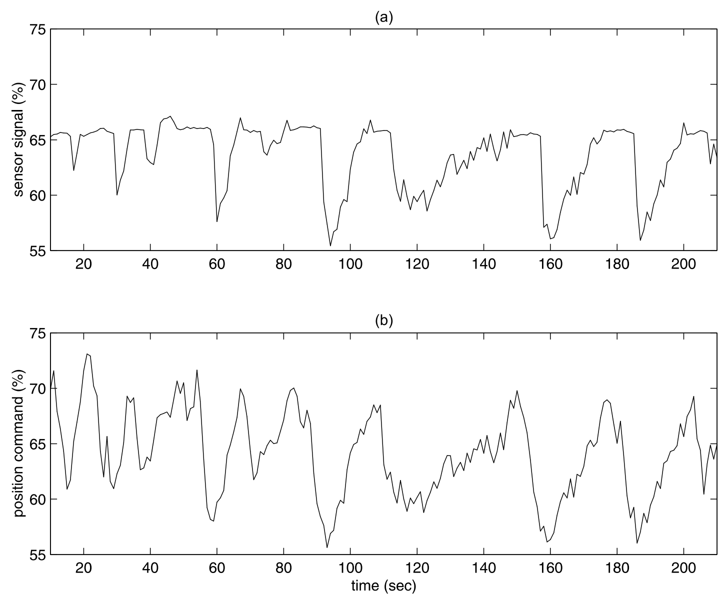

Fig. 12: Simulated malfunction of the position controller of the u_C valve. (a) position signal from the sensor, (b) control signal to the valve

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