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Practice and Challenges in Chemical Process Control Applications in Japan

Morimasa Ogawa * **Manabu Kano** **

* *Yamatake Corporation, Kurashiki, Okayama 712-8061 Japan (Tel: +81-86-448-2511; e-mail: ogawa-morimasa@jp.yamatake.com).*
** *Kyoto University, Nishikyo-ku, Kyoto 615-8510, JAPAN (Tel: +81-75-383-2687; e-mail: manabu@cheme.kyoto-u.ac.jp).*

Abstract:
This paper surveys how the three central pillars of process control – PID control, conventional advanced control, and model predictive control – have been used and how they have contributed to production activity from the viewpoint of the process control section in the Japanese chemical industry. In addition to introducing practical methods and their application results, the authors point out challenging problems, which include the development of a general model-based control technique to enhance batch process control.

Keywords: Process control; Industrial application; PID control; Conventional advanced control; Linear model predictive control; Non-linear model predictive control.

1. INTRODUCTION

This paper looks back on the projects that process control sections of a general chemical corporation of Japan have executed in the last two decades and introduces practical process control techniques successfully applied there, especially focusing on their results. In addition, challenges in the process control field, which are currently tackled jointly by engineers in industry and researchers in universities, are briefly described.

2. MILESTONE IN THE HISTORY OF PROCESS CONTROL APPLICATION

There are three phases in process control application projects in Mitsubishi Chemical Corporation (MCC), to which the first author had belonged for many years, as shown in Fig. 1. They are the advanced process control (APC) projects for large-scale continuous processes, the improvement activity of the control performance of basic control systems for small-to-medium-scale processes, and the advancement of polymer and batch process control.

2.1 Project Chronology

In the first phase in the early 90's, multivariable model predictive control (MPC) was applied to large-scale continuous processes such as olefin production units for generating a large profit. The APC project was conducted for 15 production units of 5 production sites by using DMCplus® as a standard tool, and satisfactory results were achieved. The key to success is nurturing process control engineers who can accomplish the projects independently on their own. They learned procedures and methods of planning, control system design, plant tests, tuning, and operation. In addition, they joined seminars on advanced control theory given by prominent researchers and professors. By

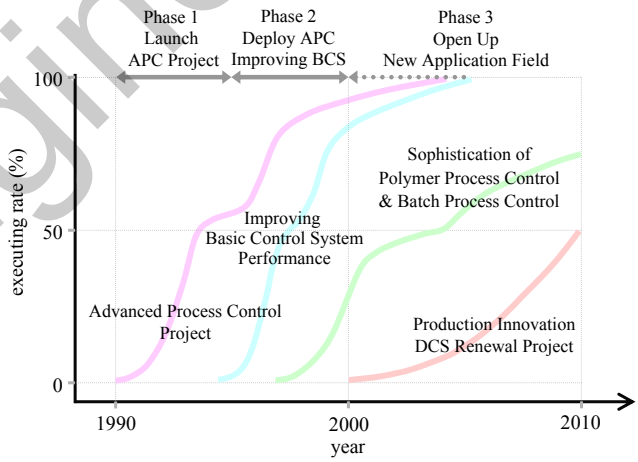


Fig. 1. Chronology of project execution in MCC

accumulating experience on the projects, they grew into capable engineers who understood theory and had business acumen. These 15 process control engineers took a leading part and accomplished APC projects.

In the second phase, the performance of PID control systems was assessed and improved. All production units which APC projects did not cover were targeted. Both the operation section and the instrumentation section jointly carried out this project as a daily improvement activity in cooperation with the process control section. As a result, the operator workload was reduced through the improvement in service factors of PID control systems and a reduction in frequency of alarms and operator interventions. In addition, the improvement in control performance contributed toward the economic profit because it made energy-efficient operation possible by changing setpoints. It was also the perfect opportunity for finding applications

Table 1. Process control applications

classification	methodology	application
modern advanced control	linear MPC non-linear MPC LQI with preview action	54 2 2
conventional advanced control	feed-forward control override control valve position control analyzer feedback control model-based control etc	500+
regulartory control	PID/I-PD control	5006

of conventional advanced control such as override control and valve position control (VPC).

In the third phase, the advancement of polymer process control was investigated. It is important to achieve rapid grade transition while satisfying quality specification in polymer plants, because transitions among a wide variety of products are made frequently. Therefore, an original control algorithm that is based on precise first-principle models of polymerization reactions and quality models relating polymerization reaction conditions and product quality has been used since the 80's. In this phase, process models such as catalyst activity were reviewed, and a new nonlinear MPC algorithm was developed and applied. As a result, the control performance was significantly improved, off-specification products were reduced, and quality was stabilized.

The focus of the process control section has shifted to problem-solving regarding process control of small-to-medium-scale processes and the maintenance of APC systems. The targets include 1) accumulating energy-saving effects by applying an in-house linear MPC algorithm to distillation, reforming furnace, and air separation processes, 2) developing soft-sensors, which are substituted for process gas chromatographs, for shortening the control period and improving control performance, and 3) adapting APC systems for reinforcement of process units.

Recently, the activity which reforms the whole production activity has started at advanced chemical companies. In addition to integration of control rooms, the movement that reviews operation management, alarm management, emergency shutdown system, maintenance management, and so on, and modernizes the control information system is becoming active. Such an activity is triggered by the opportunity for DCS introduced in the 80's to enter a renewal period. Process control engineers will be involved in this movement.

2.2 Process Control Methodology

Control methodologies which bear the central role in process control systems can be classified into PID control, conventional advanced control such as feedforward control and override control, and MPC. The number of applications of these control methodologies in the MCC Mizushima plant is summarized in Table 1. The ratio of applications of PID control, conventional advanced control, and MPC is 100:10:1. PID control is used in 5006 loops in 24 production units. The number of control loops repeatedly increases and decreases corresponding to new establishment, reinforcement, or stopping of production units. Conventional advanced control is effective in many cases, but the number of its applications is not as many as expected. MPC has become established as a standard technique for multivariable control which realizes economical operation of large-scale processes.

3. IMPROVING PID CONTROL PERFORMANCE

In Japanese chemical companies, KAIZEN activities aimed at safe and stable operation are actively continuing. One important activity is improvement in the control performance of PID control systems. The aim of this improvement activity, in which controllers are retuned appropriately, is (1) to realize stable operation by reducing the influence of disturbances, (2) to realize automatic rapid transition of operating conditions such as production rate, (3) to gain the ability to achieve economical operation, and (4) to allow operators to be released from taking care of PID controllers. Additional effects are to find out problems with sensors and actuators, and to clarify possible targets of advanced control application.

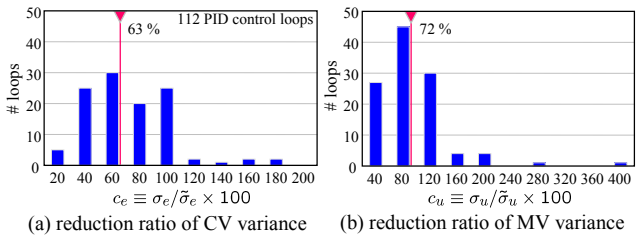
In the KAIZEN activities, improving the control performance with retuning should be stressed, rather than spending time and effort to strictly assess the control performance of PID control loops. The following simple indexes are sufficient to determine good or bad control performance. (1) Is the controller in auto mode at all times? (2) Are PID parameters in the proper range? (3) Is fluctuation of the controlled variable and the manipulated variable sufficiently small? (4) Is the PID tuning agreeable to the control purpose such as flow-averaging level control? Other than these, it is necessary to check the range propriety of sensors and actuators, the necessity of filtering of measurement noise, the presence of stiction of control valves, and so on.

Experience leads us to believe that 80% of PID control loops can be successfully tuned with a method based on rule of thumb and trial and error. For example, initial settings for PID parameters should be "wide proportional band and fast reset time" for flow control and "narrow proportional band and slow reset time" for level control. After the initial PID setting, PID parameters are tuned gradually to strengthen control action while verifying the control performance.

The control performance improvement activity introduced in this section has attracted the attention of many enterprises in chemical and oil refining industries in Japan, and the number of enterprises starting this activity has increased rapidly. Such a movement seems to be the result of the process control section not directly recognizing the reality that the operation section had an awareness of control performance issues and was dissatisfied with the control performance.

3.1 Actual Project Example

The result of a project on a large-scale monomer plant, which has 190 PID control loops, is introduced here. In this plant, 90% of the PID controllers were in auto mode for 30 days. This value outperforms the average of 70% in the literatures (Desborough and Miller [2001], Ender [1993]). Operators had adjusted PID parameters to realize



(a) reduction ratio of CV variance (b) reduction ratio of MV variance

Fig. 2. Improvement in PID control performance

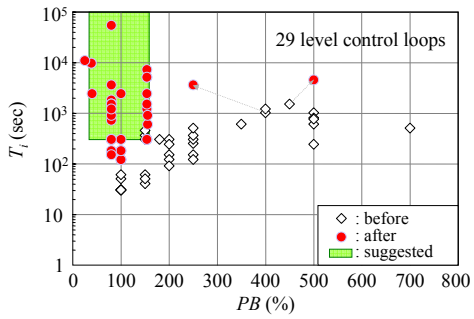


Fig. 3. PI parameters of level controllers

very loose control action. As a result, since the process was easily affected by disturbances and a long time was required for production rate changes, the operators made frequent adjustments such as setpoint changes and manual operation.

In all, 112 loops having a margin of improvement in control performance were retuned in 12 days. The variance of controlled variables (CVs) and manipulated variables (MVs) was reduced by an average of 37% and 28%, respectively, as shown in Fig. 2. The reduction is almost the same as the value reported by Shah et al. [2004]. A pronounced effect was achieved in tray temperature control loops of distillation columns. Temperature fluctuation was reduced to one-fourth up to one-seventh, and composition was also stabilized.

Fig. 3 shows PID parameters for 29 level control loops before and after retuning them. With the exception of a part such as six loops for a heat recovery boiler, the purpose of these control loops is flow-averaging level control. Operators made the proportional gain small (wide proportional band) in order for the manipulated variable not to change. However, the manipulated variable had been oscillatory due to small reset time. To solve this problem, Ogawa et al. [1998] developed a design method of flow-averaging level controllers and applied it to those loops. This flow-averaging level control, explained in section 3.3, was very effective for decreasing changes in feed/product flow rate to distillation columns and lightening the burden of tray temperature control.

3.2 Robust I-PD Controller Tuning

Since most PID controllers have the I-PD algorithm, Ogawa and Katayama [2001] derived a robust model-based PID tuning method for the I-PD controller shown in Fig. 4. This method is suitable for specific control loops such as temperature and composition control, which are required a

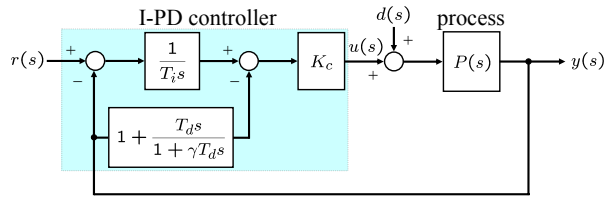


Fig. 4. I-PD control system

proper control performance in the presence of plant-model mismatch.

Here, the I-PD controller tuning method for a first-order plus time-delay (FOPTD) model is explained. The desired response $W_r(s)$ of the controlled variable y for the setpoint r is specified by

$$W_r(s) \equiv \frac{y(s)}{r(s)} = \frac{1}{(1+T_F s)^n} e^{-T_L s} \tag{1}$$

where T_F denotes a tuning parameter and $n = r + 1 = 2$ for the relative order $r = 1$ of the process model. T_L, T_p, and K_p denote time-delay, time constant, and steady-state gain of the process model, respectively. By using the 1/1 Pade approximation and ignoring the derivative filter, the partial model matching method (Kitamori [1981]) provides the following PID setting rule.

$$K_c = \frac{p - 2q + 4}{K_p\,(p + 2q)} \tag{2}$$

$$T_i = \frac{(p + 2q)\,(p - 2q + 4)}{2p + 4} T_p \tag{3}$$

$$T_d = \frac{p\,(p + 4q - 2q^2)}{(p + 2q)\,(p - 2q + 4)} T_p \tag{4}$$

where $p \equiv T_L/T_p$ represents the difficulty of control and $q \equiv T_F/T_p$ is a tuning parameter. Although the parameter q can be tuned so that ISE (Integral of Squared Error) is minimized, such tuning is not preferable in practice. To realize robust PID control that is intuitive and practical, a constraint on the maximum change of the manipulated variable $u(t)$ against a stepwise setpoint change is introduced. Given $U_{\max}(\%)$, the parameter q is determined by solving the following equation.

$$\max_q \|u(t)/u(\infty)\|_\infty \le U_{\max}/100 \tag{5}$$

where $u(\infty)$ is the steady-state value of $u(t)$ after the setpoint change. The relationship among q, p, and $U_{\max}$ is shown in Fig. 5.

This robust I-PD controller tuning method is derived not only for FOPTD models but for integral plus FOPTD models and SOPTD models with/without an unstable pole.

3.3 Flow Averaging Level Control

Consider a process described by

$$P(s) = \frac{y(s)}{u(s)} = \frac{1}{T_p s}, \quad T_p = \frac{K_m A}{K_u} \tag{6}$$

where T_p (h) denotes reset time constant, K_m (m/%) sensor gain, K_u (m^3/h/%) actuator gain, and A (m^2) sectional area.

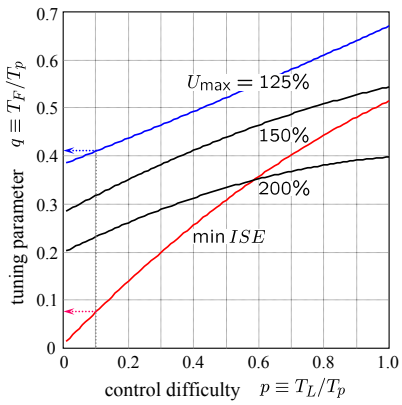


Fig. 5. Tuning of robust I-PD controllers

I-P control is used for flow averaging level control (FALC). Its block diagram is shown in Fig. 4 and derivative time T_d is set equal to 0. The control response to setpoint r (%) and disturbance d (%) becomes the following second-order standard form.

$$y(s) = \frac{1}{1 + 2\zeta T_n s + T_n^2 s^2}\left(r(s) + \frac{T_i s}{K_c} d(s)\right) \tag{7}$$

The damping coefficient ζ and the natural frequency T_n are given by

$$\zeta = \sqrt{\frac{K_c T_i}{4T_p}}, \quad T_n = \sqrt{\frac{T_p T_i}{K_c}} \tag{8}$$

By defining the performance index of FALC under a step-wise disturbance as

$$\min \ J = \frac{1}{2}\int_0^\infty \left(q^2 y^2(t) + r^2 \dot{u}^2(t)\right) dt \tag{9}$$

and solving the optimization problem, the control parameters can be related to the process parameter.

$$K_c T_i = 2T_p \tag{10}$$

As a result, the damping coefficient becomes $\zeta = 1/\sqrt{2}$ and the second-order standard form becomes Butterworth-type.

Given the size of the step-wise disturbance d_s and the maximum allowable level change y_s, the proportional gain and the reset time can be determined as follows:

$$K_c = \frac{\sqrt{2}e^{-\pi/4}}{y_s/d_s} \approx \frac{0.645}{\eta}, \quad T_i = \frac{2T_p}{K_c} \tag{11}$$

Here, $\eta \equiv y_s/d_s$ is the disturbance rejection ratio.

4. CONVENTIONAL ADVANCED CONTROL

Conventional advanced control is effective for various processes and easy to implement on DCS. However, there has been a trend for control engineers to take little account of its application. This is the result that MPC became a standard tool for the advancement of process control. However, there is no doubt that production cost can be decreased by accumulating the effect of conventional advanced control.

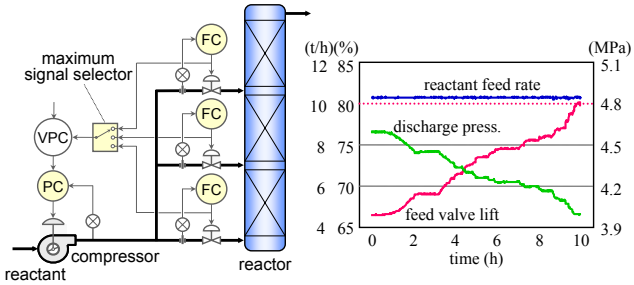


Fig. 6. Compressor power saving control

The application of energy-saving control of the compressor with VPC is described here. As shown in Fig. 6, the feed gas is pressurized with the turbo compressor and supplied to three different stages of the reactor. Each flow rate of the feed gas is controlled. The discharge pressure of the compressor is controlled by using guide vane opening as the manipulated variable. To reduce compressor power, the discharge pressure is lowered gradually with VPC, until the largest valve opening among three feed flow control valves reaches the upper limit, while feed flow rate is kept constant. In this application, the discharge pressure was decreased from 4.6 MPa to 4.0 MPa by increasing the largest valve opening from 67% to 80%. As a result, motor electric power consumption was saved by 16%.

In the enterprise, it is important to find any loss that usually has been overlooked, to make the most use of conventional advance control, and to continue the effort at minimizing the loss. In comparison with the APC project of using MPC, profitable results can be obtained much more quickly without spending any expense.

5. MODEL PREDICTIVE CONTROL

5.1 Linear MPC

The process that MPC is applied to most is distillation. A simple example of MPC for a distillation process is shown in Fig. 7 (a). The controlled variables are the purity of products extracted from the column top and bottom, and the manipulated variables are the setpoints of temperature PID control at the column top and bottom. The disturbance variables are flow rate and composition of feed. The constraints are upper and lower limits of the manipulated variables and the controlled variables and upper limits of changes in the manipulated variables.

The economic benefit that MPC brings is illustrated in Fig. 7 (b). Since the achievable performance of PID control is limited due to interaction, which is a feature of multi-variable processes, it is assumed that the current operating region corresponds to region A in the figure. In such a situation, the operating condition bound has to be set far from the real constraints to ensure a sufficient margin of safety. Using MPC can improve control performance and reduce variation. As a result, the operating region becomes small from A to B. This improvement makes it possible to move the operating region from B to C, which is close to the bound of operating conditions. Furthermore,

more economical operation D can be realized by optimizing setpoints to minimize operational costs. MPC takes on the responsibility of this set of functions. The benefit is not only the improvement of the control performance by using model-based control, but also the realization of stable operation close to the optimal point under disturbances by using optimization.

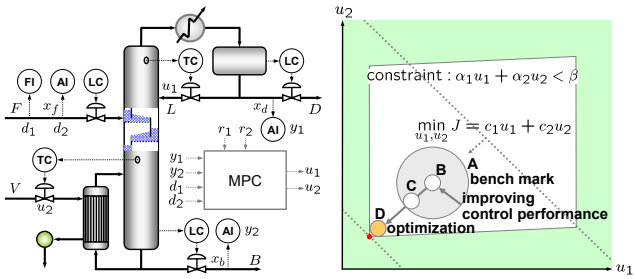
Implementation of MPC releases operators from most of the adjustment work they had to do in the past because the optimal operating condition is automatically determined and maintained under disturbances. In addition, MPC makes it possible to maximize production rate by making the most use of the capability of the process and to minimize cost through energy conservation by moving the operating condition toward the control limit. Both the energy conservation and the productive capacity were improved by an average of 3 to 5% as the result of APC projects centering on MPC at MCC.

The control performance of MPC depends on the accuracy of the process model and the appropriateness of tuning, but MPC has outstanding robustness. For example, stable operation is realized by MPC in spite of large model parameter errors of about 50%. However, it is difficult to assess the control performance of MPC due to a large number of variables. A plant test for modeling sometimes requires two weeks. The engineers who have experienced it can readily understand that the implementation of MPC including modeling and tuning is a demanding job.

MPC is highly effective, but it has several weak points (Hugo [2000]). First, it is not good at level control when the process has an integrator. For such a case, PI control is easy to design and superior to MPC in control performance. Second, the control performance of MPC deteriorates against ramp-wise disturbances because the MPC algorithm is developed by assuming step-wise disturbances. In addition, linear programming (LP) is usually used for optimizing setpoints under constraints, and the optimal point is located at one of the extreme points of a polyhedron consisting of linear constraints. When the gradient of the objective function and that of constraints are similar to each other, the optimal point jumps from one extreme point to another and the setpoints change suddenly. Research and development is continued to solve these problems.

Ohshima et al. [1995], who wrote about the state of MPC application by the petroleum and chemical enterprises in Japan, reported that 154 MPC controllers were in operation and 43 under implementation. The total number of 197 was 2.5 times as much as the number of 75 in 1990. At present, the number of MPC controllers is 169 only at MCC (Ogawa [2006]). An updated report would be expected.

At the very end of this subsection, the MPC application for energy conservation and production maximization of the olefins unit at MCC Mizushima plant is briefly explained (Emoto et al. [1994]). Qin and Badgwell [2003] reported that this application was the largest MPC application in the world, consisting of 283 manipulated variables and 603 controlled variables. The process was operated in energy conservation mode for the first four days in Fig. 8. Since the productive capacity was beyond the demand, the temperature difference between vapor and coolant in the overhead condenser was increased by making the column pressure higher. As a result, an amount of heat exchanged was increased, and the amount of coolant used was decreased. This operation made it possible to reduce the refrigerator power. On the other hand, the process was operated in production maximization mode for the last five days. To maximize the production rate for fulfilling the demand, the separation performance was improved by decreasing the column pressure and increasing the relative volatility. The feed flow rate to the cracking furnace was increased until the tray delta-pressure reached its upper limit, that is, the flooding limit. In this production maximization mode, the MPC system is large because MPC controllers for many cracking furnaces and distillation columns function in cooperation.



(a) MPC for distillation column (b) optimal operation using MPC

Fig. 7. MPC for distillation process

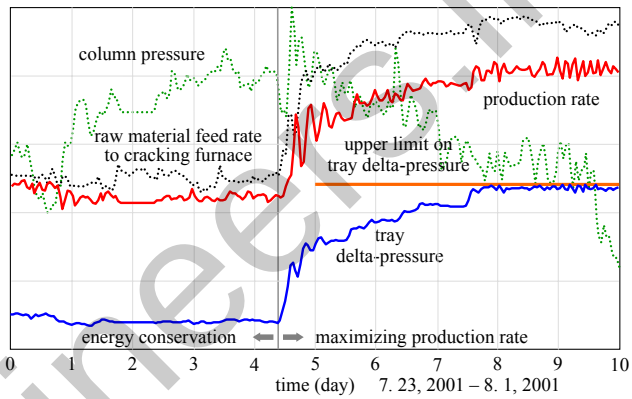


Fig. 8. Actual performance of the large scale MPC in the olefins unit

A skilled operator made the following comment on this MPC application: "We had operated the Ethylene fractionator in constant pressure mode for more than 20 years. I was speechless with surprise that we had made an enormous loss for many years, when I watched the MPC decreased the column pressure, improved the distillation efficiency, and maximized the production rate." Another process control engineer said "I had misunderstood that setpoints were determined by operation section and process control section took the responsibility only for control. I realized MPC for the first time; it makes the most use of the capability of equipments, determines setpoints for economical operation, and maintains both controlled variables and manipulated variables close to the setpoints."

5.2 Non-linear MPC

Nonlinear MPC has attracted attention in recent years (Qin and Badgwell [2003]). It is suitable for control of a nonlinear process operated in a wide range, e.g. polymerization reaction processes. In MCC, an independently developed nonlinear MPC has been applied to polymerization reactors at the polyolefin production units, and it has been put successfully to practical use (Seki et al. [2001]).

However, application of nonlinear MPC has not spread as well as was expected. It is difficult to build a nonlinear model of a process, or process control engineers have slackened their efforts at modeling nonlinear processes. On the other hand, most polymer production processes are operated without any quality problem by existing control systems supported with operators' suitable manual intervention. Therefore, it is difficult to justify any benefit of using nonlinear MPC. These obstacles should be overcome to expand nonlinear MPC application.

6. CHALLENGING PROBLEMS AND CONCLUSIONS

In Japan, a task force was launched in July 2007 to sift through problems regarding process control and investigate solutions. The task force, which consists of 30 engineers from industry and 12 researchers from universities, is supported by the 143rd committee on process systems engineering, the Japan Society for the Promotion of Science (JSPS). Currently, the following problems are listed by the members.

- Practical closed-loop system identification
- Practical tuning techniques of PID controllers
- Systematization of the control performance improvement activity based on control performance assessment
- Control system design from the viewpoint of plant-wide control
- Evaluation and maintenance of model predictive control tools
- Design and maintenance of soft sensors

In recent years, there has been a strong trend to produce polymer products having special functions in a small amount in a batch process. At the forefront of production, the necessity of practical technological development is being recognized, such as precise control of reaction temperature, estimation of reaction state, and batch-to-batch control. Process control engineers have been committed to continuous process control so far. In the future, however, they need to open their eyes to batch process control and to meeting the challenges to its advancement.

Young control engineers are eager for training in concrete techniques based on theory and experience to solve actual problems by themselves. On the job training (OJT) and self-enlightenment are not sufficient to rear them. In Japan, the activity aimed at growing engineers, who carry the destiny of future process control on their shoulders, has started through the united efforts of experienced engineers in industry and outstanding university researchers. The authors are looking forward to the result.

This paper has surveyed what process control engineers have done in the last two decades and what they might do in the future, focusing on the projects at a Japanese chemical company. The authors expect that engineers share practical methods and best practice and also that they spare no effort in developing their own methods to solve their own problems.

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REFERENCES

L. Desborough and R. Miller. Increasing customer value of industrial control performance monitoring - Honeywell's experience. *Preprints of Chemical Process Control - 6, Tucson*, pages 153–186, 2001.

G. Emoto, Y. Ohta, H. Matsuo, M. Ogawa, D.B. Raven, R.F. Preston and J.S. Ayala. Integrated advanced control and closed-loop real-time optimization of an olefins plant. *Proceedings of IFAC ADChEM1994, Kyoto*, pages 97–102, 1994.

D. Ender. Process control performance: Not good as you think. *Control Engineering*, pages 180–190, September 1993.

A. Hugo. Limitations of model predictive controllers. *Hydrocarbon Processing*, pages 83–88, January 2000.

T. Kitamori. A design method for nonlinear control systems based upon partial knowledge about control objects. *Preprints of 8th IFAC World Congress, Kyoto*, Paper no. 17.5, IV25-IV30, 1981.

M. Ogawa. Process control practice: the key to stable operation for maximizing the profit. *Proceedings of KIChE 2006 Annual Meeting, Seoul*, October 27, 2006.

M. Ogawa, M. Kano and I. Hashimoto. Design of flow averaging level control system using I-P controller with gap. *Journal of the Society of Chemical Engineers, Japan*, volume 24, no. 2, pages 259–265, 1998.

M. Ogawa and T. Katayama. A robust tuning method for I-PD controller incorporating a constraint on manipulated variable. *Transactions of the Society of Instrument and Control Engineers, Japan*, volume E-1, no. 1, pages 265–273, 2001.

M. Ohshima, H. Ohno and I. Hashimoto. Model predictive control experiences in the university-industry joint projects and statistics on MPC applications in Japan. *Proceedings of the international workshop on predictive and receding horizon control, Korea*, October, 1995.

S.J. Qin and T.A. Badgwell. A survey of industrial model predictive control technology. *Control Engineering Practice*, volume 11, pages 703, 2003.

H. Seki, M. Ogawa, S. Ooyama, K. Akamatsu, M. Ohshima and W. Yang. Industrial application of a nonlinear model predictive control to polymerization reactors. *Control Engineering Practice*, volume 9, pages 819–828, 2001.

S.L. Shah, W. Mitchell and D. Shock. Challenges in the detection, diagnosis and visualization of controller performance data. *Preprints of Control Systems - 2004, Quebec-City*, 2004.