DYNAMICS AND CONTROL OF CHEMICAL REACTORS, DISTILLATION COLUMNS AND BATCH PROCESSES

Edited by
J. E. RIJNSDORP, J. F. MacGREGOR,
B. D. TYREUS and T. TAKAMATSU

IFAC Symposia Series, 1990. Number 7

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Selected Papers from the IFAC Symposium,
Maastricht, The Netherlands, 21–23 August 1989

Edited by

J. E. RIJNSDORP
University of Twente, Enschede, The Netherlands

J. F. MacGREGOR
McMaster University, Hamilton, Canada

B. D. TYREUS
E.I. du Pont de Nemours & Co, Newark, USA

and

T. TAKAMATSU
Kansai University, Osaka, Japan

Published for the

INTERNATIONAL FEDERATION OF AUTOMATIC CONTROL

by

PERGAMON PRESS
Member of Maxwell Macmillan Pergamon Publishing Corporation
OXFORD · NEW YORK · BEIJING · FRANKFURT
SÃO PAULO · SYDNEY · TOKYO · TORONTO

U.K.	Pergamon Press plc, Headington Hill Hall, Oxford OX3 0BW, England
U.S.A.	Pergamon Press, Inc., Maxwell House, Fairview Park, Elmsford, New York 10523, U.S.A.
PEOPLE'S REPUBLIC OF CHINA	Pergamon Press, Room 4037, Qianmen Hotel, Beijing, People's Republic of China
FEDERAL REPUBLIC OF GERMANY	Pergamon Press GmbH, Hammerweg 6, D-6242 Kronberg, Federal Republic of Germany
BRAZIL	Pergamon Editora Ltda, Rua Eça de Queiros, 346, CEP 04011, Paraiso, São Paulo, Brazil
AUSTRALIA	Pergamon Press Australia Pty Ltd., P.O. Box 544, Potts Point, N.S.W. 2011, Australia
JAPAN	Pergamon Press, 5th Floor, Matsuoka Central Building, 1-7-1 Nishishinjuku, Shinjuku-ku, Tokyo 160, Japan
CANADA	Pergamon Press Canada Ltd., Suite No. 271, 253 College Street, Toronto, Ontario, Canada M5T 1R5

First edition 1990

Library of Congress Cataloging in Publication Data

IFAC Symposium on Dynamics and Control of Chemical Reactors, Distillation Columns, and Batch Processes (DYCORD+ '89) (1989: Maastricht, Netherlands)
Dynamics and control of chemical reactors, distillation columns, and batch processes (DYCORD+ '89): selected papers from the IFAC Symposium, Maastricht, the Netherlands, 21-23 August 1989/edited by J. E. Rijnsdorp . . . [et al.].—1st ed.
p. cm.—(IFAC symposia series: 1990, no. 7)
Sponsored by the International Federation of Automatic Control, Technical Committee on Applications, Working Group on Chemical Process Control and co-sponsored by European Federation of Chemical Engineering (404th event).
1. Chemical reactors—Data processing—Congresses. 2. Chemical process control—Data processing—Congresses. I. Rijnsdorp, John E. II. International Federation of Automatic Control. Working Group on Chemical Process Control. III. European Federation of Chemical Engineering. IV. Title. V. Series
TP157.I36 1989
660'.283—dc20 90–32048

British Library Cataloguing in Publication Data

Dynamics and control of chemical reactors, distillation columns and batch processes (DYCORD+ '89).
1. Chemical engineering equipment: Reactors
2. Distillation
I. Rijnsdorp, J. E. (John E.), *1928-* II. International Federation of Automatic Control III. Series
660.283

ISBN 0–08–037038–1

These proceedings were reproduced by means of the photo-offset process using the manuscripts supplied by the authors of the different papers. The manuscripts have been typed using different typewriters and typefaces. The lay-out, figures and tables of some papers did not agree completely with the standard requirements: consequently the reproduction does not display complete uniformity. To ensure rapid publication this discrepancy could not be changed: nor could the English be checked completely. Therefore, the readers are asked to excuse any deficiencies of this publication which may be due to the above mentioned reasons.

The Editors

Printed in Great Britain by BPCC Wheatons Ltd, Exeter

IFAC

International Federation of Automatic Control

DYNAMICS AND CONTROL OF CHEMICAL REACTORS, DISTILLATION COLUMNS AND BATCH PROCESSES

IFAC Symposia Series, 1990. Number 7

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Sponsored by
International Federation of Automatic Control (IFAC)
Technical Committee on Applications
Working Group on Chemical Process Control

Co-Sponsor
European Federation of Chemical Engineering (404th Event)

Organized by
Royal Institution of Engineers in the Netherlands
Division for Automatic Control (Dutch NMO of IFAC)

Further supported by
DSM, Geleen
EXXON Chemical Holland, Rotterdam
GE Plastics Europe, Bergen op Zoom

INTRODUCTION

Abstract. This volume contains the proceedings of the IFAC/EFCE-Symposium "DYCORD+'89", which was held at Maastricht (Netherlands) from August 21 to 23, 1989. The previous symposium (Bournemouth 1986) was focussed on Dynamics and Control of (chemical) Reactors and Distillation columns, hence its acronym DYCORD. For 1989 the scope has been extended to batch processes, which is indicated by the plus sign in the name. In this way, the present symposium deals with three application areas of great interest to automatic process control.

Keywords: dynamics and control of: batch processes, chemical reactors, fermentation reactors, distillation processes; process control; process modelling; AI techniques.

REACTORS
Chemical reactors usually form the critical core of chemical plants. In most cases, what is done there cannot be undone, by rerunning nor recycling. In particular catalysts have to be protected against contaminants, high temperatures, etc. A problem is the great variety of reactor types and reaction mechanisms, hence solutions tend to be tailor-made.

DISTILLATION
Distillation is still the predominant separation process in the chemical industry. This mainly holds for continuously operated tray columns in bulk sectors, but tray or packed columns are also rather common in small-scale process industries, either in batch or semi-continuous operation. Due to their multistage or distributed character, it is quite a challenge to develop simple and realistic dynamic models. This particularly holds for columns with high purity products, which tend to behave in a very nonlinear manner.

Control systems for distillation columns are inherently multivariable. Of special interest is coping with the interaction between top and bottom purity control: Effective dual quality control systems bring about a reduction in plant investment costs, as now two pure products can be realized with one column. Additional interactions appear when two or more columns are heat-integrated. Evidently, advanced process control is a must for advanced process design.

BATCH PROCESSES
In the past, batch processes have often been treated as the Cinderella of the process industries. Continuous processes were considered to be superior, not in the least owing to the well-developed regulatory control and automation technology. But, with the introduction of the PLC (programmable logic controller) in the seventies, automatic sequence control has become a practical possibility for flexible and reproducible operation of batch processes. In the eighties, the second generations of PLC's and DSC's (distributed control systems) have enabled integration of sequence and regulatory control functions. Presently, system manufacturers put great efforts into extending and improving software for configuring sequence control functions, which holds promise for more efficient development and maintenance of batch automation systems.

Evidently, there is less and less reason for preferring continuous processes over batch ones. With the growing need for high-performance and specialty products, with large added value per kilogram, batch processing will become even more important.

For automation and control experts, batch processes offer interesting and diverse tasks. Compared to continuous ones, there are additional subjects, such as plant scheduling (the timing of process steps), program control (optimum time functions for process variables), adaptable recipes (coping with changing conditions at the beginning or during the batch), and flexible switching from one recipe to another.

METHODS AND TECHNIQUES
Thus far, we have mainly looked at the object for application of dynamic analysis and control systems design. Of course, during the symposium methods and techniques will receive much attention too. Complex processes need dynamic models which can easily be adapted and updated. These models also play an important role in various types of model-based control, which represent an important field of advanced control techniques. Additional possibilities are created by modern developments in artificial intelligence, in particular with respect to expert systems and neural networks.

SYMPOSIUM SCOPE
In soliciting and selecting papers for the symposium it was not always easy to define the borderlines of the various topics. Strictly speaking, fermentation processes are biochemical rather than chemical reactors, but we felt that including them was worth while in view of their growing importance, and their interesting modelling and control aspects. Three related papers about modelling crystallizateion created some discussion. Finally we decided to allow reduction to one paper, focussed on topics of central interest to the symposium. Adsorption and absorption operations differ from distillation processes, but control requirements and model structures are quite similar. Batch processes come in very broad variety, hence we had to resort to the term "miscellaneous" in defining paper clusters for sessions.

All in all, we hope the selected papers will also be of interest to process designers in order to catch up with opportunities and limitations of modern process control, and to control scientists to obtain realistic cases for research in methods and techniques.

Prof. John E. Rijnsdorp (chairperson IPC)

Lupinestraat 32, 7552 HK Hengelo (O)
Netherlands

SUMMARY

Compared to the first DYCORD Symposium (held in Bournemouth, UK, 1986), the scope of DYCORD+'89 included dynamics and control of batch processes as an additional third topic. There were 189 participants from 19 countries, of which 123 work in industry.

For each of the three main topics there was a plenary survey paper, all emphasizing the broad diversity of processes, operating conditions, and problem solutions. A fourth plenary presentation was devoted to the interactions between process design and process control. It highlighted singular value analysis as a tool for assessing plant operability, with consequences for the process structure.

A plenary panel session focussed on "sense and nonsense of artificial intelligence for process control". The number of industrial applications still appears to be very small.
The table gives a survey of papers announces, submitted, and actually presented in the technical sessions.

TOPIC	SECOND ANNOUNCEMENT	SUBMITTED	PRESENTED
distillation	19	18	15
misc. cont. processes	6	6	6
cont. chemical reactors	12	11	8
batch chemical reactors	7	5	4
fermentation reactors	3	3	3
other batch processes	8	7	7
Totals	55	50	43
wholly or partly from industry	20	18	15

There were a good number of papers about industrial applications of advanced control techniques. Solutions were offered for control of distillation columns with high purity products. A lasting problem is the choice of control structures.

Of special interest were papers about optimal thermal processing of canned food, application of finite automata theory to sequential control, and dynamics of crystallizers.

Two technical sessions were devoted to the application of artificial intelligence techniques to process control. Two of the papers were about neural networks, introduced by a nice tutorial in a well-visited early morning session.

In general model-based techniques are making headway in industry, often combined with parameter estimation. A major problem is "coping with diversity" which requires adaption of general methods based on process insight. The symposium scope appears to attract many papers and participants, with much interest from industry. As a consequence, a third symposium in the series is being planned for 1992.

prof. J.F. MacGregor,
prof. J.E. Rijnsdorp,
prof. T. Takamatsu,
dr. B.D. Tyreus.

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THE IMPACT OF PROCESS DIVERSITY ON DISTILLATION COLUMN CONTROL

W. L. Luyben

Process Modeling and Control Center, Department of Chemical Engineering, Lehigh University, 111 Research Drive, Bethlehem, PA 18015, USA

ABSTRACT

The engineering science of distillation column design and control is considered by many academic visionaries to be a mature field that is not worthy of continued research. However, the economic importance of distillation in the chemical and petroleum industries continues to be very significant. This is true in both old and new plants and processes. The replacement of "inefficient" distillation by newer separation methods that has been predicted for over a decade has not occurred. The hard-nosed economics continue to show well-designed distillation systems can out-perform any of the alternatives in a very large number of processes.

Many universities around the world with more practical engineering research interests continue to have active research programs in distillation. This is particularly true in the area of the dynamics and control of distillation systems. The increasing diversity and complexity of distillation systems present significant design and control challenges for future engineers.

This paper attempts to point out how the extensive diversity that is encountered in distillation applications has a significant impact on the design of control systems for distillation.

INTRODUCTION

An invitation to give a plenary lecture is always greeted with mixed emotions. On the one hand, it is supposed to be somewhat of an honor to be asked, since at least the meeting organizers feel you might have something worthwhile to say. On the other hand, I remember the remark that Professor Neal Amundson made during one of his many invited lectures. He said that when you start receiving this kind of invitation it means you are rapidly approaching senility. Therefore, I don't know if I should be flattered or concerned. In any event I will do my best to convey some of my personal thoughts on why the field of distillation dynamics and control is where it is and where it may be going in the future.

Distillation is dead! So say the academic visionaries. These learned minds claim that if you are not into biotechnology, material science, polymers or electronic processing, you might as well hang it up. There is no future in anything else. Traditional unit operations and engineering are given nothing but bad press. Some of our more enlightened professors have even decided that distillation has no place in a modern chemical engineering undergraduate education, and certainly not in graduate education. The courses should cover "modern" chemical science, with increasing emphasis on computer science and artificial intelligence (as opposed, I supposed, to "real" intelligence!). The students will be computer software "junkies" and hackers who know object-oriented programming but don't know the difference between an air-to-open and an air-to-close control valve and have never heard of relative volatility.

Well, I don't believe a word of it! Distillation is alive and growing. I have spent all my professional life working in distillation, some thirty-five years, and every week brings a new situation, a new problem, a new insight, a new theory to test, a new design to evaluate or a new approach to an old problem. My prediction is that the importance of distillation in the real world of chemical engineering will not soon diminish. I am also confident that more level heads will prevail in most of our educational institutions, and the importance of distillation as a mainstay of chemical engineering education will continue into the next century.

Many plenary lectures basically provide an extensive review of the literature of the field. That is not the goal of this paper. In fact this paper will set a record for having the fewest number of references. I apologize to the organizers if a literature survey is what they expected, but I personally find it more interesting to discuss a few topics that are perhaps more philosophical than technical. When an engineer reaches my age, he or she should be permitted such indulgences. I hope that the points I make will provide a little insight, provoke some healthy constructive controversy and help to explain why we are where we are in distillation control.

DISTILLATION PROCESS DIVERSITY

The most significant thing that strikes me about distillation is the wide diversity of types of columns, processes and plants in which they are extensively used. This phenomenon is difficult for the novice in the distillation field to appreciate, but it is an important lesson to be learned. The situation is a little like the man who says that there are two types of women. What you can conclude from his statement is that he has known two women. There are some columns that are quite similar in several plants, but these are certainly the exception and not the rule.

Certainly some distillation columns can be classified into general types that perform similar separations. Some examples are given below, but this list is by no means exhaustive.

A. STABILIZERS: These are columns in which only a small distillate is produced that is usually much lighter than the rest of the components in the feed. These columns typically have small flow rates of reflux and distillate. The separation is relatively easy (modest relative volatility) and a reasonable temperature change occurs in the rectifying section. The temperature profile in the stripping section is very flat. Most of the reboiler duty goes into providing sensible heat to the liquid feed, so only a small amount of the energy consumption is due to reflux. Bottoms is used to preheat the feed to reduce energy consumption in the reboiler.

Figure 1 shows a typical stabilizer column with its standard control system. Heat input holds a tray temperature a few trays from the top of the column where the temperature change reflects the position of the light-key/heavy-key profile. A temperature near the top is not used because this is where the lighter-than-light-key components build up and affect temperatures very strongly. Reflux is ratioed to feed flow rate. Distillate comes off on reflux-drum level control if it is a liquid product or on column pressure control if it is a vapor product.

B. SUPERFRACTIONATORS: These columns separate components (usually binary) that have very close boiling points (low relative volatility). They are characterized by a large number of trays, high reflux ratios and very slow dynamics. The temperature profile in the column is very flat and the temperature difference between the top and the bottom of the column is relatively small. Because of this small ΔT, vapor recompression can often be economically used in these columns. Also because of this small ΔT, temperature can seldom be used to infer composition, so direct composition analyzers are often required.

Figure 2 shows a typical superfractionator with a commonly used control system. Reflux-drum level is controlled by reflux flow because of the high reflux ratio. Distillate flow rate controls distillate composition. Reboiler steam controls bottoms composition. Base level is held by bottoms flow rate. Because the dynamics of the column are very slow (sometimes many hours) extensive use is made of feedforward control and the column is isolated from external disturbances as much as possible.

C. PETROLEUM FRACTIONATORS: These columns usually have no reboilers. Vapor is generated by flashing hot feed which comes from either a high-temperature reactor or from a fuel-fired furnace. Several sidestreams are withdrawn through live-steam strippers. Crude fractionators (pipe stills or benches) and catalytic cracking fractionators are the most important examples. Since high temperature levels are involved, recovery of energy is very important. Multiple intermediate condensers (pumparounds) are used.

A degrees-of-freedom argument can be used to show that it is possible to control only one "major" product quality in each product. For example, if we control the 95% point of the first sidestream, we can do very little to change the 5% point of the second sidestream. Stripping steam can be used to change the initial boiling point of the second sidestream somewhat, but its effect on the 5% point is small.

Changing a liquid sidestream flow rate affects the 95% point of that sidestream and affects all products below that sidestream, but has only very small effect on products above that sidestream. This "one-way" decoupling makes the control problem easier.

Changing a pumparound affects products both above and below the pumparound since less vapor is sent up the column and more liquid is sent down. However, pumparounds are frequently held almost constant for a given type of feed at flow rates that reflect the economic optimum trade-off between energy recovery and product recovery. Increasing pumparound heat removal decreases fuel consumption in a pipe-still furnace or maximizes high-pressure steam generation in a cat-cracking fractionator. However, increasing a pumparound flow reduces the vapor and liquid flow rates in the sections of the column above the pumparound. This reduces the "fractionation" between product streams above the pumparound (the difference between the 95% boiling point of the lighter product and the 5% boiling point of the heavier product), which affects the yields of the various products.

Figure 3 shows a small petroleum fractionator with three sidestreams and two pumparounds. It is assumed that both the 5% and 95% points of the second sidestream are important and must be controlled. Therefore, neither the 95% point of the first sidestream nor the 5% point of the third sidestream can be controlled.

D. HIGH "K" COLUMNS: The vapor-liquid equilibrium K-values or relative volatilities between the two key components are very large, making the separation very easy. I like to use the example of the separation of hydrogen and peanut butter! The temperature profile is very sharp and the temperature difference between the top and the bottom of the column is very large. If we try to control the temperature on a single tray in this type of column, we will find it is quite difficult because of the high process gain (a very small change in the manipulated variable, heat input, produces a very large change in the temperature) and because of nonlinearity (the temperature saturates at either the high or low values as the profile moves past the control tray). A commonly used strategy to overcome these problems is to use temperature sensors located at several trays to control an "average" temperature.

Figure 4 shows a typical column and control system. Energy consumption is low because the separation is so easy, so reflux flow is simply held constant.

E. STRIPPERS: Feed is introduced on the top tray. No reflux is used. The overhead vapor is the distillate product, and it is almost in equilibrium with the feed. Figure 5 shows typical stripper control system. The only manipulated variable is heat input to the reboiler. It is used to keep the desired amount of light component in the bottoms product. There is no control of the recovery of heavy component, so significant yield losses can occur (heavy component going overhead) unless the separation is quite easy or the amount of light component in the feed is low.

F. SIDESTREAM COLUMNS: Both vapor and liquid sidestreams are employed, as are single and multiple sidedrawoffs. Strippers and rectifiers are sometimes coupled with the sidestreams. Sidestream columns are used in binary separations to produce three or more products with different purities. Sidestream columns are used for ternary separations when the concentration of the lightest component in the feed is fairly low (less than 10%), using a liquid sidestream, or when the concentration of the heaviest component in the feed is fairly low, using a vapor sidestream. When these feed concentrations are not low, sidestream strippers or rectifiers are usually used for improved energy efficiency.

Figure 6 shows a liquid sidestream column with a stripper. Distillate composition is held by reflux, sidestream composition by heat input to the stripper reboiler and bottoms composition by heat input to the column reboiler.

The flow rate of liquid to the stripper is used to minimize energy consumption as feed composition changes (particularly the amount of the intermediate component in the feed). In the example shown, a temperature difference is used to achieve this goal.

SOURCES OF DIVERSITY

The list given above illustrates some of the generic columns that occur in distillation and their typical control systems. However, despite the similarities, the individual columns within each of these classifications can be very different in both their design and control for a number of reasons. Some examples of conditions that vary from column to column are given in this section.

A. FEED CONDITIONS: The number of components in the feed and their concentrations can be quite different from plant to plant. A feed that has a lot of light component requires a different column design and a different control strategy than one that has a lot of heavy component. The same is true when different amounts of non-key components are present in a multicomponent mixture. The lighter-than-light and heavier-than-heavy key components can result in drastic changes in the temperature profile and require different column, reboiler and condenser designs and different control systems.

Different feed thermal conditions can also affect column design and control. A column with a subcooled liquid feed will look and act differently than one with a superheated vapor feed. Figure 7 compares columns with liquid and vapor feeds. Subcooled-liquid feed requires a bigger reboiler and larger diameter stripping section. The control system must change heat input as feed flow rate changes but must prevent flooding of the smaller diameter rectifying section. Superheated-vapor feed requires a bigger condenser and larger diameter rectifying section. The control system must change reflux flow as feed rate changes but must prevent flooding of the smaller diameter stripping section. Weeping in the larger diameter rectifying section at low feed rates must also be prevented.

The feed preheating system can vary from column to column and can profoundly affect the dynamics of the column. Some feed preheaters act as a source of positive feedback in a temperature loop which can destabilize the column. This is particularly true for columns which use a vapor sidestream to preheat the feed. The positive feedback is very large because the heat is latent heat of condensation.

B. PRODUCT SPECIFICATIONS: Product specification: High purity requirements may require a different column design and different control structure than low purity specifications. High-purity columns are very nonlinear and more sensitive to disturbances. The high-purity process may have to be isolated from large and rapid changes in feed composition and feed flow rate (using larger feed tankage) than the low-purity process. The control of the high-purity column may require more temperature-composition cascade systems and/or nonlinear controllers, whereas the low-purity column may work well with simple linear tray temperature controllers.

If the product must be controlled within a very narrow band of purity (must be "on aim"), the control structure will also be quite different than if purites must only be held at or above specification. The specified band may be so tight that it is physically impossible to achieve such small deviations using normal control systems. It may be necessary to go to some kind of blending system. Two stream with purites above and below the specification are blended to provide tight control of composition. This is very inefficient from an energy consumption point of view (second law of thermodynamics), but it may be the price that has to be paid to accomplish the control objective. Figure 8 shows a column where the overhead is deliberately made more pure than the specification, and some feed is bypassed around the column and blended with the overhead to produce the product. Of course, instead of using feed, it may be more economical to use a sidestream for blending.

C. ENERGY COSTS: One of the major sources of variability is energy costs. These can vary by an order of magnitude from one plant to another. This is not just due to variations in the cost of raw energy sources such as oil, gas or coal. These base fuel costs vary from location to location around the world because of transporation costs and national tax structures. But the production and uses of energy in each individual plant can also lead to major differences in energy costs from plant to plant in any one country.

For example, if you are operating a propylene-propane distillation column (which has a low base temperature and therefore can use very low-pressure steam as its energy source) and if there is an exothermic reactor in your plant that can be cooled by generating low-pressure steam instead of using cooling water, the energy costs in the distillation column would be very low. On the other hand, if your C_3 splitter is located in a plant where 300 psig steam must to dropped down in pressure and used in its reboiler, the energy costs will be high.

The designs and the control systems used in these two columns would be different.
If energy is cheap, you would probably build a conventional C_3 splitter that has many trays (200), uses a high reflux ratio (14) and operates at 17 atm. so that cooling water can be used in the condenser. If energy is expensive (and no source for heat integration is available), you would probably build a vapor-recompression C_3 splitter that has fewer trays (150), uses a lower reflux ratio (11) and operates at 11 atm. The control systems used on the cheap-energy column might be one in which you control only one end of the column, wasting a little energy to simplify the control system, but avoiding dynamic interaction problems and gaining the added benefit of improved product recovery. See Figure 9A. The control system used on the expensive-energy column would probably try to minimize energy consumption by using a more complex control structure (dual composition control via multiloop SISO controllers shown in Figure 9B or multivariable controllers such as DMC).

D. UPSTREAM AND DOWNSTREAM UNITS: If your distillation column is fed from a large storage tank and if its products are sent into large tanks, the design and control of the column can be treated differently than if the column is in a direct sequence of operating units. Limitations in flow rate changes (both absolute magnitude and rate of change) in upstream or downstream units connected to the column can require a different control structure than would be used if the column were operated in isolation.

For example, suppose the distillate product from a column is a vapor. If this vapor is discharged into a large header which can accommodate the swings in vapor flow, the vapor distillate stream would probably be used to control column pressure. Let us assume that the rest of the control system uses reflux to control a tray temperature near the top of the column, steam to the reboiler is fixed and reflux drum level is held by cooling water flow rate. See Figure 10A. Any changes in the energy or material balances will show up directly in frequent and large swings in vapor distillate flow rate.

However, if the vapor distillate is fed into another distillation column, large changes in flow rate could upset this downstream column. Therefore the appropriate control system on the first column might be (see Figure 10B) to use reboiler steam to control temperature, hold reflux drum level with reflux flow rate, control column pressure with cooling water flow rate and ratio distillate vapor flow to reflux flow. The reflux drum level controller would be proportional-only and a large drum would be used so that the changes in reflux and distillate flow rates are very gradual.

E. SOURCES OF HEATING AND COOLING: The heating and cooling media used in the reboiler and condenser can drastically affect column design and control. If a direct-fired furnace or a Dowtherm heater is used to provide heat, the dynamics of the column/reboiler system will be much slower than if high-pressure steam is used, and a different control system may be required. If cooling water is used in the overhead condenser, column design and optimum operation will be different than if refrigeration is used in the overhead condenser. For example, if the condenser is water or air cooled, the feed should normally be preheated as much as possible to reduce energy consumption in the reboiler (note, total energy consumption is not reduced as feed preheat is increased). However, if the condenser is refrigerated, the feed should be kept as cool as possible to reduce refrigeration costs.

The use of heat integration among various processes and columns usually requires process designs and control system designs that are quite different than when energy sources and sinks can be treated as utilities. Floating pressure in the high-pressure column is sometimes used; but in other systems auxiliary reboilers and condensers must be used to break the coupling between the units. Figure 11 illustrates the use of an auxiliary condenser to hold pressure in the high-pressure column and an auxiliary reboiler in the low-pressure column to control bottoms composition. The pressure and composition controllers are both split-ranged so that auxiliary cooling and heating are only used when necessary. The opening of the valve in the vapor line to the reboiler/condenser can to reduced through the low selector by either the pressure controller (if the pressure is too low in the high-pressure column and after the condenser is completely flooded) or by the composition controller (if less heat is needed in the low-pressure column and after the steam valve to the auxiliary reboiler is completely shut).

Intermediate reboilers and intermediate condensers (pumparounds) are often used to permit the use of lower-temperature heat sources or to recover some of the energy that would otherwise be thrown away to cooling water in the overhead condenser. This intermediate heat addition and heat removal increase the total energy consumption (or increase the number of trays required to achieve a given separation), but they can reduce net energy costs. The column design and the control systems used on these column are much different than for a conventional column. As sketched in Figure 12, the control system tries to preferentially use low-level energy instead of high-level energy. The split-ranging of the bottoms composition controller may have to be done so that there is some overlap, i.e. the low-pressure steam valve is not completely open before the high-pressure steam valve opens. This may be necessary because the minimum cost operation may not correspond to the maximum use of low-pressure steam since some high-pressure steam is always needed because of vapor-liquid equilibrium limitations (the bubblepoint temperature of the bottoms product is higher than the saturation temperature of the low-pressure steam).

F. OPERATING LIMITATIONS: The column can be limited by a number of items. Let us take the simple example of flooding. The location of the flooding in the column and cause of the flooding can call for different actions. In many columns flooding can be reduced by *increasing* pressure. This increases vapor densities and reduces vapor velocities. However, in other columns flooding can be reduced by *decreasing* pressure. This occurs (a) if the flooding is caused by insufficient downcomer vapor-liquid disengaging capacity (which occurs in high-pressure columns where vapor and liquid densities are closer together), or (b) if the increase in pressure affects the relative volatility so much that the reflux ratio must to increased significantly, resulting in a net increase in the load on the column when pressure is increased.

If the column begins to flood in the rectifying section, the feed temperatures should be decreased to reduce vapor load in the top of the column. If the column floods in the stripping section, feed temperature should be maximized.

Other limitations include condenser heat removal (the usual one encountered in summer operation) and reboiler heat input (often encountered when the material has a tendency to foul the reboiler). Effective control systems must be able to handle all of these situations.

G. FEED MAKEUP SYSTEMS: A final example of a source of diversity is the variations in distillation column design and control caused by differences in where the feed to a plant is introduced and in its phase (vapor or liquid). The idea is best illustrated with an example. Suppose we have a plant which consists of a liquid-phase reactor and a column as shown in Figure 13. In the reactor components 1 and 2 react to form component 3, which is the product. There is an excess of component 1 to avoid undesirable side reactions, so we assume component 2 is completely consumed. The reactor effluent contains only components 1 and 3. These are separated in the column, with component 1 going overhead since we assume it is lighter than component 3. Distillate is recycled back to the reactor.

Reactant component 2 is flow controlled into the reactor. Clearly reactant component 1 also must be fed into the system somewhere to satisfy the stoichiometry of the reaction. How this makeup feed is introduced will affect both the design and the control of the column.

Let us assume that the basic control system is to flow control fresh component 2 into the system and to ratio the total flow (fresh plus recycle) of component 1 to the flow of component 2. Distillate composition is held by reflux. Bottoms composition is held by reboiler steam. Column pressure is held by condenser cooling water flow.

In this reactor/column system the inventory of component 1 is mostly in the reflux drum of the column. The level in this tank can be used to set the makeup of fresh component 1 into the system. If the makeup is a liquid, it might be added directly into the reflux drum on level control. This would lead to a simple and direct control structure as shown in Figure 13a.

However, suppose the source of fresh component 1 is not pure but contains some heavy impurity that we must keep out of the reactor due to catalyst contamination problems. Now it would make sense to add the liquid makeup stream into the column instead of directly into the reflux drum. As sketched in Figure 13b, this leads to an unusual control system since the liquid makeup stream has no direct effect on the level in the reflux drum. Distillate composition is now held by makeup feed, and reflux flow controls reflux drum level.

Note that if the makeup feed where introduced into the column as a vapor, the original control system could be used. The level in the reflux drum would be affected by the vapor makeup feed, assuming of course that the pressure controller is in automatic.

THE OVERRIDING IMPORTANCE OF PROCESS UNDERSTANDING

The result of all this diversity is that almost each column requires its own control system. This means that engineering time is significant. We do not have the situation that is common in the aerospace field where 20 engineer-years can be devoted to each control system, and then it is duplicated in hundreds of the same types of devices.

Therefore, the use of the simplest system that will do the job is highly desirable. And the understanding of how the particular column operates in its individual plant environment, its limitations and its constraints is vital to developing an effective control strategy. Hence, process understanding is much more important than understanding some elegant control theory. This is one of the reasons why most columns are controlled by simple SISO controllers with a sprinkling of override controllers added where necessary. Nonlinearity and constraints are more important than interaction and optimal performance.