Multi-Loop Control of Temperature for TV Glass Furnace

Un-Chul Moon

Department of Electrical Engineering
Woosuk University,
WhoJoung-Li 490, SamYe-Op
WanJu-Koon, JeonBook, Korea, 565-701
Tel) 82-63-290-1478 Fax) 82-63-291-9312
E-mail) ucmoon@woosuk.ac.kr

Kwang. Y. Lee

Department of Electrical Engineering
The Pennsylvania State University
103 Electrical Engineering East
University Park, PA 16802, USA
Tel) 814-865-2621 Fax) 814-865-7065
E-mail) kwanglee@psu.edu

***Abstract* - This paper presents a practical application of the conventional multi-loop control method for the temperature control of a television glass furnace. To control the temperature of the TV glass furnace, major input-output variables are selected, and simple First-Order-Plus-Dead-Time(FOPDT)models are established with the process experimental data. Based on the FOPDT models, a multi-loop control, which is a combination of cascade and single loops, is used. Practical implementation using the Distributed Control System (DCS) showed satisfactory results of the proposed control algorithm for the 150-ton/day-production furnace.**

I. INTRODUCTION

In spite of the mature development of modern control theory, conventional multi-loop control techniques that contain feedforward, cascade, override, selective, decoupling, etc. is one of the most popular control method in process industry [1]-[6]. Conventional multi-loop controls have advantages that cost is relatively low, design and tuning procedures are relatively simple, and process operators can understand its behavior easily. Therefore, it is possible for the operator to deal with the changing plant condition. These advantages are mainly from the fact that the conventional controller is designed with simple input-output models of the process instead of the detailed mathematical model. Because of this advantage, the conventional multi-loop control has been applied to various plants.

However, careful analysis and clear understanding of the process are necessary to apply the conventional multi-loop control to real plants. Furthermore, much trial and error approach should also be performed, especially to a Multi-Input Multi-Output (MIMO) process, which has strong interactions among variables.

In this paper, a practical application of the multi-loop control is introduced to control the temperature of television glass furnace. Haber, *et al.* [7] presented a modeling and control of a glass melting furnace. They elaborated a 3-input 3-output model with experimental data and applied a self-tuning regulator to the glass level control. Aoki, Kawachi and Sugeno [8] proposed two application methods of fuzzy logic control to a glass-melting furnace. They used estimated plant output in the fuzzy PI controller and modified the fuzzy control action based on the estimated plant output to overcome the dead-time characteristics of the furnace. Hadjili, *et al.* [9] presented an identification procedure of fuzzy models for a glass furnace process, and investigated the Takagi-Sugeno fuzzy system for the gas input in the molten glass temperature control. Moon and Lee [10] suggested a practical integration method with fuzzy logic and PI control for the temperature control of a glass furnace.

However, all of the phenomena that occur in the glass-melting furnace have not been explained fully because of their complexity. The furnace has many kinds of unmeasurable disturbances and 3-dimensional stream in molten glass. Therefore, it is still hard to establish a mathematical model, which reflects the chemical and mechanical characteristics of the furnace. This kind of furnace management is still delegated to a furnace operator

II. TV GLASS FURNACE

A. Overview of a TV Glass Furnace

Figs. 1 and 2 show a structure of a typical TV-glass furnace. A glass furnace has two rooms, a melter and a refiner, and two regenerators are connected with the melter. Two regenerators, which have 6-port, are shown in the upper and lower rectangles of the melter in Fig. 1. The melter and regenerator are hexahedral structures, and the refiner is a cylindrical structure. Usually, the size of melter is about 10×40×5 meters. The glass melting furnace has left-right symmetric structure as shown in Fig. 1.

The raw material fed into the melter is changed in its chemical characteristics to become molten glass and flows into the direction of refiner. Each regenerator is connected with a melter through 6 ports in Fig. 1; hence, Figs. 1 and 2 represent a "6-port furnace". Each port supplies either fuel and preheated combustion air, or cooling air into the melter. The heat energy to the furnace is supplied by burning the Bunker-C oil (BC). The internal temperature of the melter ranges from 1000 °C to 1500 °C depending on the location. In the refiner, temperature of molten glass is regulated to a proper level and impurities in the molten glass are settled to the bottom for product forming.

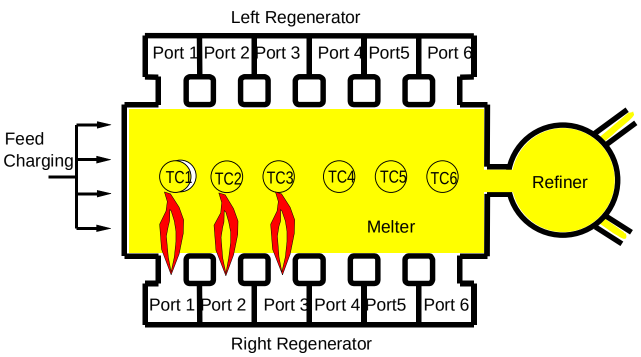

Fig. 1. Birds eye section structure of a glass furnace.

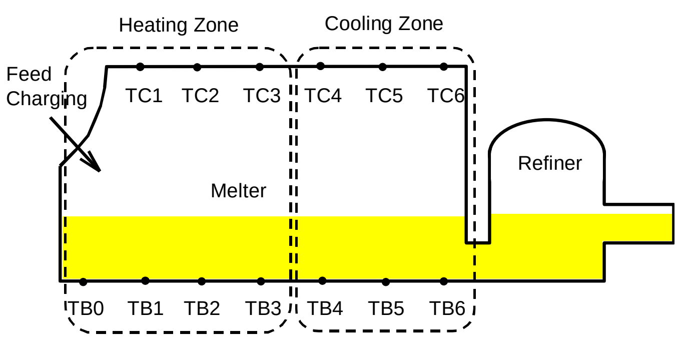

Fig. 2. Longisection structure of a glass furnace.

The firing and air flow directions are reversed from left to right at every 20 minutes. This change is called a reversing. Air for combustion is charged through the hot bricks of regenerator, which was used as a discharger of the waste gas during the previous half cycle. Heat contained in waste gas leaving the furnace is used to preheat this air, which leads to higher flame temperature and better heat transfer to the raw materials in the furnace.

In Fig. 1, the ports 1, 2 and 3 are heating ports which supply the fuel and combustion air, and the ports 4, 5 and 6 are cooling ports which supply the fresh air into the melter to cool the molten glass. While the direction of flame and combustion air of the ports 1, 2 and 3 is reversed at each half cycle, the fresh air without preheating is supplied from the ports 4, 5 and 6 in both sidewalls at every cycle. For this reason, the ports 1, 2 and 3 are called heating zone ports and the ports 4, 5 and 6 are called cooling zone ports in Fig. 2. As a result, the raw material is melted into the molten glass in the heating zone, and the molten glass is cooled in the cooling zone and flows into the forming process through throat, refiner and fore hearth.

Figs. 1 and 2 show the locations of the crown and bottom thermocouples. In Fig. 2, the crown thermocouples and bottom thermocouples are shown as TC (Temperature Crown) and TB (Temperature Bottom), respectively. The crown thermocouple indicates the radiation energy in the melter. The bottom thermocouple indicates the temperature of molten glass.

B. Manual Operation of a Furnace

The temperatures of TCs and TBs at normal condition are increased starting from the feed charging end (port 1) in the direction of port 2, and decreased starting from port 2 in the direction of port 6. For this reason, TC2 and TB2 are called as Hotspots. They serve as the upper and lower limits for all TCs and TBs, respectively. The furnace operators work to maintain the temperature profile to be within the upper and lower limits in all sensors, by adjusting the heating and cooling of the melter.

The BC oil in ports 1, 2 and 3 are adjusted proportionally with a constant ratio, instead of adjusting them independently. That is, only the total amount of BC is controlled, which then is distributed with a predetermined ratio to keep the temperature profile. Amount of cooling air (CA) in ports 4 and 5 is fixed and only the amount of CA in port 6 is adjusted. This is to reduce the interaction between the heating and cooling zones by creating a buffer zone.

In the heating zone, the TCs respond first to the change in BC, then the TBs react to the radiation energy in the crown. In the cooling zone, the TCs respond first to the change in CA, then the TBs react to the TCs in the cooling zone and TBs in the heating zone simultaneously.

III. ANALYSIS OF FURNACE PROCESS

A. Selection of Input-Output Variables

Since the furnace melter has many redundant sensors which have similar characteristics, only the major sensors are selected. The TC2 and TC6 are selected as representative sensors for TCs and TB2 is selected as for TBs. The CAs of port 4 and 5 are also excluded from the input or manipulated variable because they supply the constant amount of CA. Thus, the temperature control problem is formulated for the system with 2 inputs and 3 outputs:

Outputs: y={TC2, TB2, TC6}
Inputs: u={total BC, Port 6 CA}

Total BC is distributed with constant ratios as

$$\begin{aligned} BC_1(k)&=r_1 BC(k), \\ BC_2(k)&=r_2 BC(k), \\ BC_3(k)&=r_3 BC(k), \end{aligned} \tag{1}$$

where k is the discrete time step, BC, BC_1, BC_2 and BC_3 are the amount of total oil, port 1 oil, port 2 oil and port 3 oil, respectively, and r_1, r_2 and r_3 are the distribution factors for port 1, port 2 and port 3, respectively, such that $r_1 + r_2 + r_3 = 1$. The distribution factors r_1, r_2 and r_3 are determined from experience, and do not change much during the life of the furnace.

B. Input-Output Modeling of the Furnace

Since the mathematical model for the furnace is not available, identification is performed based on the experimental process data. One of the most common identification methods is to minimize a performance index which represents the error between the real plant output and the model output [11],[12]. In this paper, several process tests are done carefully to generate the perturbed input-output data from the furnace in operation. The model structure is assumed as a FOPDT system [1]-[4], then the parameters of the model are found with the root mean squares method [11],[12].

At first, the reaction of TC2 from the total fuel, BC, can be found from the reversing characteristics of the glass furnace. Fig. 3 shows the response of TC2 from BC. In Fig. 3, the horizontal axis is time and the unit is minute, therefore 200-minute data is presented. The vertical axis is temperature [°C] for TC2 and the liter per hour [*l*/hour] for BC. Since the TC2 and BC data show a large difference in scale, they are scaled with proper constants to plot in one figure. The valleys at every 20th minute represent the reverse of the left-right firing. There is a 6~9 minutes interval from the time that the flames of the left (right) regenerator are distinguished to the time that the flames of the right (left) regenerator supply the heat.

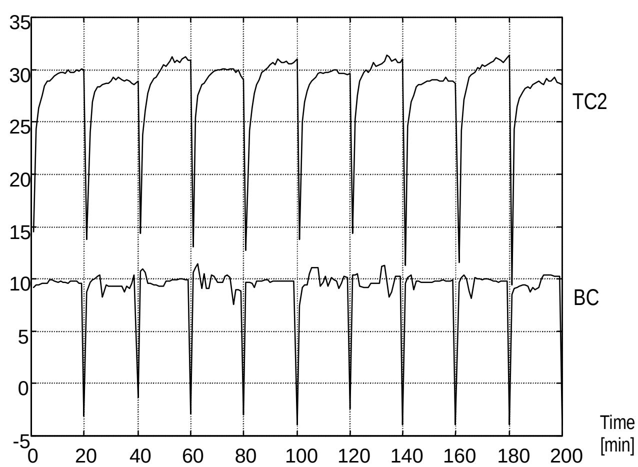

Fig. 3. The response of TC2 from BC.

Fig. 3 shows that the TC2 can be modeled adequately with the linear first-order system with some dead time, that is, FOPDT system [1]-[4]:

$$G_1(s) = \frac{TC2(s)}{BC(s)} = \frac{K_1 e^{-T_1 s}}{\tau_1 s + 1} \tag{2}$$

where

K_1 : steady state gain of TC2 from BC,
T_1 : dead time of TC2 from BC,
τ_1 : time constant of TC2 from BC.

Then, the modeling problem is to determine the model parameters, K_1, T_1 and τ_1. When there is no noise or disturbance in the TC2 data, the identification problem is to adjust the model parameters to match the model output data to the TC2 data. However, the existence of noise and disturbance makes the identification problem difficult. In the furnace, parameters are estimated quite differently with the choice of input-output data due to the noise and disturbance. Therefore, the process input-output data for identification should be chosen carefully to avoid the effect of noise and disturbance.

One of the reactions of TB2 and TC6 from step change of BC is shown in Fig. 4. As in Fig. 3, TB2, TC6 and BC data are scaled with proper constants to plot in one figure. The BC is decreased for 4 hours to perturb the furnace in this test. Compared with TC2, the TB2 and TC6 do not respond clearly because of unknown disturbances in the test. However, it can be found that the TB2 and TC6 start to react after about 1 hour of the BC change, and the time constant is about 2~3 hours. This test with BC was done iteratively. With the assumption that the transfer functions of BC-TB2 and BC-TC6 are represented with the FOPDT systems, the model parameters are found by the root mean squares method.

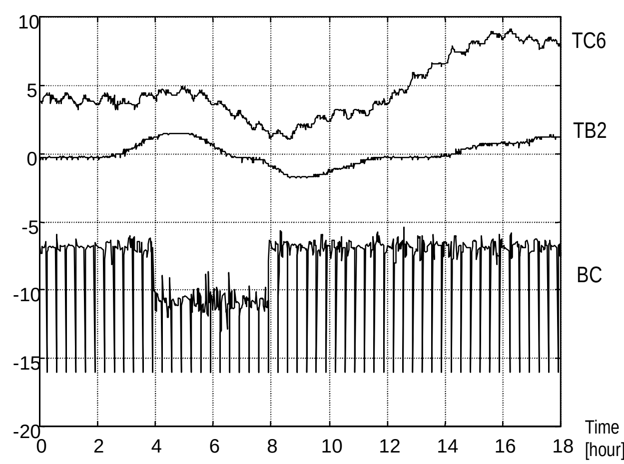

Fig. 4. The process test data with BC.

To identify the effect of CA, several process tests are also done with CA. Fig. 5 shows one of the test data. The reactions of TC2, TB2 and TC6 from step changes in CA are shown for 20 hours in Fig. 5. As before, TC2, TB2, TC6 and CA data are scaled with proper constants to plot in one figure, where the vertical axis is temperature [°C] for TC2, TB2 and TC6, and volume per hour [m^3/hour] for CA. The CA is increased for 4 hours and decreased for 4 hours to perturb the furnace in this test. Compared to TC6, the TC2 and TB2 do not respond clearly to the CA. The TC6 starts to respond after about 5 minutes of the CA change, and the time constant is about 1 hour. The transfer function of CA-TC6 is also represented with the FOPDT system, and the model parameters are found with the root mean squares method.

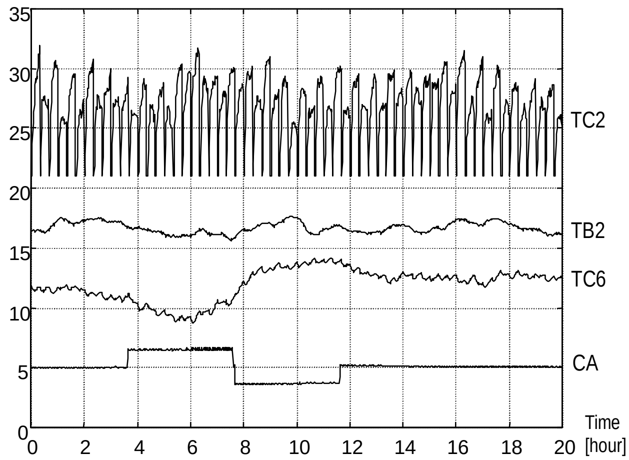

Fig. 5. The process test data with CA.

The bottom temperature TB2 is heated by the radiation energy of the melter, TC2, according to the characteristics of the furnace. Therefore, the reaction is the sequence of BC→TC2→TB2. To describe this sequence, the transfer function for TC2→TB2 needs to be found. The transfer function for TC2→TB2 can be induced with the transfer functions BC →TC2 and BC→TB2, which are found from the process test data.

Let G_a be the transfer function for BC →TB2:

$$G_a(s) = \frac{TB2(s)}{BC(s)} = \frac{K_a e^{-T_a s}}{\tau_a s + 1} \tag{3}$$

With (2) and (3), the transfer function for TC2 →TB2, G_2, is calculated as follows:

$$G_2(s) = \frac{TB2(s)}{TC2(s)} = \frac{G_a}{G_1} = \left(\frac{K_a}{K_1}\right)\left(\frac{\tau_1 s + 1}{\tau_a s + 1}\right) e^{-(T_a - T_1)s} \tag{4}$$

$$= \left(\frac{K_a}{K_1}\right)\left(\frac{\tau_1}{\tau_a} + \frac{1 - \left(\tau_1 / \tau_a\right)}{\tau_a s + 1}\right) e^{-(T_a - T_1)s}. \tag{5}$$

And, considering that τ_a is lager than τ_1, the τ_1/τ_a is neglected, then G_2 becomes

$$G_2(s) = \left(\frac{K_a}{K_1}\right)\frac{e^{-(T_a - T_1)s}}{\tau_a s + 1} \tag{6}$$

$$\triangleq K_2 \frac{e^{-T_2 s}}{\tau_2 s + 1}. \tag{7}$$

Therefore, G_2 can be roughly represented as a FOPDT system. The transfer functions for CA→TC6 and BC→TC6 are represented with G_3 and G_4, respectively, as follows:

$$G_3(s) = \frac{TC6(s)}{CA(s)} = \frac{K_3 e^{-T_3 s}}{\tau_3 s + 1} \tag{8}$$

$$G_4(s) = \frac{TC6(s)}{BC(s)} = \frac{K_4 e^{-T_4 s}}{\tau_4 s + 1} \tag{9}$$

As a result, the furnace is modeled with 4 FOPDT sytems as (2), (7), (8) and (9). And G_1 and G_3 have better quality than G_2 and G_4 in the modeling procedure.

IV. CONTROLLER CONFIGURATION

The furnace was modeled as a Multi-Input Multi-Output (MIMO) system with 2 inputs and 3 outputs. Compared with the modern control theory, which uses detailed mathematical model, the multi-loop control uses a combination of single loops of inputs and outputs. The proper pairing of controlled and manipulated variables and interactions among control loops in a MIMO system has been the subject of much research over the last 20 years [1]-[4]. Various types of decouplers were explored to separate the loops. Rosenbrock presented the Inverse Nyquist Array (INA) to quantify the amount of interaction [13]. Bristol developed the Relative Gain Array (RGA) as an index of the loop interaction [14]. These kinds of works are based on the premise that interaction is undesirable. Therefore, inputs and outputs, which have dominant relationships, are recommended to be paired.

This concept is also applied in this paper. In the case of the glass furnace, the paring is relatively simple. The furnace is divided into heating and cooling zones. The BC has a dominant effect on the heating zone, and the CA has a dominant effect on the cooling zone. Therefore, the basic loops are BC-heating zone and CA-cooling zone, and the interaction is considered on this basic configuration.

A. Control of the Heating Zone

When the heating zone is considered as a sub-system, the input of the sub-system is BC and the outputs are TC2 and TB2. This is a system where G_1 and G_2 are connected in cascade. It is well known that the cascade control scheme is a very useful technique when the inner or secondary loop is faster than the outer or primary loop [1]. In the case of the heating zone, the time constant of G_1 is smaller than that of G_2. Therefore, the cascade control is applied to the heating zone.

The inner loop, H_1 is the controller that tracks the set point of TC2 with BC. It is well known that when the dead time is smaller than the time constant, the FOPDT system can be controlled effectively by a PI controller [1], [15]. Therefore, a PI controller is applied as H_1 in this paper. The controller H_1 is described as follows:

$$H_1(s) = K_{p1}(1 + \frac{1}{\tau_{i1} s}) \tag{10}$$

where, K_{p1} is the proportional gain and τ_{i1} is the reset time of H_1.

The proportional gain and the reset time are determined by minimizing the Integral of the Time-Weighted Absolute Value of the Error (ITAE) performance index:

$$ITAE = \int_0^\infty t|e(t)|dt \quad (11)$$

where $e(t)=y(t)-y_{setpoint}$. Calculation shows the following results for G_1 [1],[15],

$$K_{p1} = \frac{0.859}{K_1}\left(\frac{T_1}{\tau_1}\right)^{-0.977}, \quad (12)$$

$$\tau_{i1} = \frac{\tau_1}{0.674}\left(\frac{T_1}{\tau_1}\right)^{0.686}, \quad (13)$$

where K_1, T_1 and τ_1 are the gain, dead time and time constant of G_1, respectively.

The outer loop, H_2, is to provide the setpoint of TC2, and has the structure of the PI controller as

$$H_2(s) = K_{p2}(1+\frac{1}{\tau_{i2}s}) \quad (14)$$

where K_{p2} is the proportional gain and τ_{i2} is the reset time of H_2.

The initial values of K_{p2} and τ_{i2} are determined to minimize the ITAE performance index (11) while using G_1, G_2 and H_1. Since G_2 does not have good quality, parameters K_{p2} and τ_{I2} need to be tuned again in practical implementation.

B. Control of the Cooling Zone

When the cooling zone is considered as a sub-system, the input of the sub-system is CA and BC, and TC6 is the output. The controller H_3 to control TC6 with CA is designed as a PI controller as

$$H_3(s) = K_{p3}(1+\frac{1}{\tau_{i3}s}) \quad (15)$$

where K_{p3} is the proportional gain and τ_{i3} is the reset time of H_3. The parameter values of K_{p3} and τ_{i3} are also determined to minimize the ITAE performance index while using G_3.

To avoid the interaction between heating and cooling zones, a decoupling loop needs to be applied in the furnace. General decoupling method is known for the 2 input - 2 output system [1] -[4]. The essence of decoupling is the imposition of a computing network which will cancel the interaction in the process, permitting independent single loop controls. However, the decoupling system need be an inverse of the process to cancel, and the inverse is not realizable because it requires leads and time advances to compensate lags and dead time [3].

In this paper, H_4 is applied based on G_4 to reduce the interaction between heating and cooling zones. The controller H_4 is also designed as a PI controller,

$$H_4(s) = K_{p4}(1+\frac{1}{\tau_{i4}s}) \quad (16)$$

where K_{p4} is the proportional gain and τ_{i4} is the reset time of H_4. The initial values of K_{p4} and τ_{i4} are determined to minimize the ITAE performance index (11) while using G_3, G_4, and H_4. Then, the parameters are tuned in the practical implementation.

Fig. 6 shows the overall configuration of the proposed control system. A limiter of y_1^{ref} is included in this figure. The role of this limiter is to avoid the TC2 being too far from the normal operation range.

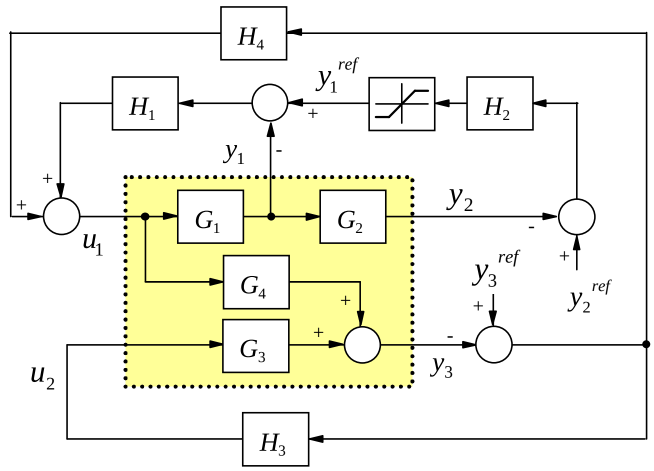

Fig. 6. The overall configuration of proposed control system.

V. IMPLEMENTATION RESULTS IN A REAL FURNACE

The proposed controller has been applied to a 150 ton/24 hour glass melting furnace in Samsung-Corning Company in Suwon, Korea. The Rosemount Distributed Control System (DCS) RS3 is installed in the furnace. The PI controllers, H_1 ~ H_4, are implemented in the internal module of the DCS.

Fig. 7 shows typical temperature variations of TC2, TB2 and TC6 in the manual operation. In Fig. 7, the horizontal axis is time and the unit is day, therefore 3-day data are presented. The vertical axis is temperature [°C], and TC2, TB2 and TC6 data are scaled with proper constants to plot in one figure. Fig. 8 shows typical temperature variations of TC2, TB2 and TC6 when the proposed control system is implemented for 3 days.

The variations of proposed control show relatively stable operation of the furnace. The standard deviations of TC2, TB2 and TC6 in the manual operation are 3.08 [°C], 1.00 [°C] and 0.95 [°C], respectively. And, the standard deviations of TC2, TB2 and TC6 in the automatic operation are 2.79 [°C], 0.63 [°C] and 0.76 [°C], respectively. The standard deviations of TC2, TB2 and TC6 are reduced to 90%, 63% and 80% in the automatic operation, resulting in a much smoother and tighter operation.

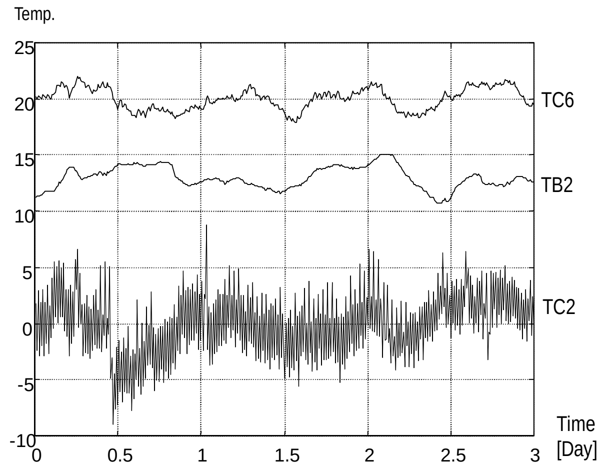

Fig. 7. Typical manual operation for 3 days.

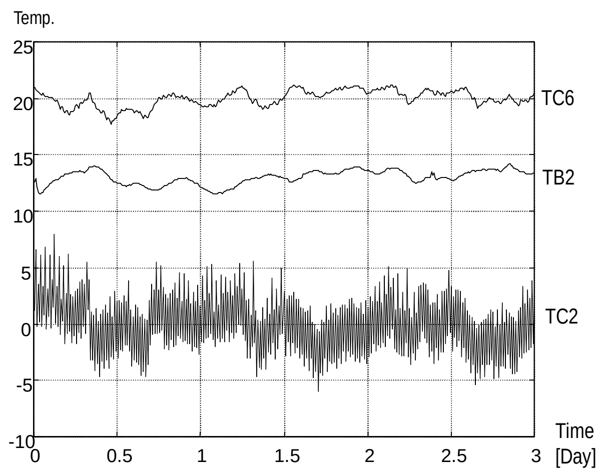

Fig. 8. Typical automatic operation for 3 days.

VI. CONCLUSIONS

A practical application of the conventional multi-loop control is proposed in this paper to the temperature control of a TV glass furnace. The major input and output variables are selected for the furnace, and the dynamics for dominant input-output pairs are modeled as the First-Order-Plus-Dead-Time (FOPDT) systems using the process experimental data. Based on the furnace model, a conventional multi-loop control, which is a combination of cascade and single loops is used to control the major input-output variables. Each single loop controller is designed with a PI controller, and heating zone is controlled with a cascade of controllers. A control loop is also added to reduce the interaction between the heating and cooling zones. The proposed control configuration has been successfully implemented in a real manufacturing furnace in Samsung-Corning Company in Suwon, Korea.

ACKNOWLEDGMENT

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REFERENCES

[1] C. A. Smith and A. B. Corripio, *Principles and Practice of Automatic Process Control*, John Wiley & Sons, 1985.

[2] William L. Luyden, *Process Modeling, Simulation and Control for Chemical Engineers*, McGraw-Hill, 1990.

[3] F. G. Shinskey, *Process Control Systems*, McGraw-Hill, 1988.

[4] Donald R. Coughanower, *Process Systems Analysis and Control*, McGraw-Hill, 1991.

[5] Dale E. Seborg, Thomas F. Edgar and Duncan A. Mellichamp, *Process Dynamics and Control*, John Willy & Sons, 1989.

[6] R. Stenz and U. Kuhn, "Automation of a batch distillation column using fuzzy and conventional control", *IEEE Trans. on Control Systems Technology*, vol. 3, no. 2, pp. 171-176, June 1995.

[7] R. Haber, J. Hetthessy, L. Keviczky, I. Vajk, A. Feher, N. Czeiner, Z. Csaazer, and A. Turi, "Identification and adaptive control of a glass furnace", *Automatica*, vol. 17, pp. 175-185, 1981.

[8] S. Aoki, S. Kawachi and M. Sugeno, "Application of fuzzy control logic for dead time process in a glass melting furnace", *Fuzzy Sets and Systems* 38, pp. 251-265, 1990.

[9] M. Hadjili, A. Lendasse, V. Wertz and S. Yurkovich, "Identification of fuzzy models for a glass furnace process", *Proc. of the 1998 IEEE International Conference on Control Applications*, Trieste, Italy, pp. 963-968, September 1998.

[10] U.-C. Moon and K. Y. Lee, "Temperature Control of glass melting furnace with fuzzy logic and conventional PI control", *Proc. the 2000 American Control Conference*, Chicago, IL, pp. 2270-2724, June 2000.

[11] T. Soderstrom and P. Stoica, *System Identification*, Prentice Hall, 1989.

[12] R. Johansson, *System Modeling and Identification*, Prentice Hall, 1993.

[13] Rosenbrock, *Computer-Aided Control System Design*, Academic Press, 1974.

[14] E. H. Bristol, "On a new measure of interaction for multivariable process control", *IEEE Trans. on Automatic Control*, vol. 2, No. 1 pp.133-134, 1966.

[15] A. M. Lopez, P. W. Murrill, and C. L. Smith, "Controller tuning relationships based on integral performance criteria", *Instrument Technology*, vol. 14, no. 11, p. 57, Nov. 1967.