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UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION

PROCESS INSTRUMENTATION
AND
CONTROL SYSTEM MODERNIZATION

PETROCHEMICAL COMPLEX IPCL
BARODA, INDIA

FINAL REPORT, OCTOBER 1986

SYSTEMATICS
SYSTEMS AND AUTOMATICS ENGINEERING
BULGARIA - SOFIA

UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION

PROCESS INSTRUMENTATION AND CONTROL

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PETROCHEMICAL COMPLEX IPCL

BARODA, INDIA

FINAL REPORT, OCTOBER 1986
FOR UNIDO PROJECT No DP/IND/84/001
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S Y S T E M A T I C S
SYSTEMS AND AUTOMATICS ENGINEERING
BULGARIA - SOFIA

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SECTION 1

1. INTRODUCTION

1.1. Subject of the report.

The subject of the report is to present some new control strategies, which could improve significantly the DMT Plant, Ethylene Glycol Plant, Acrylonitrile Plant, Linear Alcyl Benzene Plant and Industrial Utilities Plant operation in IPCL BARODA Complex, as well as their integrated option, on condition that suggested modern instrumentation and distributed control systems, available today, is used.

An eventual revamping should consist of a substitution of the actual instrumentation pneumatic system with a microprocessor based digital control system, which can assure better basic control, advanced control and optimization in order to increase the production plants efficiency and its operational stability.

This report is not a design document. Its objective is to evaluate the potential benefits, which advanced control and optimization using a distributed digital control system could bring to plant optimizations.

1.2. Purpose of the report

The main purpose of the present report is to advise and to assist the Indian Petrochemical Corporation Limited /IPCL/ -BARODA in development work involved in revamping existing pneumatic instrumentation with electronic equivalents with possibilities to addopt microprocessor based distributed control system.

It can be used for advise the opportunity for closed loop, advanced control and optimal control strategies for the plants specified , so as to achieve optimum plant performance.

Additionally can be used to advise and assist the IPCL engineering staff in the specification, installation and commissioning of new electronic instrumentation, distributed process control systems, integrated computer control system and suitable on-line analytical instruments as well as of the Industrial Utilities Management software system.

1.3.Description of the results.

The suggested final specification of the recommended new electronic field instruments and distributed process control systems can be used for preparation of enquaries for the purchase of a microprocessor based digital system not only on the base of TDC -3000 produced by HONEYWELL,but also on the base of available on the market digital system from other producers.

The list of recommended digital control system and its producers, as well as list of preferable factures of electronic field instruments, control valves and on-line analytical instruments is attached in the report.

However HONEYWELL is the world's leader in producing and applying distributed microprocessor based systems:The TDC-3000 has been conceived to be applied in all kind of petrochemical plants and it has been developed in accordance with the requirements of the major world's industrial companies.

The content of this report nas to be intended as a starting base for future deeper analysis of the control problems and the relevant possible solutions.

However, because of SYSTEMATICS and HONEYWELL wide experience on such kind of industrial units and plants, we can assure (like it is shown in the report) that significant savings can be achieved implementing higher level control and optimization strategies, which have been demonstrated in innumerable applications.

1.4. Summary of the report by items

Section 2 of this final report first explains the IPCL - Complex-Baroda like an object for possible implementation of new control systems based on microprocessor distributed control. The hierarchical approach - basic control, advanced control and optimization, for realization of enhancenments of control systems is described. Finally, description of the project area work and of the recommended control solution is made.

Section 3 suggests hierarchical approach for plants optimization, describes the technical support in the future engineering work, which can be given by SYSTEMATICS and explains the evaluation of economical and social benefits, as well as possible risks.

Section 4 contains the main system decisions for modernization of the instrumentation and the process control system for DMT, Acrylonitrile, Ethylene Glycol, LAB Plants and Industrial Utilities. The detail description of suggested advanced control schemes and optimal control strategies for each of the above plants is also presented. Finally, evaluation of possible economic benefits and savings is made and specifications for new process control systems based on distributed microprocessor are attached.

Section 5 contains the architecture, functional characteristics, hardware and software solutions for realization of an Integrated Computer Control System in IPCL Complex.

Section 6 is a summary of the general recommendations for implementation of distributed control system in DMT,EG,LAB,AN and Industrial Utilities Plants with recommended strategies of practical realization.Comparative evaluation of instrumentation and digital control system producers for petrochemical industries is also made.

The recommendations to the on-line analytical instruments and for on-line conversion implementation of a distributed digital control systems are included in this section.

ANNEX 1 contains the detailed suggestions for the new electronic instruments application in DMT Plant as well as the advanced and optimal control strategies with the technical specifications of the process control system for its realization.

ANNEX 2 to ANNEX 4 contain the suggested advanced control strategies and the technical specification of distributed digital system for ACN,EG and LAB Plants.

ANNEX 5 contains the technical specification for realization of Integrated Computer Control System and for integrated industrial utilities management software system.

APPENDIX A is an overview of the series ST-3000 (Smart Transmitters), produced by HONEYWELL and suggested for application.

APPENDIX B is an detailed overview of the functional and capacity characteristics of the different modules from the TDC-3000 distributed control system.

APPENDIX C describes the suggested Industrial Utilities Management Software Sytem.

SECTION 2

2. SUMMARY

2.1. Description of the IPCL Complex- Baroda, as an object of a research

The Indian Petrochemical Corporation Limited /IPCL/ was formed by the Government of India 1969 to operate the large Aromatics and Olefines Manufacturing units built at Baroda in Gujarat State. The complex has been expanded until now, it is composed of the thirteen major units,shown on Table 2.1.1. The product flow diagramme of the existing plants in the complex with the new projects /expansions/ is presented on Fig.2.1.1.

A number of these plants are now 7 years old. They are mostly equipped with pneumatic instruments and have no modern optimizing equipment. But they are required to work to capacity and major expansion/debottlenecking projects are planned or are in hand.

The current national plant - the Sixthe Five Year Plan - envisages rapid expansion of the petrochemical industry in India, using IPCL as the base. However, with increasing energy costs and rapidly changing nature of feedstocks and technology worldwide, the profile of this industry is undergoing a rapid change. Large plants with continuous production, precise operational control using sophisticated instruments , microprocessors, computers require a level of experience in operation and preventive maintenance as well as environmental control not conceived before. Extreme reliability in engineering a familiarity with computers and the systems approach to plant operation, an ability to anticipate failures and take corrective action by analysing the output of very precise measuring and alarm devices will all require a general upgrading of technological and operational perceptions as well as specific skill with respect to each of the above mentioned disciplines. The Sixth Plant proposes new petrochemical plants in India. These new plants will require manpower and IPCL will be the main source to cover the basic needs of trained manpower. A beginning has already been made in this area by providing a task forced for the proposed Maharashtra Gas Cacker and Salimpur Petrochemicals plants.

Keeping in view the aims of the above mentioned, Subcontract No 1 , the substantive terms of reference, dated 26 March 1985 and the strong requirements of IPCL management and engineering staff the objects of the research and preparation of the present report are as follows:

1. To examine and give detailed suggestions for a revamp of existing pneumatic instrumentation to electronic type with implementation of a distributed microprocessor based process control system for DMT Plant.

TABLE 2.1.1.

SR. NO	PLANTS	CAPACITY IN MTA	COLLABORIATION	DETAIL ENGG.	DATE OF COMMISSI-ONING
1.	AROMATICS P-XYLENE O-XYLENE DMT	 17 000 21 000 30 000 (to expanded to 40 000)	KRUP KOPPERS DYNAMIT NOBEL FRG	EIL	MARCH 1973
2.	OLEINS ETHYLENE PROPYLENE BUTHADIENE BENZENE	 130 000 71 000 22 000 23 600	LUMMUS UK	EIL	MARCH 1978
3.	LOW DENSITY POLYETHYLENE	80 000	AQUITANE TOTAL ORGANIC, FRANCE	EIL	MAY 1978
4.	POLYPROPYLENE	30 000	MONTEDISION+ TECHNIMOUT, ITALY	EIL	MAY 1978
5.	POLY BUTADIENE RUBBER	20 000	POLYESTER LTD CANADA	EIL	AUGUST 1978
6.	ETHYLENE GLYCOL	20 000	HALCAN INTER - NATIONAL,USA	EIL	JUNE 1978
7.	ALKYL BENZENE	30 000 (to be expected TO 43 500)	UNIVERSAL OIL PRODUCT, USA	EIL	AUGUST 1978
8.	ACRYLONITRILE	24 000 (to be expected TO 30 000)	PROSPECT INT+ BADGER BV, SOHIO,USA	EIL	DEC 1978
9.	ACRYLIC FIBRE	12 000	ASAHI CHEMICALS JAPAN	EIL	APRIL 1979
10.	ACRYLATES	10 000	ASAHI CHEMICALS, JAPAN	EIL	OCTOBER 1982
11.	PETROLEUM RESIN	5 000	CDF CHEMIC FRANCE	EIL	MARCH 1984
12.	VINYL CHLORIDE	57 000	STAUFFER,USA	EIL	APRIL 1985
13.	POLYVINYL CHLORIDE RESIN	55 000	GOODRICH,USA	EIL	APRIL 1985

SECTION 1

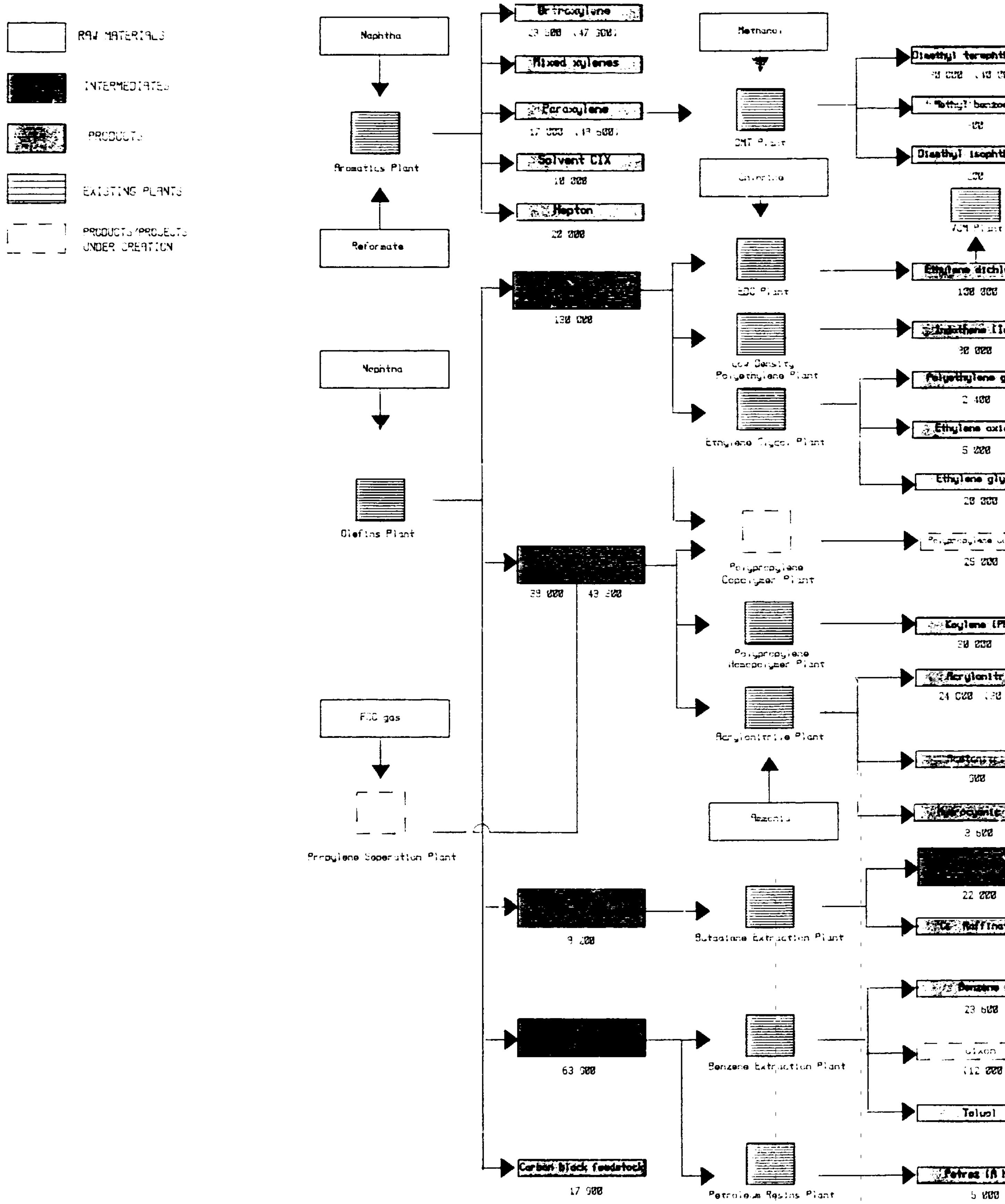


FIG. 2.1.1.

PRODUCT FLOW DIAGRAM
IPCL (VADODARA COMPLEX)

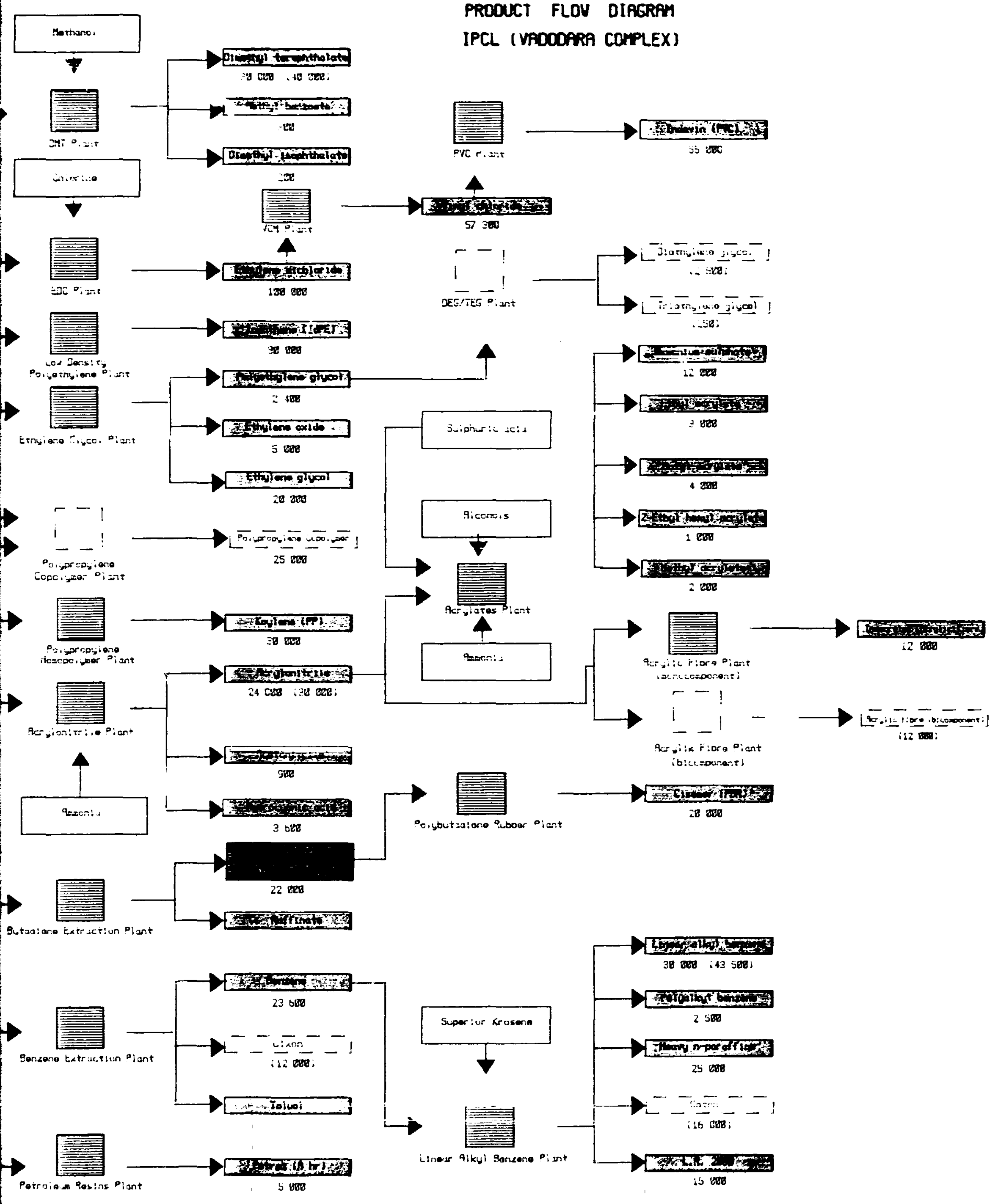


FIG. 2.1.1.

2. To study the present pneumatic instrumentation in the Linear Alcyl Benzene PLant, Ethylene Glycol Plant and Acrylonitrile Plant and suggest electronic microprocessor based distributed control systems implementation for this plants.

3. To study and suggest possibilities for realization of Integrated microprocessor based software system for IPCL boilers control and for the all industrial utilities.

4. To study the present on-line analyzers at IPCL Complex and to suggest better instruments.

5. To improve the capability of IPCL engineers to continue the development work on modernization of the process control systems by training on implementation of distributed digital control systems both on site /at IPCL Complex/ and abroad/ in Bulgaria/.

2.2.Basic definitions .

Before examing a real application it is preferabble to define some basic concepts.

In general, the objectives of an automation system are to guarantee the correct operation of a production process.

In sequence of priority the objectives are :

- to ensure the continuity of operation

The plant must run as long as possible to ensure maximum return of the construction investment in the shortest possible time.

-to guarantee the safety of plant operators and installed equipment

The operators have to be phisically protected from accidents which could occur in the plant. Therefore they must operate in a safe zone.

The most expensive equipment in the plant must be protected from operational conditions which could create damage that cannot be repared quickly.

- to maintain stable and/or repetative operating conditions

It is obvious that the more stable the operation of a continuous plant, or the higher the repetitiveness in the case of a discontinuous plant, the more stable will be the product quality.

- to maximize production

As a consequence of a higher stability reached in plant operation, the plant can operate closer to the limits of the theoretical specifications of the process

- to minimize the cost of production

This implies the selection of operational conditions which allow the determination of a production operation that maximizes the obtainable profit.

To satisfy the above-mentioned operational objectives, an automation system can use a series of independent but reciprocally related subsystems.

The system approach is to structure the control strategies in independent hierarchical modules(fig.2.2.1.)

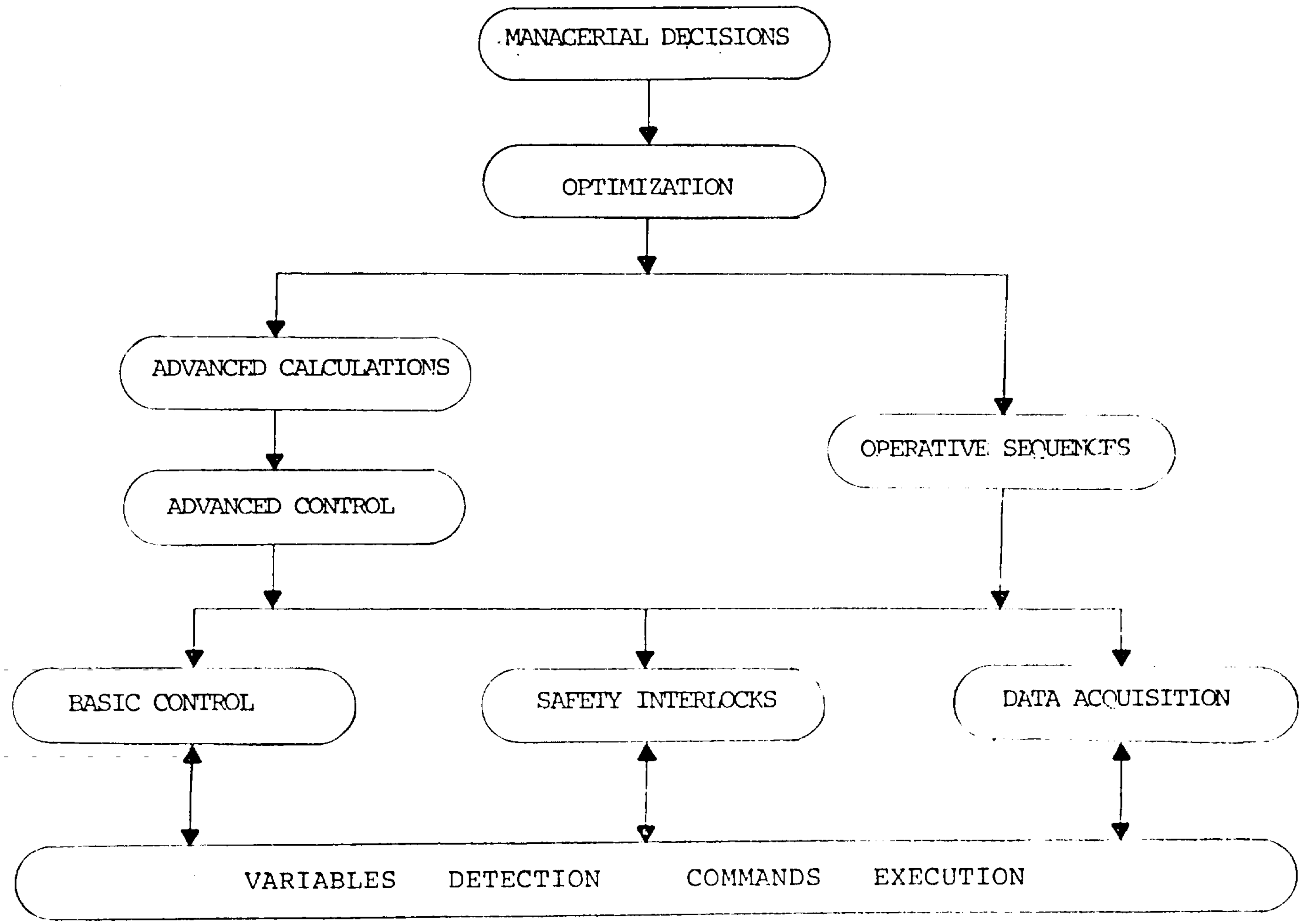

MANAGERIAL DECISIONS
OPTIMIZATION
ADVANCED CALCULATIONS
ADVANCED CONTROL
OPERATIVE SEQUENCES
BASIC CONTROL
SAFETY INTERLOCKS
DATA ACQUISITION
VARIABLES DETECTION COMMANDS EXECUTION

FIG.2.2.1.

Each module sends information to the higher level modules and receives from them its own set-point.

The lowest level modules are basic control elements with direct access to the field measurements and valves.

The advanced control modules are located at an intermediary level. The relevent calculations can be executed by distributed microprocessors or centralized process computers.

At higher level the optimization modules can operate"on line" by direct elaboration of the information and automatic determination of the inferior set-points, or " off line" as operator guidance. These modules are normally located in computers or distributed microprocessors with high calculation capability.

Generally, basic control - advanced control - optimization are defind as follows:

Basic Control

is the first regulating level. Its objective is to assure the return of the process to its normal operating conditions when a disturbance has provoke an upset. The relevant control strategies are usually implemented with well known algorithms like PID (Proportional + Integral + Derivative), simple arithmetic calculations, signal selections , signal selectors, etc...

Traditionally these strategies were and are executed by means of conventional analog controllers.

However nowadays beeing an actual tendency to implement more and more control strategies of superior level, the use of microprocessors is becoming popular because they are more accurate, reliable, flexible and suitable to easily communicate with other digital elements.

Advanced Control

is the ensemble of all control strategies devoted to aviod that an upset occures in the process, and not to correct it, this is a basic control task!

Therefore its objective is to obtain a higher level of operational stability.

The advanced control strategies can be defind as feedforward (predictive) ; by means of them, upstream disturbances, of the process unit under control, are compensated with corrective actions to avoid their downstream transfer.

Advanced control schemes could be implemented to avoid upset caused by fuel gas pressure and composition variations.

It is intuitive that some calculations in real time are needed, and that they will be more complex than the ones needed for basic control.

Typical algorithms are, for example, [illegible] flow calculation of the process dynamics, etc...

The more accurate the prediction and compensation of the effects of an upstream disturbance, the higher process operational stability will be reached.

Efficiency calculations, mass balances, energy balances, signal validations, etc.., are also domain of the advanced monitoring and control.

Optimization

of a process, which has already reached a good level of stability, is a control straregy which assigns set point values to a series of controlled independent process variables in order to maximize or minimize objective function.

The most common functions are for example the profit, the efficiency, the product cost, the losses, the production, etc...

Optimization must not be confused with advanced control. The first one implies an iterative search of independent variables' values to drive the objective function to a maximum or minimum limit; the second one executes all the calculations and actions needed to maintain stable the operation.

High process stability does not mean optimal operation, but it is a necessary condition to allow the research of optimal operative values.

For instance, optimization of a single industrial furnace means to search the combustion excess of air value (independent variable), wich minimizes the heat losses (objective function).

2.3. Description of the tasks and control solutions performed.

2.3.1. Project area services at IPCL - Baroda as per P.2.02.a/ of the Contract.

Under sub-contract No 1, project area services J.V.SYSTEMATICS team during the period from January 19,1986 to February 19,1986 consisting the following members:

Name	Function	Duration
Mr.G.Nikiforov (Team Leader)	Automation Engineer	19.01.86 to 19.02.86
Mr.V.Donchev	Process Techno-logy Engineer	19.01.86 to 09.02.86
Mr.P.Skenderov	Process Control & Instrumentation Engineer	19.01.86 to 09.02.86
Mr.V.Pankov	Optimization System Engineer	02.02.86 to 19.02.86
Mr.D.Dimov	Digital Control System Engineer	02.02.86 to 19.02.86

carried out the following services:

1. Study in detail the existing pneumatic instrumentation in DMT Plant and suggest in detail changes for implementation of electronic instrumentation integrated with distributed digital microprocessor based process control system. For this subject a recommendation report was prepared and given to IPCL. This report includes a specific detailed suggestion for electronic instrumentation implementation and a configuration of a distributed process control system

2. Collecting a background information for the existing on-line analyzers at IPCL.

3. Organizing training of IPCL instrumentaion and process engineering staff on implementation of distributed microprocessor based system integrated with supervisory level computer for appropriate period at IPCL site

4. Study the present pneumatic instrumentation in the LAB,ACN and EG Plants and give some general and particular suggestions for improving the instrumenation and for use of electronic instrumentation and process control systems.

5. Collecting the necessary background information for suggeting of integrated microprocessor based system for IPCL Boilers control.

6. Discussion and suggestion of some improvements in the instrumentation and process control strategy for particular IPCL problems.

The expert team recieved for use at IPCL Complex a background information for above mentioned plants only/P&I diagrammes for LAB,EG,DMT, ACN and Boilers/ as per IPCL requirements.

2.3.2.Home - office services and training of IPCL engineers

Due to the discussions and requirements of IPCL, SYSTEMATICS prepare the present Final Report on the basis of the above mentioned backgroung information.

The main control solutions suggested in this report can be summarized as follows:

- Recommendations for revamping of the existing pneumatic type control systems with implementation of modern digital distributed control systems, based on TDC-3000 System of HONYWELL, are made.

The hardware and software configuration for succesfull implementation of these new process control systems for DMT, Ethylene Glycol, Acrylonitrile and Linear Alkyl Benzene Plant are presented.

- Suggestion for implementation of new advanced control schemes and optimal control strategies for the abovemensioned plants are given together with the required new instrumentation for its realization.

- Detailed suggestion about installation, implementation and commissioning procedures for digital distributed control system realization are included in the report.

- An Integrated Computer Control Sytem is developed, presenting in detail the hardware and software modules of the system, the communication system and physical distribution of its modules.

- Detailed recommendation for the revamping of the existing and planned to be installed instruments of the DMT Plant with application of electronic type instrumentation is also presented in the report.

- Options, if feasible, are given regarding the step by step implementation of hardware and software specifications of the modern automation system and the field instrumentation.

- Possible economical benefits and savings which can be achieved implementing advanced and optimal control strategies procesed for each plant are justified.

- An Integrated Industrial Utilities Management Software system is suggested together with the required for it application hardware solutions and expected benefits.

- General recommendations about modernization of the control system for each plant and for revamping of used in IPCL on-line analytical instruments are given at the end of the report.

- SYSTEMATICS and HONEYWELL engineering capabilities in the future realization of the suggested process control and energy management system modernization are presented.

As per SYSTEMATICS proposal from August 15,1985 a one month /from April 12, to May 12,1986) training of the following four IPCL Engineers :

Mr. C.Z.Mehta - mechanical engineer from planning department

Mr.Gorana - instrument engineer from GAP

Mr.A.N.Shah - instrument engineer from GOP

Mr.Rajgopalan - technical engineer from GOP

in Microprocessor Based Distributed Process Control Systems Implementaion at SYSTEMATICS training facilities in Sofia and at the petrochemical complexes in Bourgas and Yambol (Bulgaria) where such digital systems are implemented , was organized.

SECTION 3

3. APPROACH FOR AUTOMATION SYSTEM ENGINEERING WORKS.

3.1. Approach for plant optimization

On-line optimization requires three programs: the process model, the executive function and the constraint driver (Fig.3.1.1.)

3.1.1. The process model predicts the value of the dependent variables for every set of independent variables. Industrial processes can rarely be described by an exact analytical expression. Most frequently one uses approximate models which are then constantly updated from process data. These models can be very simple provided that they are used in the neighbourhood of the current operating point. Prior to their use in the calculations, process data must be adequately averaged and compensated to correct for fluctuations and inaccuracy of measurement.

3.1.2. The executive function determinates the new operation point which will improve the objective function. Step by step the process is driven gradually towards the optimum. Two cases are encountered (Fig. 3.1.2.)

Unconstrained optimum is a local optimum such that any departure from the optimal operating point is a decrease of the objective function.

Constrained optimum is met when the operating point hits against a process constraint. The optimum will then depend on the variation of the constraints.

3.1.3. The constraint driver checks the new values of the independent variables versus process constraints. When the operating point hits a variable constraint, the program keeps the process against that changing limit.

The problem is that the concepts of optimization and advanced controls have overlapping boundaries and they can become nebulous, confusing terms. For instance, if advanced controls are arranged in a hierarchical structure, in which the upper level controls, can reset the set points of the lower level controls, then the upper levels can easily include optimization techniques. Because the different levels in the hierarchy work together , the entire structure is referred to as optimization even though some of the controls are purely regulatory in nature.

To avoid misunderstanding, it is important to clearly define what is meant by the term optimization. It should be explicit whether the task is single unit operation optimization (i.e.single furnace efficiency maximization, these are

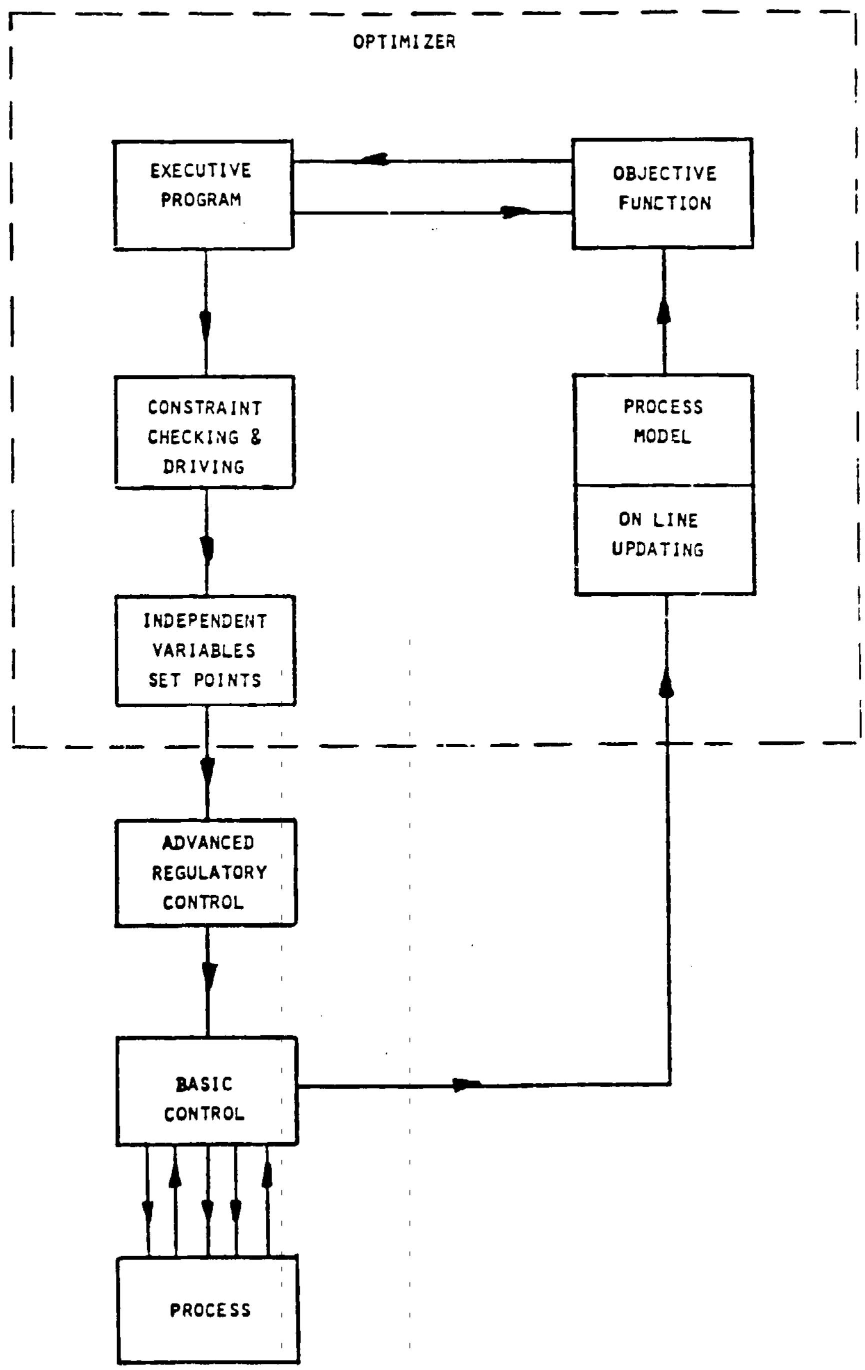
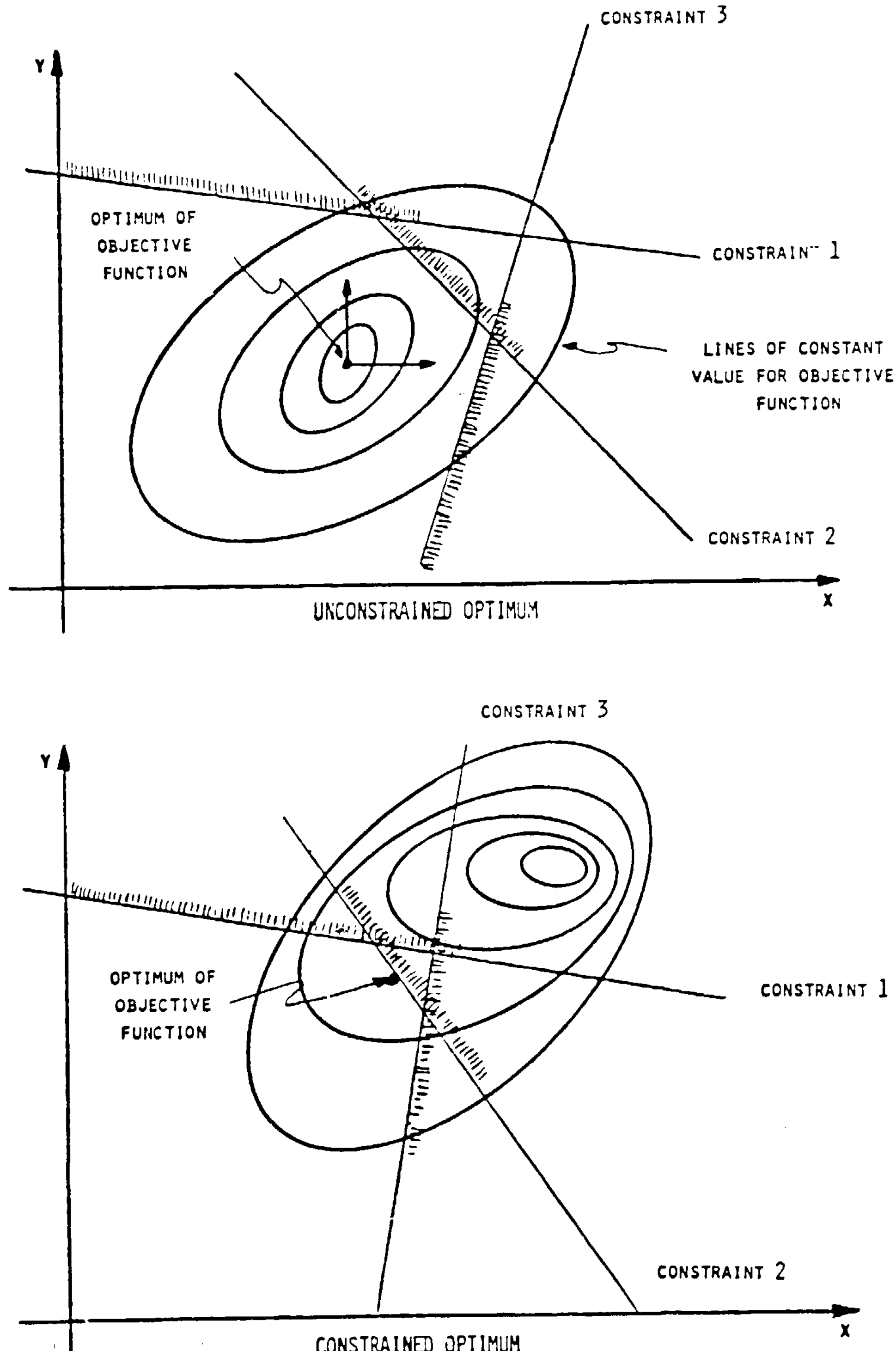

OPTIMIZER
EXECUTIVE PROGRAM
OBJECTIVE FUNCTION
CONSTRAINT CHECKING & DRIVING
PROCESS MODEL
ON LINE UPDATING
INDEPENDENT VARIABLES SET POINTS
ADVANCED REGULATORY CONTROL
BASIC CONTROL
PROCESS

FIG. 3.1.1.



X,Y = INDEPENDANT VARIABLES

FIG. 3.1.2.

possibly sub-optimal in terms of overall plant economics) or whether the task is overall plant optimization (i.e.: crude unit optimization).

Furthermore there are two methods of approaching the optimization of an entire process:/see Fig.3.1.3./

- the Total Plant Model Approach
- The Modular Hierarchy Approacn

Six points can be made about these approaches:

Total Plant Approach	Modular Hierarchy Approach
1.The entire optimization procedure must work or none of it does.	1. Parts of the control scheme and some optimization procedures can be turned off without jeopardizing the remaining parts.
2.Development time would be exessive .It would be years before any benefits would become apparent.	2.Some benefits could be obtained quickly.Continuing project management could decide whether or not to commit further resources to obtain more benefits faster
3.Total Plant Model Optimization has never been satisfactorily demonstrated by anyone.	3.Techniques for hierarchical control have been demonstrated on many industrial plants.
4.Total Plant Model could over tax computer(primarily computer free time of the CPU)	4. Modular Hierarchy Computer resources allocation can be estimated by known methods.
5.Total Plant Model would be incomprehensible to plant operators, management and to all except two or three technical people.	5. The modules can be made understandable to every one.

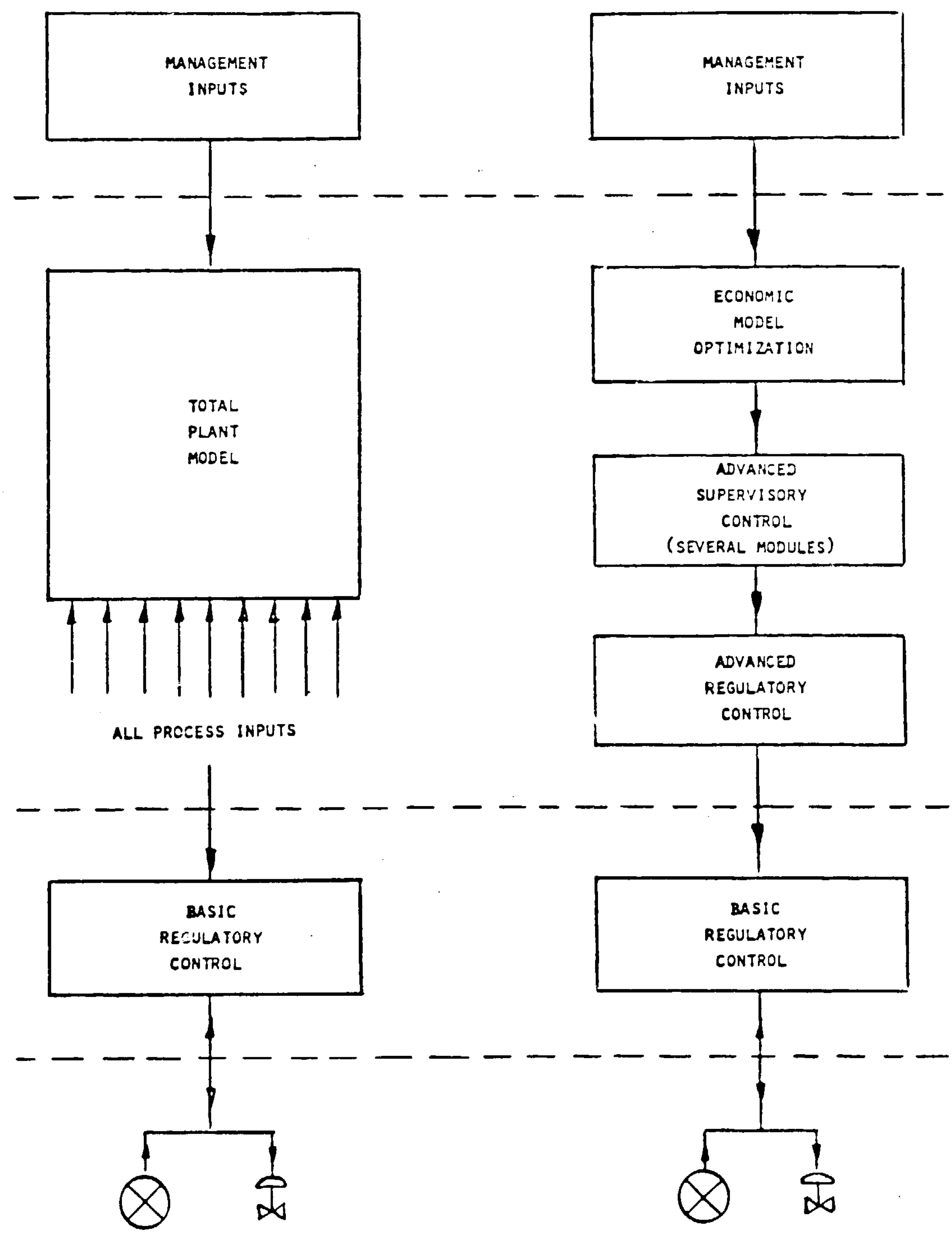

MANAGEMENT INPUTS
TOTAL PLANT MODEL
ALL PROCESS INPUTS
BASIC REGULATORY CONTROL
MANAGEMENT INPUTS
ECONOMIC MODEL OPTIMIZATION
ADVANCED SUPERVISORY CONTROL (SEVERAL MODULES)
ADVANCED REGULATORY CONTROL
BASIC REGULATORY CONTROL

FIG.3.1.3.

6. Total Plant Model should only be applied after the plant is under good regulatory control. Thus the regulation problem still has to be addressed.	6. Advanced regulatory controls are lower hierarchical levels. Thus, good regulatory control is achieved before reaching the upper hierarchical optimization control levels as the project develops.

These six points are sufficient to show that the Total Plant Model Approach is a high risk approach to plant optimization. The Modular Hierarchy Approach is definitely preferable.

3.2. Approach for technical support in the future engineering works.

The aim of the study presented was already described in item 3.1.of our proposal.However, this material can serve only as a basic information for a brief scope of engineering tasks, which are to be solved successfully with common efforts from the customer and the contractor. It is essential to be understood that no one of this two sides are able to carry out all work itself.There is a common belief that application programmes exist in the form of packages. It would be desirable if they were indeed available, but they are not, because of the following reasons:

1.Plants have different kinds of unit operations(equipment)

2.Plants have different process design arrangement.

3.Customers have different objectives and operating procedures.

4.Customers have different feeds and products.

5.No vendor has a large enough volume of specific process application work to justify application package development.

The lack of existing packages results in the contractor and/or customer having to design many applications specifically for each project or to make extensive revisions in previously designed applications.However,contractor experience with similar plants and processes can be valuable in minimizing the design efforts and insuring the project success.

The design requires most of the manpower resources prior to the computer startup.It generally begins with a plant visit by the contractor, which lasts from several weeks to several months,depending on the size of the project. The contractor and customer must agree on the functional specifications for the selected applications.After the functional specification are issued, the detailed design is completed by the contractor and the basic information is prepared by the customer.

The design normally coincides with the control system constru ction period.It closes with a test demonstration of all of the controls at the vendor site.

The project engineering work can be devided in two main parts,as follows:

1.Basic Engineering

General-to specify standard forms for application engineering.
Process related-to lay-out the system from basic PID

2.System and application engineering.

Control room design (operator interface)

-controls divisions

-control room functions

-operator stations lay-out

Measurement and control design.

-detailed loop specification

-loop drawings

-logic diagrams

Configuration

-programming of process control schemes.

-data base generation

Construction and system design

-system diagrams

-cabel connection drawings

-power supply

Installation design.

-mounting drawings

-mounting material lists.

Supplies design

-project purchasing.

-documentation design

-assembly of documentation

-shipping of documentation

Test engineering.

-application testing

-correction of loop DWGS and configuration lists (testing document)

Installation engineering

-supervision of site installations

-start-up co-operation

-installation documents completition

This two main parts are performed in accordance with the appropriate time shedule in five phases.
In summary ,the contractor sees its role in the project phases to be the following:

I.JUSTIFICATION -To help the customer get approval for the project by justifying all necessary basic information.

II.DESIGN -To work with the customer to specify the functions and to design the details of the controls to be implemented

III.TRAINING -To train the customer to use the system quoted.

IV.IMPLEMENTATION-To provide start-up assistance by working with the customer to turn on those programmes designed in Phase II and to initially tune them.

V.MAINTENANCE -To provide backup assistance if the customer has problems which he cannot resolve

From this summary it should be evident that we cannot recommend a traditional turnkey project approach to projects like these. A turnkey approach means that the contractor would do everything through implementation. Such a project is almost assured to fail in the maintenance phase because the customer would not be able to maintain or add control function.

The only way to project success is the development of a team approach in which the customer and the contractor are commited to work together to make the project profitable both short and long range.

Phases I and II are essentially clarified in the present material.There are some considerations which have to be kept in mind,looking on the rest phases.

Phase III-Training
Also during the system construction period, the customer engineers are trained at a contractor training center.Training is important for the implementation and maintenance phase which follow.It is important that the customer make a significant and continuing commitment to the training of personnel.Projects can easily fail if only one person is trained and he changes his job assignment.

Phase IV-Implementation
This phase begins when the system is installed in the plant and the hardware is tested and accepted by the customer.The implementation period ends when all of the programs in the functional specification are operational.This phase can last for many months depending on the scope of the functional specifications. The reasons for the length of this phase include the following:

1.All controls cannot be started at the same time. Each module requires a check-out period during which the parameters are tuned.

2.Problems and delays may be encountered with measurement devices.

3.Plant operating procedure may have changed.

4.Unexpected design deficiencies may require design changes.

5.Manpower problems may be encountered (either customer or contractor)

As a result of these potential problems , this phase tends to be open-ended.Even though applications are written and ready for implementation , the actual start-up of some programmes may have to be delayed.For this reason, the contractor has to limit his implementation phase manpower to a fixed amount, normally called start-up assistance.

Phase V-Maintenance

After specified functions are implemented,the project enters the maintenance phase which is the responsibility of the customer.The contractor is available for assistance in the case of problems which the customer cannot resolve.This phase requires more customer manpower than generally thought because of the following reasons:

1. Application modules are not static; they are dynamic. Changes in the programmes are necessary to handle new problems,program improvements, changes in plant operations and procedures, and so forth.

2. New functions beyond the scope of the initial project are constantly desired.The flexibility and power of the computer allows for many new ideas which are only limited in application by the available customer manpower.

3. Additional engineers must be trained as replacement for transfers, vacations and sickness relief.

System control project have failed because the maintenance phase was undermanned or manned with untrained personnel.Manpower is the usual limiting factor in this phase, not control system capabilities.

3.3. Approach for definition of economic and social benefits.

For a given unit potential savings depend on factors like:

- Status of equipment /age, design, etc./
- Type and condition of the instrumentation
- Control schemes used.
- Quality and experience of the operating personnel

The last three of these factors determines how far one is from the maximum possible unit efficiency, while the farest one sets the upper limit to the efficiency, which is possible to achieve for this unit.

Project for improved process control by distributed control and/or computer system should be justified by the evaluation of the credits relative to the costs of installation and implementation of the control system.There are generally three kinds of credits which arise from improved control. These can be refered to as "hard","soft" and "intangible" credits.

Hard credits are those which are claimed with relatively good assurance that they can achieved. They are often associated with energy usage in the petroleum and petro-chemical industries. A useful technique to substantiate hard credits is to plot an energy consumption variable against a capacity indication variable to establish the average performance and best operator performance.The assumption is made that the product specifications are never violated. The difference between the two performance lines i[illegible] realistic, albeit conservative , measure of the amount of improvement that can be realizes with advanced control techniques. In order for a credit to be considered hard , the variable which will be improved must be translated unambiguously into cost savings. An example of a hard credit is reb[illegible]er steam consumption on a distillation column. The average and best operator performances of steam as a functio[illegible] column feed can be readily obtained and the steam saving[illegible] easily calculated .

Soft credits are those which are cla[illegible] with less confidence than the hard credits.Soft credits results from the lack of good , available data, the lack of firm economic bases, the possibility of encountering unknown constr[illegible]nts , or the combination of these factors.An example of a soft credit can also be provided by a distillation column. In this case,the improved control benefits are not apparent from past and present operating conditions, because the tower will be run in the future under different conditions with improved control.For example ,additional energy could be saved if overfractionation was minimized so that products were not purified more than their specifications allow.However the question arises as to how much energy can be saved by doing this. To answer the question , the choice is between plant test and simulations studies.Both of these are time consuming and expensive.The possibility of encountered difficulties of the values for intermediate product streams further complicate the situation.

Intangible credits are those which are known to exist, but for which no possible quantitative estimates of savings are possible. For example it is generally agreed that computers provide better information to operators and engineers in the form of graphic displays and reports. It is agreed that using computers allows the operators to spend their time checking and optimizing the process rather than recording numbers on log sheets. It is agreed that better regulatory control leades to longer equipment life and fewer incidents of product losses.It is also agreed that no one can calculate the magnitude of these intangible savings.

If at all possible, projects should be justified by hard credits.The dependency of projects on soft credits increases the project risk and leads to difficulties in verifying the benefits of improved control after project completion.If further justification is needed beyond hard credits, the firmer of the soft credits can considered.However, the risk to the project increases rapidly as the dependancy on soft and intangible credits increases.The presented study should work credit estimation only far enough to justify the project.The additional soft and intangible credits only enhance the probability of increased profits and the chance of project success.

There are some well known ways to find the possible hard credits,if old instrumentation is replaced by a digital system, and modern control system schemes as described in Chapter 4 are implemented.

This is due to:

- the performance of digital electronic wich is more accurate than the analog one.
- the high level of operational stability assured by the advanced control strategies.
- the higher unit stability because of operator safety margin from the constraints minimized.

The normal way to find out the possible hard credits is to determine the energy consumption variable against the capacity production of the propriate unit (see fig.3.3.1.)

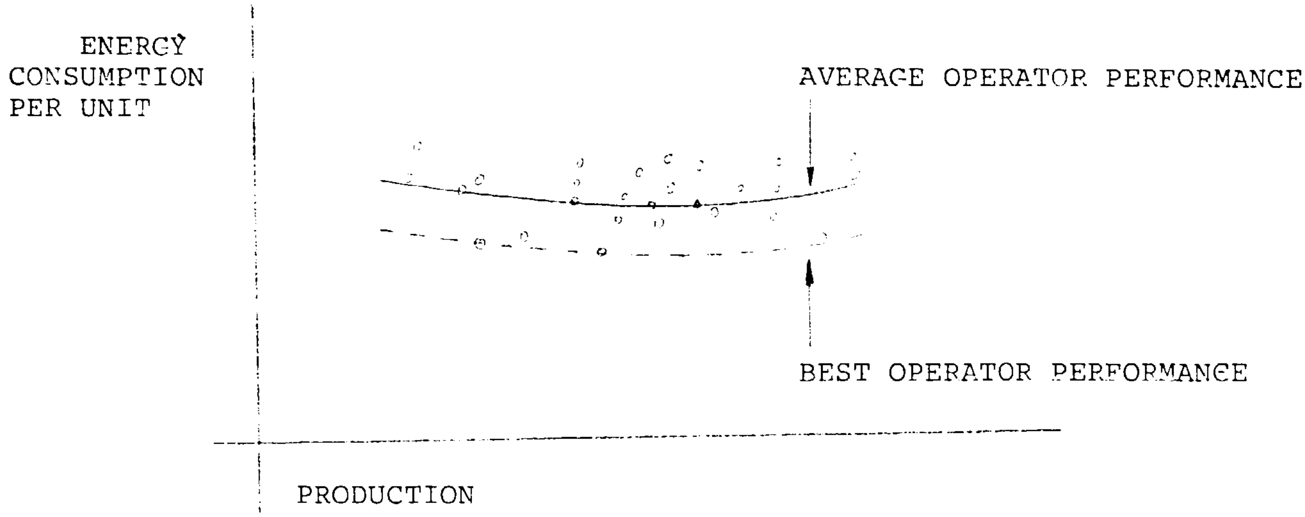


FIG.3.3.1.

The next step is to clarify the possible throughput due to the best operator line.(see fig 3.3.2.)

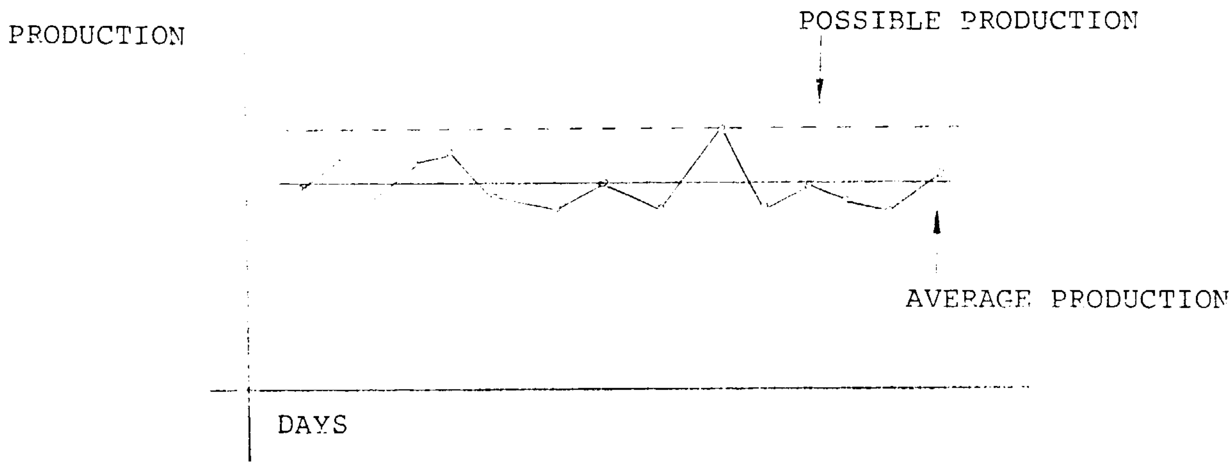
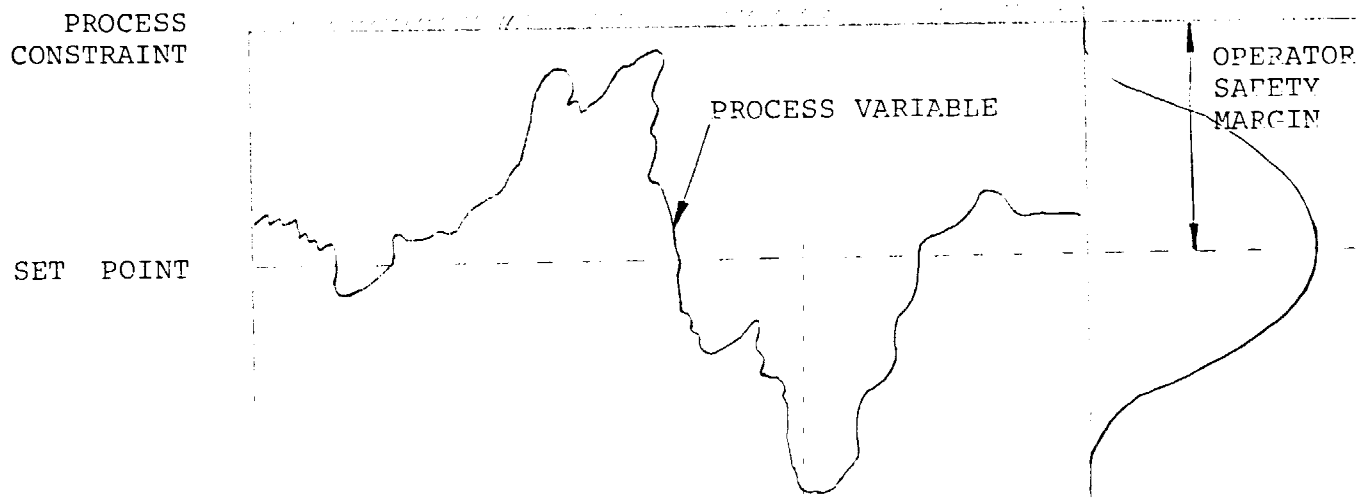
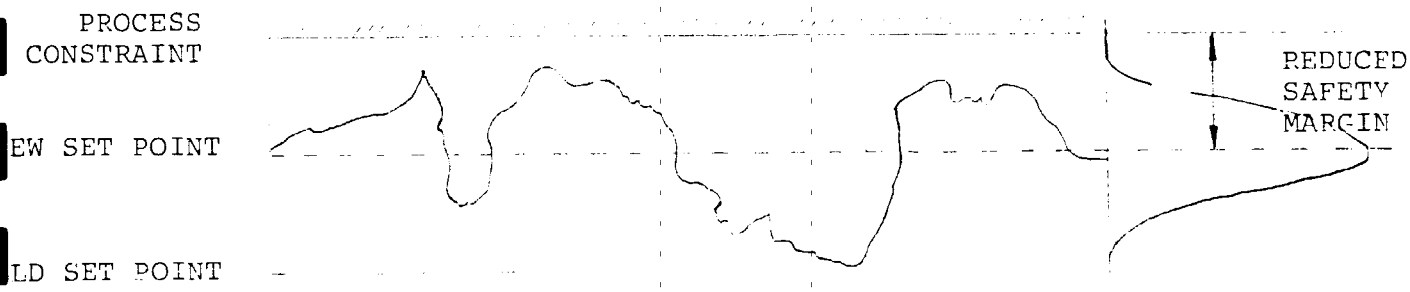


FIG.3.3.2.

These two justifications are to be made for the main energy consumable units and units with big production capacity in the complex.The justification have to be made at least for one month duration for both "Summer" and " Winter" period.Usually it leads to proving savings from 5-2% due to the replacement of analog with digital instrumentation only.
Optimal operation requires maximum unit stability to permit operation closer to the constraints.
Instabilities occur when changes in process conditions are not well controlled and result in oscillations around the steady state. It is clear that the smaller the oscillations,the closer to the constraints the unit can operate.(see fig 3.3.3.)



WITH UNSTABLE PROCESS



WITH STABILIZED PROCESS

Fig.3.3.3.

In practice, the operator is familiar with the dynamic behaviour and sets the set point to stay safety away from the constraints.Similary , he sets the unit always to make products purer than required.

Obviously this leads to some loses from unused unit's efficiency.

For all process disturbances, the principle of the solution is the same

a/ exactly measure the process perturbation(temperature, flow rate, feed composition...)

b/ accurately predict the corrective action (dimension of the change , when should be it applied)

c/ send the corrective action to the relevant process control valves (one or several)

d/ check the result of each action versus the desired result and correct for any deviation.

e/ for control valves and transmitters,select characteristics and operating ranges appropriate to the type of control.

The effect of this solution can be calculated only after the implementation of the propriate system. It can be determined on the basis of comparison between previous and new ones set points. Evidently this effect causes the soft credits.

3.4.Possible risks.

There are several things that can go wrong with computer control projects.Comments are made about each potential risk.

1. Hardware may not work.

The HONEYWELL TDC System is fully demonstrated. Many install- ations already exist . There is no chance that the hardware will not work.

2. Software may not work.

A.System.The operating system programs have all been fully demonstrated and debugged.

B.Applications. In light of the previous discussion, all application programs are unique.However the tools for building the programs have fully tested and similar programmes to those described herein after have been implemented in other plants.

There is no chance that software will not work.

3. Personnel may change jobs.

There is always a chance that trained people will become unavailable to a project for one of several reasons.JV SYSTEMATICS has enough application engineers that there will be an available replacement for a transferred engineer.A bigger risk is the loss of a trained customer engineer.The customer must have at least one backup engineer so that the project does not depend on one person.The customer must commit to a continuing training effort to ensure the existance of trained personnel.

4. The plant may change.

Short of the completely shutdown of the plant, the HONEYWELL TDC System is versatile enough to handle changes in plant operations, shifts in control emphases, in changes in constraints and economics.In fact it is common for projects to make more money than originally forecast but to do so in ways completely different than forecast .Here is where the flexibility and ease of programming prove themselves valuable.

In summary, the primary risk to this project will not be due to hardware or software problems.The primary risk will be the possible shortage of trained customer manpower .

4.SYSTEM DECISION.

4.1. DMT Plant

The existing process units and the project documentation for Phase I expansion of the plant have been deeply examined and some areas where possible benefits can be obtained by application of advanced control strategies have been identified.

4.1.1. Background information

A/ Analysis of the existing automation system

The existing process instrumentation and process control system in DMT PLant is analog type, partly centralized in the main control room.It consists of mainly pneumatic instruments, manufactured in India under license from Taylor and few electronic ones. The temperatures for indication are measured by RTD's , connected to electronic temperature indicators and recorders.

In the plant there are implemented three Oxygen analyzers , produced by Hartman & Braun /FRG/.

The control valves applied are produced by Gulde. The main part of the control loops are centralized in the control room, but there are also few local /field mounted/ control panels. The existing instruments are in a relative good condition. However the pneumatic instrumentation due to its inherent limitations does not allow an application of advanced control.

B/ General recommendations to the process instrumentation and control system.

1. On the basis of the detailed study of the existing control system and instrumentation in the plant as well as their own experience in the field of implementation of distributed digital control system (DDCS) in the DMT Plant in Bulgaria Systematics experts are suggesting that the instrumentation in the existing DMT Plant requires to be modernized to meet the requirements of proper measurement, accuracy and control of (Process) parameters (i.e. temperature, Flow, Level, Pressure, Analysis of O2 etc.), utilities and energy consumption in near future - During the expension of the production to 40000 T/Year.

2. Systematics advises the implementation of a distributed Microprocessor based digital control system for the whole DMT plant (i.e. all 10 process stages) The Units of the system should be placed and mounted in the existing control room after modernization.

3. Systematics suggests, on the basis of their experience in DMT and Aromatics Plant control, increasing of the centrallization in the control room. Local instrumentation panels and local profile programmers should be changed with a new instrumentation using microprocessors in the control room. The number of centralized control loops including cascade and ratio control as well as the number of centralized indications and remote control for binary variables (e.g. Remote control of pumps, switches and alarm indicators) should be increased.

Total number of controlled and remote operator from the control room parameters (variables) is suggested to be about two times more compared with the existing number of the centralized (in the control room) parameters.

4. Appropriate combination of electronic and pneumatic instrumentation is preferrable but the number of electronic instruments and loops should prevail. For some points pneumatic field instruments can be used with appropriate P/I converters.

5. Programmable logic control and Mathematical modelling on the base of Microprocessor equipment can be utilized for the optimization of batch type stages and units e.g. Oxidators, Crystalizers, etc.

6. Systematics has got about 6 years experience in using of Microprocessor based distributed control system for stabilization control used with implementation of sophisticated supervisory computer functions. With this hierarchical approach and energy management functions, Material and Energy balances, etc. can easily be realized.

7. The design of the digital control system is recommended to be organized in few stages with care for minimum disruption of the production and minimization of the start-up time. The functional capabilities of the new system should fulfil all requirements and functions of the existing conventional instrumentation system such as :

- implementation of closed loop control as well as cascade ratio and sequence control;

- on/off remote control of pumps and valves;

- analog and digital (e.g. temperature selector, etc) displays in the control room;

- trend curves for analog variables, e.g. the existing recorders will be changed with trend functions of the digital control system;

- logical, sequence and programmable logic control of set-points and of binary variables'

- different type of alarming with logging functions/printing of alarm reports;

- manipulation of all type control parameters. From an information processing point of view the electronic process monitoring and control system should be designed to perform the following main functions:

- Distributed process interface using different Microprocessor based devices;

- Process control and optimization;

- Data acquisition and transmission;

- High- resolution and easy for operation

Man-Machine interface by implementation of specialized operator stations with keyboard;

- Digital Back-up of the most critical control loops and programable control functions.

8. The detailed suggestions of the instrumentation revamp and implementation of electronic instruments, positioners for control valves, temperature measurements, etc. are given in section 4.1.2.- Instrumentation recomendations.

9. General advices and recomendations for using of microprocessors based process and production control system are specified in a section 4.1.3.

10. Systematics is suggested to realized the process control system modernization in steps with paralel training of DMT staff in operation and maintenance of the system.

11. The design of the DMT digital process control system should give a possibility for extenssions also for the Xylene production unit with decreasing of the necessary capital investments.

4.1.2. Instrumentation Recomendations.

4.1.2.1.General

1. Systematics team suggests replacement of most of the existing pneumatic field and control room instrumentation with electronic type ones,e.g.Temperature,Pressure, Differential Pressure & Level Transmitters.

2. We would suggest to modify the existing Nitrogen(N2) supplying system for every production Unit in the Complex as follows:

- To realize a separate Nitrogen supply system for instrumentation needs in each production process unit such as to realize a direct Nitrogen supply of the instruments from the inlet receivers.

- Nitrogen pressure to be controlled automatically by PID control loop with low-pressure alarm.

- To study the possibility of modification of the existing N2 system in DMT Plant with implementation of high pressure vessel and old existing piston compressors.

We would suggest to make better maintenance of the existing Nitrogen rotameters system by testing in the maintenance workshops (Especially the rotameters, the relays and the needle valves) and after that the second step can be the replacement of the existing in the DMT Plant differential pressure transmitters with Electronic type.

3. The implementation of Microprocessor based electronic control system does not require replacement of the existing control valves and actuators if there are not other constraints and considerations.International practice is when electronic instrumentation is supplied to use a separate I/P converter and pneumatic positioner due to easier adjustment and low sensitivity to variations and dynamics.

In the cases when Kv and dP are high usually the use of double seated valves is recommended if high tight shut off is not required. Considering decreasing of the valve sizes we suggest the often use of valves equipted with actuators and positioners for higher pressure /2,5 to 4 Atm/

4. Cut-off valves used in interlocking systems and in batch type process units have to be equiped with limits switches /Microswitch or proximity switch type/ for the closed position and if it is necessary for the open position too. These valves have to be from 90o rotating or single seated

type supplied with appropriate sertificate for tighteness or leak percentage.

Electromagnetic type valves have to be mainly implemented for small sizes with sealed coils/epoxy sealed or safety sealed type/.

5. Application of infrated and paramagnetic type analyzers is recomended for measurement of O_2,CO_2,CO contents of the gasses.Because of the high dust content in the circulating Nitrogen in stage 8 additional devices for automatic cleaning and powder removal in sampling systems is preferable to be implemented.

We would recommended the use of "Magnus" and "Uras" type analyzers of H&B - FRG or " Servomex" of Taylor U.K., specially for O_2-Measurement. The connection of the output signal of O_2 analyzers to the electronic process control system can be realized via mV/I converters in the control room.

6. Critical for the process operation alarms can be hardwired to the control room and equipped with individual alarm indicators.

The same digital signal can be wired to the process control system as well.

7. The list of all instrumentation which shall be used in DMT Plant after the Phase I expension /to 40000 MTA production capacity/ with detailed Systematics suggestion is attached in ANNEX 1 ,Table 4.1.2.1. to 4.1.2.9.

4.1.2.2.Flow Measurement

1.In the specific conditions of the DMT Plant, where flowrate of high viscosity fluids in high temperature conditions are measured, it is necessary to implement rotameter type transmitters with steam jacketing and pneumatic signal ("Krone" can be a suggested supplier).In this cases the pneumatic signal can be converted in this case via P/I transduccers, mounted in the control room.

2.For relative clean fluids it is preferable to decrease the number of supplied rotameter type transmitters by using of dP transmitter with orifice plate or with integral orifice with electronic output signal for flow measurement.

3.Systematics suggests to use electromagnetic flowmeters with electronic output signal for high accuracy measurement of liquids,for catalyst dosing or for accurate ratio control,etc.

The implementation of rotameter type transmitters is avoided because of the low accuracy measurement (1,5 to 2,5 %) and possible sediments.

In accordance with the international practice all rotameters have to be tested every year.We would suggest to IPCL to consider a possibility of creation of rotameter testing station .This test station can be extended in future for other type of flowmeters also and first of all for positive displacement meters.

If the implementation of rotameter transmitters can not be avoided then the recommended supplier is "KROHNE" with electric output signal for nonjacketed and low-temperature services.

4.For the material and energy balances of the DMT Plant we suggest to implement measuring instruments with temperature and pressure compensation including inlet and outlet flow of utilities. Instrument output signals should be centralized for indication and control in the control room. For proper selection of this equipment the following practical advices can be given:

- The turbine counters could be implemented for big flow rates / over 100M3/Hour /of small viscosity products. The high velocity rotation, the sensitivity toward the product and the gas-bubbles limit their wide application for DMT Plant;

- For the case where better accuracy with totalising of the measured flow rates or where setting of integrated quantity is needed, implementation of positive displacement meters equipted with local indication /Dial & Totalizer / and remote transmition unit is suggested.

- dP measuring devices could be implemented for measurement of steam , vapour and normal liquid flows.

- For cooling water flow measurement annubar type sensors are suggested.

We suggest the implementation of silicon type electronic dP transmitter for these measurements.(suggested supplier is Honeywell).

4.1.2.3. Level Measurement

1. By the opinion of the DMT instrumentation maintenance staff a lot of problems result from level measurement with differential pressure transmitters with Nitrogen purging systems. The reason for that is the common nitrogen system for the instrumentation purging and for the equipment blanketing.

2. There is a positive international experience of using of radioactive and capacitive type level transmitters mainly for specific applications like cases of measurement of high corrosive mediums, suspending fluids, high temperatures and pressures, etc. In plants like DMT production electronic type differential pressure transmitter with remote sealing diaphragms are often used/preferable supplier is Honeywell/, considering the following limitations:

- A special flange and nozzle at the bottom part of the vessel is necessary to be supplied;

- Temperature compensation is needed;

3. For clean products a displacement type level measuere devices can be used. The replacement for part of the existing differential pressure level transmitters with displacement type is possible considering the following:

- The tanks have to be equipped with external chambers /and additional LG for control/;

- The level is not a fast changing process parameter and if good/stable/ pneumatic type level transmitters are available / from Eckard -FRG or Masonelan -USA/ then this type can be implemented with P/I converter, specially for high temperature applications or jacketed type.

4. Existing in the plant float-type level switches/which create problems due to vapour deposition or suspension/ can be partly replaced with:

- <u>Radioactive type</u>

Suggested supplier is E&H or Dr. Berthold -FRG. It is preferable to use Cs 137 instead of Co 60 source due to the longer half-life time. The implementation of radioactive level instruments has to be done after proper tank inspection and corresponding measurement "full-empty". In this case the fact that the receivers are very sensitive to the high temperatures has to be considered.

- <u>Capacitive Type</u>

Usually we suggest the E&H ,FRG level switches.In this case the main problem to be solved is the realization of good grounding of the vessels.

When Microprocessor based system is applied in the control room it is possible to provide additional alarm contacts,which can be used as redundant for level switches.

5. According to the international practice most part of the level measurements are equipped with local LG/Glass or Magnetic type/ for cross checking of the level transmitter output and indication.

4.1.2.4. Pressure Measurement

1. For high and low pressure alarms Systematics experts suggest the implementation of contact type pressure gauges with proximity switches/proximity inductive system/ supplied from Wika-FRG.

2. Implementation of separating diaphragms for pressure gauges and pressure alarms is highly recommended for corrosion active and desoblymated fluids.

3. As per international experience most of the pumps and each pressure vessel should be equipped with local PG and indicator. As it is indicated on P'I diagrams in the existing DMT plant the number of pressure gauges is not corresponding to /is much less than/ this practice.

4.1.2.5.Temperature Measurement

1. After detailed study of the existing instrumentation in the DMT Plant we/Systematics experts/ suggest a replacement of the filled temperature transmitters with combination of RTD and transducer like more efficient solution.

2. All RTD's, including the existing ones; should be connected via 3-wire scheme.Generally RTD's should be delivered duplex type, particularly when it is necessary to use more than one instrument/TI/TR or TI/TIC etc/

3. Implementation of filled thermostats is suggested to be limited only to the package units or for the casses when great vibration occures near rotating equipment.

4.1.2.6.Electronic Field Instrumentation

1. The general requirement for the Electronic Transmitters are as follows:

a/ To be two-wire type without force balanced and current feedback; sensitive element to be semi conductive, strain gauges or capacitive types.

b/ To be equipped with indicators or at least with indicator terminals without output signal breakout; equipped with damping devices; to be without mechanical adjustment devices; zero adjusting potentiometer to be accessible from outside.

c/ To be provided with technical documentation sets including:

- Certificate for intrinsical safety requirements & tests;

- Electronic and electrical wire diagrams;

- Installation and mounting manuals with necessary drawings; outline and installation dimensions;

- Service data and maintenance manuals;

- Workshop testing reports;

d/ When the proper mounting place for dP level transmitters must be chosen it is preferable to sellect type with zero suppression and elevation.

2. In the case of implementation of electronic equipment in hazardous locations it is necessary to achieve the appropriate explosion proof protection of the field instrumentation as follows:

a/ To use the intrinsical safety approach as the most advanced for instrumentation. Field instruments and connecting cables should correspond to the requirements of local or international safety codes (Cenelec - Europe, Baseefa - UK, PTB - Germany etc./

b/ The wiring design and mounting to be done according to the capacitive and inductive limitations for electrical signal cables /4-20 mA or RTD cables/ . Each pair should be twisted with own shielding and each cable itself is shielded and steel-belted.

c/ The cables installed on cable trays have to be protected from direct sun-snine influences, etc, we suggest the cverground racking installation separate for IS-and non-IS cables.

d/ Electronic transmitters, junction boxes, solenoid valves should be protected against direct sun radiations and water, steam and chemicals falling through cable glands and cover sealing surface;

e/ The grounding system of the control room should be designed in such a manner that the cable belt and the cable shield to have a common ground point.

f/ IS - protection for analog signals is usually realized by z-barriers by optoelectronic couples or Galvanic Isolation.For digital input signals / including proximity switches/ the safety is achieved by sertified power supply modules ,having non-IS output contacts.

3. Together with revamping of the process equipment in DMT plant it would be necessary to reduce the number of locally controlled and indicated parameters by centralization in the control room. Our recommendation is all local controllers/exept few in flaking section and stage 9/ to be centralized in the existing control room. The same centralization should be provided also for the local programmer units, recorder, interlock etc. Only local indicators, push - buttons, alarm lamps and ,manual back-up stations could remain on the local panels. We suggest to centralize in the control room the important signals/parameters/, alarm signals,on-off switch signals for the stage and the new flaking section.

4. Interlocking system should be realized on the basis of solidstate logic equipment with high reliability-suggested supplier can be Siemens ,Hima-FRG or Telemechanique - France .

The realization of complicated logic functions including the programmable and programmable logic control could be realized by the distributed microprocessor based system.

5. The power supply for electronic instrunmentation and for the distributed computer control system should be calculated with spare capacity of up to 50% for fluctuations /deviations/-220 VAC/ + 10% and -15% /50HZ +/-1 HZ/ A battery back-up supply is needed for up to 30 min. operation.

6. The implementation of the digital microprocessor based system can be realized on the running plant with using /implementation/ of the required 24VDC power

supply units, safety Z- barriers and I/P converters for the output signals from the existing in the control room pneumatic controllers. The same I/P transducers can be after that utilized for the output signals from the digital system applied.

Mounting and start-up of the distributed digital control system can be realized on the running plant also.

4.1.3. Recommendations for advanced control strategies for application of advanced controls

A/ Suggestions for application of advanced control

The description of the suggested advanced control strategies is given below and the Process and Instrumentation Diagrammes /P & ID/ in ANNEX 1 ,where the existing classical control strategies, implemented with analog controllers are represented by circles and the advanced controls - in rectangles.

Using the general approach of hierarchical control at the lowest control level, we keep the current control loop with which the operator is familiar. Then we add a succession of increasingly complex control functions. The strategy is structured as a hierarchy of separate modules. Each module is giving the set-point to the lower one and receives its own set-point from the upper one. At any point of the chain, the operator can break the cascade. The upper modules are then disabled and the SP /set-point/ remains fixed at the last value.

At the highest level, we find very complex algorithms performing on-line process optimization and data validation.

These modules are using process models and have a sophisticated programme structure described later.

<u>Stage 0</u> : Methanolysis and Catalist recovery
Advanced control strategy not identified.

<u>Stage 1</u> : DMT Oxidation.

The oxidation is carried out in batch reactors with continuous aeration.The main incentives for application of advanced control and optimization of this unit are as follows:

- The composition of the pT Ester, that is fed to the reactors, varies widely as a result of static resycles in the plant;

- The throughput of the units in relation to the composition of the oxidate or its acid number varies in a wide range, often wider than specified due to :

a/ varying pT Ester composition;

b/ controlling the process without on-line information for PTE and oxidate composition;

c/ without realtime information for current composition of reactor mixture it is not possible to apply optimal temperature and air flow control.

Objective: To minimize the batch cycle time /i.e. maximize the throughput/ and stabilize the oxidate composition.

In order to achieve this objective the following approach to optimal control is suggested (Fig.1 and 2 in ANNEX 1).

- The objective function is to achieve desired oxidate composition in minimum time;

- The manipulated parameters are :

  p Xylene/pT Ester ratio;

  temperature controller set point;

  air flow controller set point.

- The oxidate composition is characterized by its acid number. This variable cannot be measured online, so we can suggest to use an estimator, which utilizes online measurements to calculate an estimate of the acid number online;

- The maximum air flow is limited by the reactor level and p Xylene vapour carry-over which cannot be measured online, so again it is necessary to use an estimator;

- The estimator parameters are tuned periodically using laboratory analysis data;

- The optimization procedure is based on non-linear programming techniques.

Stage 2 : Esterification

The oxidate esterification is carried out in tray columns. Besides the main endothermal reactions, yielding DMT target product, side reactions, also take place producing undesided products like resins, methyloxide, etc;

The process is characterized by wide variability of raw ester composition.

Objective : To stabilize the raw ester composition around its optimum value (i.e. to maximize the yield of target product).

Optimal control elements (Fig.3 and 4 in ANNEX 1)

- Estimator of rawester composition/acid number, utilizing on-line measurement;

- Estimator parameters, tuning subsystem, which uses laboratory analysis data;

- Optimiser, including an objective function and an optimization procedure.

The manipulated parameters are :

- oxidate flow controller set point;

- methanol flow controller set point;

- feed temperature flow controller;

- outlet hot oil temperature controller;

Stage 3 : Raw ester distillation

a/ Raw ester distillation column advanced control.

In such kind of binary distillation tower the major problem is to stabilize the internal flow of liquid and vapour. This is achieved by regulating the reflux valve and the heat input valve.

The advanced control structure proposed for raw-ester distillation column C-308 control is shown on Fig. 5 in ANNEX 1.

The internal reflux is calculated and controlled in ratio with the mass flow compensated feed. The ratio is adjusted by a temperature controller at the top of the column, thus ensuring relatively constant top product quality.

The bottom temperature controller is cascaded with feedforward loop of the raw ester feed. The structure is intended to maintain a constant ratio between feed flow and reboiler heat duty.

b/ Raw-DMT Distillation advanced control

Fig. 6 in ANNEX 1 shows the proposed advanced control strategy for the raw-DMT distillation of reflux advanced control, reboiler heat duty advanced control and optimization control of the binary distillation column.

1. Reflux Advanced Control

The internal reflux is controlled by a hierarchy of modules. The lower level is the external reflux flow control as with classical control strategies.

The next module is a flow controller where the process variable in the calculated internal reflux.Indeed we want to stabilize the internal reflux which depends on the external reflux but also on the subcooling of the condensate vs.the liquid on the return plate.

Fluctuation in condensate temperature will cause variations in vapour condensation at the top and the net flow of vapour out of the column will therefore change.

The internal reflux is calculated as:

$$RI = RE\ (1 + Cp\ /\ \lambda\ (Tout - Tin)$$

where

Re = External Reflux
Cp = Specific heat of the condensate liquid
λ = Latent heat of vaporization of the condensate
Tout = Temperature of the outgoing vapour
Tin = Temperature of the reflux

The next module is a feedforward line which maintains the internal reflux in constant ratio to the feed. The column mass feed is calculated and ratioed tc the reflux with proper lead/lag and dead time compensation to account for tower dynamics.

The ratio between feed and internal reflux can be adjusted from two parallel hierarchical lines:

- from the interaction decoupling in feedforward, which reads the changes of heat input.

If the heat duty set-point changes while the feed remains constant (e.g. to change bottom product quality) the module automatically compensate the reflux set-point.The scope is to balance the internal flow liquid/vapour to keep a stable column profile.

- from the module of prediction of the internal reflux in feedforward. In case of variation of the desired product quality, it predicts the amount of increase/decrease of the reflux ratio.

The next level is the optimization algorithm which determines the distillate quality corresponding to optimum economic operation.

2. Reboiler Heat Duty Advanced Control

The reboiler heat input is controlled by the hierarchy of modules shown on Fig.6 in ANNEX 1.

The lowest level is a cascade control loop where the temperature of a sensitive plate/or sensitive tower bottom/ goes to a temperature controller resetting the hot oil flow controller. This is the strategy usually implemented with analog instruments.

At the next level is a kcal flow controller which computes the amount of heat actually transferred to the system and compares in with the required heat duty.

The heat duty is calculated as:

$$QH = F \cdot \rho \cdot Cp\ (Tin - Tout)$$

where
F = Flow rate of hot oil
ρ = Density of hot oil
Cp= Specific heat of hot oil
Tin,Tout = temperatures of the hot oil before and after the reboiler.

Because of the dual character of ther lowest level, the control hierarchy will also developed into two parallel structures:

- A series of modules from a feedforward line from the mass flow of the column feed. The structure is intended to maintain a constant ratio between feed flow and reboiler heat duty. This feedforward is not always necessary because in plate /bottom/ temperature due to the small capacity of the column. In this case, we will suppose that the section of the column is of the significant capacity. The temperature controller controls in feedback the ratio between feed and reboil.

- This ratio can also be changed from a module of interactive decoupling in feedforward, which reads the variations of the internal reflux of the column and automatically compensates the internal flux of liquid/vapour by action on the heat flow to the column.

- A next level of parallel action is the correction of the temperature controller set-point to compensate for tower pressure variations The pressure correction is calculated with the equatuion of Antoine:

$\ln(P+a) = b-c/(T+d)$

- Also in the control system of the bottom of the column, the highest hierarchical level is the optimization modules which determines the bottom product quality corresponding to optimum economic operation.

Stage 4 : First crystallisation

Advanced control strategy not identified.

Stage 5 : Second crystallization

Advanced control strategy not identified.

Nevertheless all sequence and logic control functions of the crystalyzers in the both crystallization stage is suggested to be realized by distributed digital control system using the specialized programming capabilities.

Stage 6 : Multidistillation

Advanced control strategy not identified.

Stage 7 : Flaking section

Advanced control not identified

Stage 8 : Vacuum and vent system

Advanced control strategy not identified.

Stage 9 : Hot-oil section

a/ Combustion advanced control

On Fig. 7a in ANNEX 1 a scheme for combustion advanced control for the hot-oil furnaces 0-941/1 and 0-942/2 is suggested.

Both furnaces are burning two fuels - oil and gas; the air draft is forced with a fan. The suggested control scheme shows the minimum required control strategy needed to obtain an acceptable regulation of the outlet temperature and of the excess of air.

The oil and gas volumetric flows are measured and regulated independently, then summed to obtain the total fuel feed.

The combustion air is regulated in ratio with the total fuel.

The ratio is trimmed in feedback by an oxygen percent controller (zirconim oxide type) installed derectly in the stack.

Two high and low signal selectors allow to give always priority to the excess of air when variations of fuel are requested from the furnace outlet temperature controller.

This basic control strategy represents what is commonly implemented with analog instrumentation. But the digital instrumentation allows to improve significantly the control quantity by means of additional sophisticated modules which, as we will see, will assure a higher stability to the process making it ready to be optimized.

The total furnace heat losses result from the exess of air losses and from the number fuel losses (CO presence in the stack)

It is clear ,that it exists only one excess of air value with which the furnace can be operated at the maximum efficiency. This value depends on the type of furnace, on its vetusty state, on its fouling degree, on its burners' adjustment, etc... In conclusion it is variable and different from the design one.

Let's make an ipothesis that this value can be individuated, i.e. gradually reducing the excess of air unit presence of unburnt fuel is detected in the stack with a CO analyzer or more empirically when smoke appears.

Suppose also that, when the optimal excess of air is individuated, the same be controlled by means of an oxygen percent controller in the stack, utilizing the basic control strategy.

In this case some inconvenient would be faced:

Considering that, for abvious reasons, amission of smoke during the furnace normal operation cannot be admitted, and that an oscillation of the process operation around the optimal point, when negative, would produce it, the operator tends to run the furnace with an excess of air safety margin which will assure that process oscillations, due to external disturbances, never bring the furnace in condition of insufficient combustion.

This safety margin, which determines a significant heat loss, depends on the amplitude of the observed oscillations.

Consequently, the more stable the operation, the smaller the safety margin needed and the higher the furnace efficiency.

The principal disturbances which can affect the furnace operation stability are the possible variations of fuels and air physical characteristics:

- Temperature of fuel oil;

- Pressure, temperature and combustion calorific of fuel gas;

- Temperature and humidity of combustion air;

Possible thermal content variations of the fluid to be heated by the furnace are another source of instability.

Variations of feed flow and temperature determine oscillations of the consequent oscillations.

Usually, in a furnace where two fuels are burnt, one tends to maximize the combustion of the fuel which cost results to be the lower or the one which is the most available.

In this case the combustion of fuel gas results to be more advantageous. Hence, the task of the advanced control strategy will be also, for a given request of combustion heat, to maximize the utilization of fuel gas and minimize consequently the fuel oil one.

When an acceptable level of stability is reached, it will be possible to calculate "on-line" the furnace efficiency and search, with an iterative program, the excess of air value to which the maximal operative efficiency corresponds.

<u>b/ Total furnace advanced control</u>

The more complicated advanced control scheme realizing the total combustion heat advanced control, fuel/air ratio advanced control,fuel gas maximization and feedforward control from furnace feed to combustion is suggested as a second stage for digital control implementation of the hot-oil furnaces in the DMT Plant .This scheme is shown on Fig. 7b in ANNEX 1.

- <u>Total combustion heat advanced control</u>

The advanced control strategy presented in Fig. 7b, allows to compensate all the disturbances determined by fuels, physical characteristics variations.

The general objective is to supply to the furnace a total fuel quantity which can produce a combustion heat corresponding to the one requested by the furnace outlet temperature controller.

In order to achieve it, the oil and gas combustion heat flow are calculated and summed to obtain the total heat flow, then the set points of the oil and gas volumetric flow controllers are corrected to equalize the total supplied combustion heat to the requested one.

The advanced control modules are represented by the rectangles, the basic control ones by the circles.

There are many interactions between basic modules and hierarchical superior modules; therefore the implementation of such strategy is possible only if a microprocessor based system is utilized.

The mosaic is set in such a way that, at any moment, the operator can activate/disactivate the superior modules without provoking process perturbations.

Evident result of this control strategy is that disturbances determined by fuel oil temperature variations and fuel gas temperature, pressure and composition variations are immediately compensated.

- Fuel/air ratio advanced control

The control strategy in Fig.7b allows to compensate the disturbances determined by the combustion air physical characteristics variations.

Its objective is to supply a quantity of combustion air corresponding to the one requested by the stack oxygen percent controller which determines the ratio between the combustion heat and the comburent oxygen flow.

This way to solve the problem is approximate. In effect the excess of air should not be kept in constant ratio with the combustion heat flow, but it should be determined also in function of the composition of the total fuel supplied (carbon/hydrogen ratio)

If sudden fuel composition variations occur, mainly in the fuel gas, this calculation and the consequent compensation could bring additional benefits to the process stabilization.

Otherwise, slow composition variations can be easily compensated in feedback by the stack oxygen percent controller.

In any case the result of this control strategy is to avoid that combustion air temperature, pressure and humudity variations can destabilize the control functions. This frequently occurs during weather perturbations.

- Fuel gas maximization

The control strategy module gradually increments the fuel gas consumption because in this case it results to be cheaper than fuel oil.

The advanced control module gradually increments the gas flow. This increase is compensated by a consequent decrease of the fuel oil flow, because the control strategy described in point a/ above, keeps the combustion heat flow to the requested one. The gradual increase operation ends when the following constraints are encountered:

- General fuel gas header minimum pressure

- Minimum fuel oil flow accepted by the furnace burners.

- Fedforward from furnace feed to combustion

The control strategy of Fig 7b allows to compensate the disturbances determined by the fluid to be heated.

It is evident that feed temperature or flow variations will required correspondent combustion heat flow variations to keep constant the furnaced outlet temperature.

Therefore this strategy acts directly on the fuel demand and increments/decrements it proportionally to the requested heat variation occured.

This avoids violent feedback corrections by the outlet temperature controller, which certainly cause oscillations in the system, and therefore assures a high outlet temperature stability.

- Furnace optimization

After having implemented the advanced control strategies described in the preceding items, and consequently a high level of furnace operational stability has been reached, it is advisable to proceed to calculate the efficiency of the operative unit.

The necessary data are all present in the above described schemes.

The calculation methods depend on the chosen standard: heat loss or input/output efficiency.Some times both calculation are implemented and confirmed before being utilized in an optimization strategy.

In any case, as this value is obtained by means of a relatively simple calculation, with an iterative search program the oxygen percent controller set point is incremented or decremented until the furnace efficiency will be driven to its maximum.

Presented furnace advanced control structure is applicable for the hot-oil furnaces of DMT Plant, after the appropriate technological modernization.

c/ <u>Hot-oil distribution system advanced control</u> .

In the DMT Plant the hot-oil furnaces are transfering heat to the so called "hot-oil" used by several reboilers. A low oil flow requirees a higher reboiler inlet temperature to transfer a given flow of heat to the column.And the higher inlet temperature in turn requires a higher flue-gas temperature in the heater and therefore causes higher stack losses. Maximum efficiency will be realized when oil flow is always maximum and oil temperature is minimum.

The control system should be arranged so that the reboiler demanding the most heat will recieve full oil flow, whith oil temperature set to deliver that heat. Flow to the other reboilers may then be throttled to match their requirements. Such kind of distribution system is presented on Fig. 8. ANNEX 1.

Each column in DMT Plant has its own heat input controls manipulating hot-oil flow. The valve-position signals are compared in the high selector, and the highest is send to the valve-position controller (VPC).This device adjusts oil temperature until the highest valve signal is at or near full opening. Then the oil temperature will be at its minimum acceptable value, as will the hydravlic power loss through the control valves. The valves are free to be manipulated by the individual column controls for fast response in the short term while the slower-acting valve-position controller minimizes energy loss in the long term.

Conventional heater control include a bypass, recirculating hot-oil back to the cold-oil line, to protect against loss in flow through the heater. The bypass valve is manipulated to control pressure in hot-oil line.As less heat is required by the reboilers, the bypass valve opens to maintain a constant flow through the heater. However, bypass hot oil represents a loss in available work since it is blended with the cold oil. In the scheme shown in Fig. 8 ANNEX 1, bypass is not normally full open. However, a bypass valve that fails open should still be used to protect the heater against a control failure.

B/ Suggestions for implementation of distributed digital process control system of DMT Plant.

An implementation of a modern distributed microprocessor based control system in DMT Plant is described in the following section. It presents some consideration on the function capabilities, architecture hardware solution, input/output characteristics, information flows, etc.

1. Functional description

We assume that the main function to be handeled by the automation system in the DNT Plant (independent of the system type- analog, digital or hydride) can be defined as follows:

1.1.
Precise and correct measurements of the process variables such as flow, pressure, temperature, status,position, etc.The field mounted instruments that perform these measurements should be capable of accepting analog and digital variables.

They must be satisfactory for the specific environtment in the DMT Plant processes. It is preferable to have the possibility for multiple outputs from a transmitter, because of reservation, compensation, registration, etc.

Final control elements such as valves, actuators, contactors, etc. should be highly reliable, made of resistant to the process fluids materials.

1.2.
Indication, registration,automatic stabilisation, compensation, interlocking, etc.

A main point is that some process variable should be continuously recordered. Another important point is that the provided interlocking system should be external to the automation control system of the processes.

1.3.
Dispatching of process information in a centralised location namely the control room. The monitoring and control of the process and the equipment should be done easily from the control room or from satelite rooms, where cabinets may be located. Nevertheless intermixing should be avoided in any case, and significant concentration of the process information is required.

1.4. General notes
Obviously the provided automation control system should have such characteristics as low cost and highly reliable

equipment, minimum wiring and cabling, easy operation and maintenance, etc.

2. Architecture

The automation control system of DMT Plant could be divided in some more or less independent system as follows:

2.1. Process controlo system

It is assumed that this subsystem is a digital microprocessor based one with a high degree of reliability. It should posses the following characteristics.

- The units should be distributed throughout the plant, which improves the control system reliability and should be capable of controlling the processes even if the communication with the central control room is lost;

- The control system should provide a centralisation of the operations, i.e. the operators should be able to control and supervise the process parameters from a central control room via CRT based operators stations, even in manual mode;

- The alarm capabilities of the system should be in a position to ensure the indication of prealarm situations. In this way the operators should have the possibility to control the process, avoiding hard emergency events.

- The trending and documentation of the process should be capable of giving a full picture of the process history during current shift and the preceding one;

- The process information (Data Base) should be distributed throughout the plant and should be available at a central control room at any moment;

- A digital information transfer between the process interface units and the central control room should be realized.

Thus, the distributed architecture eliminates the risks associated with a conventional centralized control system implementation. Digital control algorithms and communications ensure high accuracy of the control system.

2.2. Interlocking system

This subsystem should be based on solid state logic. It should be completely independent of all other sub-systems included. Nevertheless, it uses single or double sensors and should be capable of running interlocking sequences at the occurance of an event or on operators request via the control subsystem operator's stations.In the same time, an occurance of a shut-down procedures should be immediately reported to the operators through the CRTs and lighting annunciators, mounted in the central control room.

In this way the necessary redundancy and re'iability of the alarm events handling should be reached.

2.3.Dispatching system

This subsystem solves the problems of the process operator interface. It should consist of CRTs with corresponding keyboards with the possibility of graphic displays indication. It should indicate also some printing devices for documentation of the processes information. Thus it ensures a significant concentration of the process information in the central control room, which is preferable, from the point of view of mopdern process control strategies.

2.4. Back - up system

This subsystem is not strictly separate from all listed above in terms of technical realisation, but it should perform the function of automatic control, operator interface, power supply, information transfer, etc.redundancy.

3. Input/output characteristics of the digital automation system.

These characteristics correspond directly to the scope and size of the system to be implemented, and are strongly related with its hardware technical specifications.

For DMT Plant they are given in Table 4.1.3.1 to 4.1.3.10 for each process stage.

4. Distributed control system configuration

The proposed Honeywell TDC-3000 system configuration fully perform the listed in Table 4.1.3.10 above, total input/output characteristics of the digital control system for DMT Plant. In addition this system configuration decreases significantly the lenght of the control cabling needed and gives unlimited possibilities for preconfiguration (without additional investments) of control strategies (automatic and sequence control, interlocking, alarm handling, etc), following any improvements or changes in the existing technology in future or possible new ideas of control strategies (cascade, feedforward, advanced control, etc)

It consist of multifunctional controllers, distributed throughout the installations, a centralized operator centre in the existing control room, battery back-up power supply system and an uninterruptable automatic control system. Thus it ensures very high degree of reliability and redundancy and remains functioning even in case of a module's malfunctioning. Something more to say is its selfdiagnostic function, which decreases greatly the time needed for repairing, thus increasing the plant productivity.

Considering the high degree of accuracy, reliability and concentration of the operator interface and step by step representing of the process information, it enables the operator staff immediate intervention if necessary yet in prealarm situations. Thus avoiding alarm or emergency events.

The architecture of the system is shown in Fig.4.1.The abbreviations used are as follows:

EOS 1,2,3 - Operator station REL 520 with floppy disc drives
LP 1,2 - Line printing device
ATR - Analog trend recorders
HTD - Hiway traffic director
UAC 1,2 - Uninterrupted automatic control system
RCD 1,2 - Redundant controller director
RMC 1,2 - Reserve multifunctional controller
MC - Multifunctional controller
PCFA - Point cards file assembly
LEPIU - Low energy process interface unit

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 0 - Methanolysis/Catalist Recovery TABLE 4.1.3.1

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	39	8	-					
FLOW	13	10			2			
PRESSURE	12	5				6		
LEVEL	15	8				24		
OTHERS						33	4	2
TOTAL NUMBERS	79	31			2	63	4	2

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 1 - OXIDATION

TABLE 4.1.3.2.

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	89	16						
FLOW	18	8			21	4		
PRESSURE	20	8				15		
LEVEL	25	16				35		
OTHERS	8	6		9		33	14	
TOTAL NUMBERS	160	54		9	21	87	14	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 2 - ESTIRIFICATION TABLE 4.1.3.3.

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	47	7	1					
FLOW	14	10	2					
PRESSURE	14	2				7		
LEVEL	7	5				4		
OTHERS						9	3	
TOTAL NUMBERS	82	24				20	3	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 3- RAW - ESTER DISTILLATION TABLE 4.1.3.4

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	29	3						
FLOW	11	6	1					
PRESSURE	13	5				2		
LEVEL	7	7	1			5		
OTHERS						13	3	
TOTAL NUMBERS	60	21	2			20	3	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 4 - FIRST CRISTALIZATION TABLE 4.1.3.5

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	20	3						
FLOW	6	4	3			8		
PRESSURE	3	2				7		
LEVEL	9					13		
OTHERS	8	4				36	9	4
TOTAL NUMBERS	46	13	3			64	9	4

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 5 - SECTION CRISTALIZATION TABLE 4.1.3.6

TYPE \ FUNCTION	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	21	3						
FLOW	6	2				6		
PRESSURE	8	4				9		
LEVEL	8	3				18		
OTHERS	6	3				47	16	2
TOTAL NUMBERS	49	15				80	16	2

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 6 - MULTIDISTILLATION TABLE 4.1.3.7

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	54	11						
FLOW	20	13	3		2			
PRESSURE	13	5				6		
LEVEL	14	6	1			21		
OTHERS						23	10	
TOTAL NUMBERS	101	35	4		2	50	10	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL .OCESS CONTROL SYSTEM

DMT PLANT - STAGE 7 - FLAKING SECTION TABLE 4.1.3.8

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	2	1						
FLOW	3	1			4			
PRESSURE	3							
LEVEL	1	1				14		
OTHERS	5					28	4	
TOTAL NUMBERS	14	3			4	42	4	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE 8 - VACUUM AND VENT.SYSTEM TABLE 4.1.3.9

TYPE \ FUNCTION	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	17	3						
FLOW	12	6						
PRESSURE	4	2				11		
LEVEL	4	1				7		
OTHERS	5					8	1	
TOTAL NUMBERS	42	12				26	1	

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

DMT PLANT - STAGE TOTAL TABLE 4.1.3.10

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	318	55	1					
FLOW	103	60	9		29	18		
PRESSURE	90	33				63		
LEVEL	90	47	2			141		
OTHERS	32	13		9		230	64	8
TOTAL NUMBERS	633	208	12	9	29	452	64	8

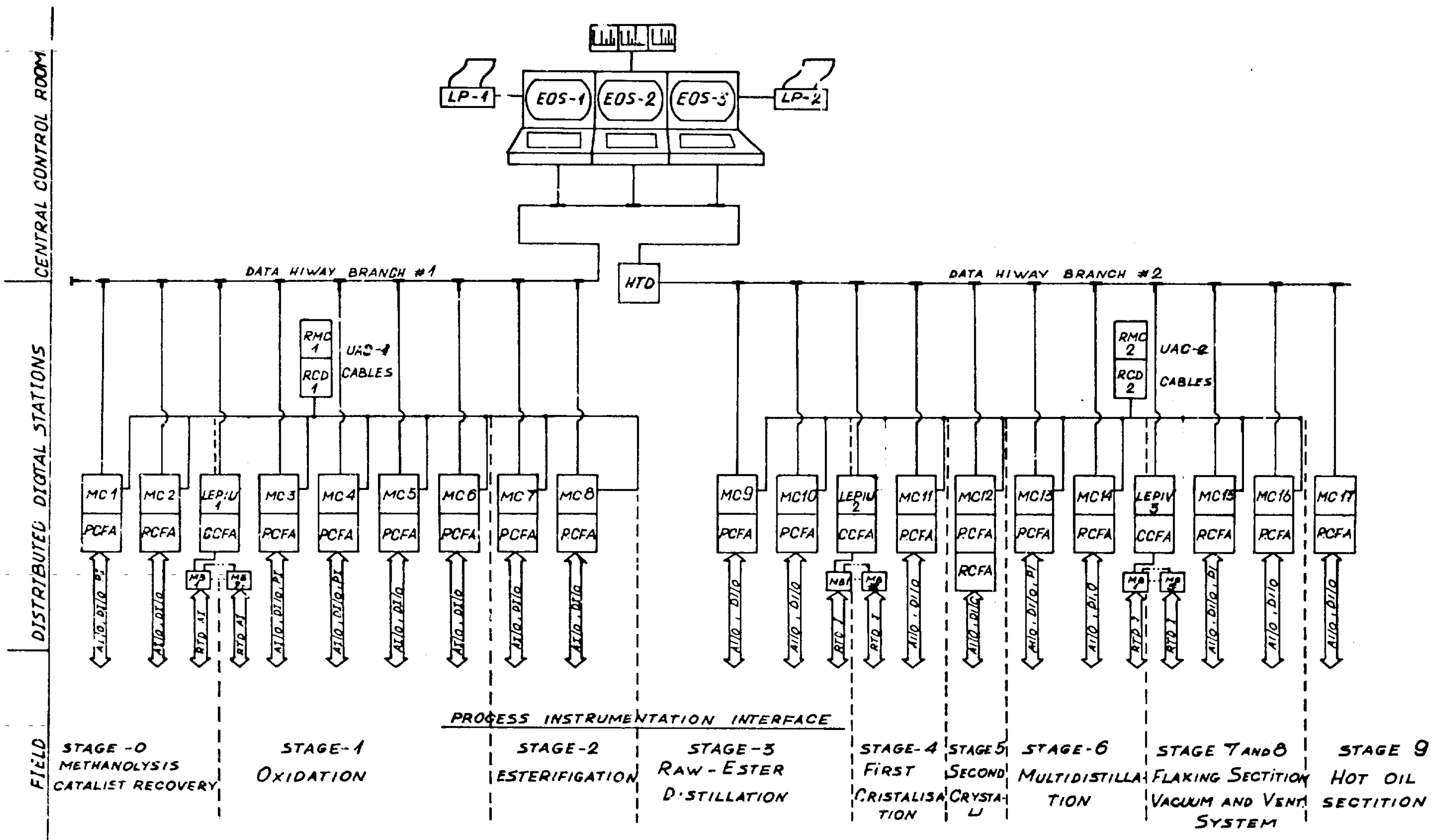


Fig. 4.1 Distributed process control system for DMT plant

CCFA - Common card file assem'ly
MB 1,2,...14. - Multiplexor boxes for RTD signals
A I/O - Analog input/output signals
D I/O - Digital input/output signals
PI - Pulse input signals

The description of the main modules of TDC - 3000 system is attached in ANNEX B.

The first EOS -1 is assigned for alarm handling and alarm printing on LP-1.

The second EOS - 2 is assigned to basic control and monitoring of the process. It is connected to three 3-pen recorders, which are capable for recording for a chosen by the operator period of time every eight process variables.

The third one EOS - 3 is implemented for displaying process mnemoschematics (graphics) with live process data. It is connected to a printer, currently reporting the process status.

Those three operator stations are fully independent and they can have the same data base. In the cases when one of the EOS fails the other two remain in operating mode and fully back-up its functions. The main modules of the suggested system configuration, such as MC's communication system using hiway data trafic and power supply units are fully reserved, ensuring in this very high degree of reliability and availlability of the distributred digital system as a whole.

A detailed specification of the TDC - 3000 system is given in Table 1, ANNEX 1.

C/ Specification of additional field instrumentation

For the implementation of the suggested control strategies only a few additional field instruments are required:

Stage 1 : DMT Oxidation

- a flow transmitter for steam generated in oxidizers' cooling system - 1 per oxidizer;

- a flow transmitters for oxidizers' off-gas - 1 per oxidizer;

- O_2, CO_2 and CO analyzers for oxidizers' off-gas;

Stage 3 : Raw DMT distillation

- a flow and a temperature transmitters for raw-DMT residue feed of the distillation column C-601;

- a flow transmitter for hot-oil in the outlet of the column C-601 reboiler 321.

Stage 9 : Hot-oil section

- a flow and a temperature transmitter for fuel oil- 1 per furnace;

- O_2 analyzers for furnaces stack;

A detailed specification of the above mentioned additional field equipment with preferable suppliers is listed in Table 2,ANNEX 1.

D/ Recommendation for field instrumentation implementation

For implementation of advanced and optimal control strategies, proposed above, it will be necessary to provide some additional field instruments, mainly transmitters. Here we can make a note that all temperature measurements which are connected with the digital system for regulation or are included in the advanced control schemes and strategies are signal conditioned in the special auxiliary cards. These cards are integrated in the specification and in the proposed distributed control system configuration. Therefore no separate temperature converters are required.

Using the existing pneumatic instrumentation with a digital control system is not possible and it will cause relatively low accuracy, repeatability, dynamic responsibility, etc.

Having this in mind we are suggesting to implement electonic transmmiters for most part of the closed control loops and for the main part of parameters for indication with alarm. Detailed suggestions for field instrumentation are given in Tables 4.1.2.1 to 4.1.2.9., ANNEX 1, and the list of preferable instrumentation vendors for petrochemical industry- in Table 3, ANNEX 1.

E/ Possible benefits and savings.

Stage 1 : DMT Oxydation

The analysis of this unit leads to the following conclusions:

a/ a lot of the main process indexes/e.g. acid number, product compositions/ vary in a wide range.

b/ the infrequent and time-intensive analysis introduce considerable time delay in the detection and calculation of process deviations.

c/ there is a considerable dispersion in the different bathes' duration, product composition and througput

It can be expected that the application of the optimal control strategy will result in :

1. Decreasing the batch cycle time with 2.5-3%, that will yield an annual increse in troughput of about 1,5% with the same raw-materials consumption. In such a case the annual benefit is expected to be:

30000 T/YEAR X 0.015 X 950$ /T = 427 500$/YEAR

2. Stabilization of oxidate composition. Hard credits cannot be calculated but this will have clearly a beneficial effect on the whole DMT plant operation.

Stage 2 : Esterification

The application of the proposed optional control strategy is expected to result in stabilization of the raw ester composition thus achieving an increase in the yield of target product of 1.0-1.5% annually.

In this situation the annual net increase of the production can be estimated as about 0.5% DMT per year which gives the following economic benefit per year:

30000 T/YEAR X 0.005 X 950$/T = 142 500 $/YEAR

Stage 3 : Raw ester distillation

The suggested advanced control strategies for this stage will give about 2-6% annually savings in consumption of hot-oil in this process section.

Stage 9 : Hot-oil section

According to the various implementations of advanced control strategies for hot-oil heaters /furnaces/ the annual saving for fuel-gas and fuel-oil can be estimated about 7-15%.

After implementation of the suggested above advanced and optimal control strategies the total annual economical benefits can be estimated at least as 570 000$/year.

These advanced control implementation in distillation and hot-oil stages will give additional hard and soft-credits. Finally , we can assume that the time for repaiment of the investments will be about 3 years.

4.2. ACN Plant

4.2.1.Background information

The acrylnitril production unit is designed on the basis of a process license from SOHIO U.S.A. by the engineering company Badger from Netherland. The licensed production capacity of the plant in MTA is as follows :

- Acrylnitrile - 24000
- Acetonitrile - 900
- Hydrocyanic acid- 3600

The ACN Plant consist of the following production units and process sections:

A/ Acryl unit

- Reactor section
- Recovery section
- Purification section
- Waste treatment section

B/ Acetonitrile purification unit

C/ Hydrocyanic acid purification unit

In the acrylonitrile production uinit the main process section is the reactor section, where the most important vessel is the reactor MR -1001. Gaseous amonia and propulene from the superheaters TT-1027 and TT-1026 feed the bottom part of the reactor. Feed air for the reactor is provided from the compressor PC-1001 in the bottom also.Propylene, amonia and oxygen air in contact with a special catalyst react and as a result reactor effluent containing acrylnitril, acetonitrile, nydrocyanic, water, CO,CO2,nitrogen, oxygen, acetone, etc is obtained. The gas mixture /effluent/ after cooling in the quench column AS-1001 feeds the bottom part of the absorber AS -1003 in the acrylo recovery section. The main purpose of the recovery column AS 1004 is pumped for feeding the upper part of the heads column AS-1006 in the acrylo purification section. This column is used for purification of the acrylnitrile and separation of the hydrocyanic acid to the HCN purification column AS-6001. Purified from the waste water acrylonitrile feeds the product column AS-1007 where fine purification of the acrylo product is obtained.

Acetonitrile purification is made by acetonitrile heads column AS-4001 and batch still column AS-4002.

The existing process instrumentation and control subystem is analog pneumatic type, centralized in the ACN plant control room. The field and panel mounted instrumentation is mainly produced in India and there are only some instruments in the control room which are electronic type. In ACN plant are used two oxygen analyzers, produced by Tailor- model Servomex and two pH analyzers from Foxboro. The control valves used in this plant are produced by Fisher.

The existing pneumatic instrumentation does not allow the application of advanced control strategies. The input/output and functional characteristics of the existing process control system are summarized in Table 4.2.1.

The numbers, shown in brackets are related to the suggested below new advanced control schemes.

In the control room a multipoint themperature indicator for 144 signals is used, also at present.

The existing process units have been studied and some areas where possible benefits can be obtained through implementation of new advanced control schemes and strategies have been identified.

INPUT/OUTPUT CHARACTERISTICS OF THE DIGITAL PROCESS CONTROL SYSTEM

ACN PLANT - TOTAL

TABLE 4.2.1.

FUNCTION / TYPE	ANALOG SIGNALS				PULSE INPUT SIGNALS	DIGITAL SIGNALS		TIMERS
	ANALOG INPUT	CONTROL LOOPS	CASCADES	ADDITIONAL ANALOG OUTPUTS		DIGITAL INPUTS	DIGITAL OUTPUTS	
TEMPERATURE (RTD TYPE)	70/ 12/ ADD./+144/*	14				18		
FLOW	67/+2/	31			10			
PRESSURE	39/+5/	24				13		
LEVEL	67	23				73		
OTHERS	60/+2/	50		6		10		
TOTAL NUMBERS	303/+21/ ADD./+144/*	142		6	10	114		

4.2.2. Recommendations.

A/ Suggestions for implementation of advanced control strategies.

The process and instrumentation diagrams of the suggested new control strategies are given in ANNEX 2 of this report, where the existing basic analog control elements are represented by circle and the advanced controls in rectangles.

Section 1 : Acrylo reactor section.

The main process unit in this section and the most crytical one for the operation of the whole plant from the process control point of view is the reactor MR -1001, where a catalytic reaction of conversion of the propylene , amonia and oxygen from air to acrylonitrile, acetonitrile, HCN,CO_2 and CO takes place.

The main characteristics of the reactor are its conversion degree and selectivity. Thus parameters are mainly influenced by feeding ratio between the propylene, amonia and air, catalyst status, temperature conditions, etc.

The objective of the process control strategies is to maximize the acrylonitrile yield per unit of propylene.

In order to achieve this objective the implementation of advanced control strategy, shown on Fig.1, ANNEX 2, is suggested.

It acts as follows :

- The signals from the field transmitters/analyzers are processed by process variable algorithms like mass flow calculation of the propylene and amonia feeds and oxygen flow calculation.

- The ratio calculation are made on the basis of on-line oxygen analysis rezults and quality calculation. By this advanced scheme it will be possible to realize the optimal ratio of the feeding to the reactor materials.

Like second stage of the advanced control strategy implementation an optimal control of the reactor temperature conditions is suggested. Due to the complexity of the reactor conditions to realize its optimal control, a mathematical model and an optimizer procedure need to be implemented.

The estimator parameters are periodically updated utilizing laboratory and on-line data.

The suggested optimal control strategy on Fig . 1,ANNEX 2 acts as follows:

- Input processing and varification of the signals from the field transmitters, i.e. transmitter limits check, filtering, converting into engineering units, alarm generating and after that storing in a data base.

- The values of the signals are fed in the estimator which includes a mathematical process module. It outputs the estimated values of current catalyst selectivity and conversion degree to the optimizer. The reactor optimizer, which includes the optimization program, constraints and the objective function, calculates and outputs to the controllers the optimal (for the each particular unit) setpoints.

By this approach it will be possible to realize the optimum moving control system and thus realizing the continuous optimal operation of the unit.

<u>Section 2</u> : Acrylo recovery section

The main unit in this section is the absorber column AS-1003 where the cold water absorbs ACN, acetonitril, HCN and the gas containing nytrogen, CO, CO2, etc. are outlet to the atmosphere.

The objective of the suggested advanced control scheme shown in Fig.2, ANNEX 2, is to stabilize the operation of the absorber by implementing the following advancerd control elements:

- Mass flow calculation of feed;

- Internal reflux control for both water streams;

- Dynamic feedforward compensation from feed to the bottom heat input;

- Dynamic feedforward from feed to internal reflux controllers.

The principle of action of these advanced control elements have been described in detail in the section 4.1. of the present report.

<u>Section 3</u> : Acrylo purification section.

The main process units in this section are the acrylo heads column AS-1006 and the acrylo product column AS-1007. The objective of the advanced control strategy, shown on the

Fig.3, ANNEX 2 ,for the heads column AS -1006 is to stabilize the operation of the column and to minimize the steam consumption.

The purpose of proposed control scheme is the reboiler heat duty advanced control by implementation of the following elements:

- Heat flow calculation and control of the steam inlet to the column reboilers TT-1016 A/S.

- Mass flow calculation and control of the crude acrylo feed to the column AS -1006;

- Dynamic feedforward compensation from column feed to the reboilers heat input;

- Dynamic correction of the bottoms temperature controller set point to compensate for tower/column/ pressure variations.

On Fig.4,ANNEX 2 the suggested advanced control strategy for acrylo product column AS-1007 is shown.The main objectives of this strategy are reflux advanced control, reboiler heat duty advanced control and total column advanced control by application of the following advanced control elements:

- Internal reflux calculation and control;

- Feedforward control of the internal reflux from the column feed;

- Mass flow calculation and control of the column feed;

- Heat flow calculation and control of the steam to the reboiler TT-1019;

- Dynamic feedforward compensation from column feed to reboiler heat input;

- Dynamic compensation from column bottoms temperature controller to the reboiler heat input;

- Internal decoupling in feedforward, which reads the variations of the internal reflux of the column and automatically compensates the internal flux of liquid/vapour by action on the heat flow to the column AS-1007;

- Mass flow calculation of the acrylo product flow to the MT-1021/ the outlet product flow/.

Section 4 : Acrylo waste treatement section.

Advanced control strategy not identified

Section 5 : Acetonitrile purification unit.

This unit is not of the main production importance and advanced control strategy is not identified.

Section 6 : HCN purification unit.

This unit is very crytical from environmental protection point of view and the main process unit is the HCN purification column AS -6001. The advanced control strategy for this column is shown on Fig.5 in ANNEX 2.

The main objective of the proposed advanced control is to stabilize the operation of the purification column and to minimize the steam consumption for it.

This objective can be fulfiled by implementation of the following advanced control elements:

- Internal reflux calculation and contr

- Mass flow calculation and control of the feed flow to the column AS-6001.

- Heat flow calculation and control of the steam flow to the column reboiler TT-6002;

- Dynamic feedforward compensation from column feed to reboiler heat input.

- Interactive decoupling in feedforward from the heat flow to the column to the internal reflux flow.

B/ Specifications of a Distibuted digital automation system.

The specification of the TDC-3000 digital automation system, which functional and technical characteristics are described in APPENDIX B of the present report is listed in Table 2.1, annex 2. This specification of the system modules is necessary for the implementation of the functions, listed in Table 4.2.1, above in the present section as well as those of the advanced and optimal control, proposed in point A/

The configuration of the proposed system is shown on Fig. 6 in ANNEX 2.

The connection of the 144 temperature measurements, which now are used on a multipoint temperature indicator in the existing control room, with the proposed digital autmation system is suggested like an option.

The suggested optimal and advanced control functions can be implewmented using the sequence control capabilities and SOPL programming larguage of the multifunctional controllers /MC/.

C/ Specification of additional field instruments.

For the application of the proposed advanced control strategied only a few additional field measuring instruments /shown on Fig.1 to Fig.5, ANNEX 2 with "new"/ are required:

Section 1 :

- a temperature and a pressure transmitters for propylene, amonia and air feeds to the reactor MR-1001;

- a CO and a CO2 analyzers for the acrylo effluent output from the column AS-1001.

Section 2 :

- a flow and a temperature transmitters for feed to the absorber AS-1003.

Section 3 :

- a temperature transmitters for crude acrylo feed to the heads column AS-1006,

- a pressure and a temperature transmitters for steam to the reboiler TT-1016 A/S.

- temperature transmitters for the feed and reflux flow of the product column AS-1007;

- a pressure and a temperature transmitters for steam to the reboiler TT-1019

Section 6 :

- a temperature transmitter for feed HCN flow to the column AS-6001;

- a temperature and a flow transmitter for HCN column reflux;

- a temperature transmitter for steam flow to the reboiler TT-6002;

- a control valve for reflux flow of the column AS-6001;

D/ Recommendation for the field instrumentation

To interface the suggested digital control system to the field instruments two approaches can be suggested:

1. Revamp all the existing pneumatic instruments with new electronic ones;

2. Install P/I converters (and I/P converters for output signals to the final control elements/valves/)

The first approach seems to be superior from technical point of view. At present time the available on the market electronic transmitters, especially the new so-called "smart" ones, have much better accuracy (up to 0-1% of the full scale) no drift , better temperature and pressure compensation, higher reliability, much easier maintenance and they can be directly interfaced, without a necessity of converter, to the digital process control system.

For this ACN plant the second approach is also applicable , because the existing pneumatic transmitters are in relatively good condition. This solution will be a cheaper one.

The existing analyzers, thermocouples and swithes can be directly interfaced to the digital system using the signal conditioning cards.

Like a best solution we suggest to implement the combination between the first and second approaches. In this case all parameters fore closed loop control and for the implementation of the suggested above advanced and optimal control strategies is recommended to be with electronic transmitters. The other part of the process parameters which are used only for indication/with or without alarm/ can be interfaced via P/I converters.

For the loops equiped with electronic transmitters we recommend the use of Honeywell services ST-3000 Smart transmitters for pressure, differential pressure and level measurements. The description of these transmitters is given in APPENDIX A.

E/ Possible benefits and savings.

Section 1 : Acrylo reactor section.

1. Due to catalyst aging, i.e. decreasing of catalyst activity it is required to change the operating conditions for composition usually once per 20-25 days.

The duration of the transition period can be prolonged due to improper actions of the operators. From the data collected on-site/at the ACN-BARODA Plant/ the hourly production of acrylonitrile at steady state is around 3t.

On the base of the implementation of the suggested optimal control strategy in p.A/ above we can assume, that the transitional periods during which the conversion degree of propylene to acrylonitrile is reduced, can be considerably reduced.

This reduction can be estimated with greater accuracy if an economic feasibility study is prepared. From our experience and other case studies we can pressume that at least 50% of the time can be saved for normal operation of the reactor section. This will result an increasing of the mounthly acrylonitrily production of about 42 tons.The annual benefits will be about:

42t x 12 (mounths) x 720 $/t = 362 880 $/year.

2. The main characteristics of the catalyst , strongly influencing the process, is its selectivity.

Selectivity is the ratio of moles produced acrylonitrile to the moles reacted propylene.Because the selectivity depends not only on the nature of catalyst, but also on working conditions, mainly on temperature, it is possible to maximize the selectivity using the optimal control strategy and thus to encrease the annual production of acrylonitrile with about 1-1.5%.This will give the annual benefits of about:

360 t/ACN X 720 ($/T) = 259 200 $/YEAR.

Total expected annual benefits for the acryl reactor section after the implementation of the suggested optimal control strategy on the basis of digital control system will be about 586 080$/YEAR

Section 1 : - Acrylo recovery section.

Section 3 : - Acrylo purification section

Section 6 : - HCN purification section

For all these section advanced control strategies proposed in p. A/ are concerning the process advanced control of distillation columns and they are aimed for stabilizing the operation of the columns through stabilizing the internal flow of liquid and vapour. Thus it can be obtained a better product quality control and minimizing the steam consumption.

The heat input to each column bottom /reboiler/ is controlled by a series of modules, which from a feedforward loop from mass flow of column feed to a heat flow controller, manipulating the set-point of the steam flow controller. Proposed control structure for each column is intended to maintain a constant ratio between feed flow and heat input flow, thus realizing stabilization of the internal flow by keeping also a constant temperature in the bottom part of the column.

Applying this approach the control strategy is better than the existing,basic control structure, because the real heat transfer is controlled and it is compensated dynamically for variations in feed flow.

Implementing this advanced control scheme typical savings of 5 to 8 % of steam can be achieved, or annually about:

151 200mT/YEAR X 0,05 = 7 560mT X 10 $/T = 75 600 $/ YEAR.

The suggested control structure stabilize the internal reflux and product quality, however no hard savings can be estimated.

Summary of annual expected savings in US Dollars.

Section 1 :	586 080
Section 2 :	75 600
Section 3 :	
Section 6 :	
TOTAL :	661 680 $

F/ Options:

From the analysis of the control system and the possible savings expected it is clear that the most important technological unit in ACN Plant from process point of view is the ACN reactor unit.

Like an alternative we would suggest to start the revamping of the existing pneumatic control system from implementation of the optimal control strategy suggested for this unit whith changing the instrumentaion with electonic one. In this case the suggested TDC -3000 system configuration will be as follows:

1.	Multifunctional controller package, including electronic package system firmware application data base terminal panel	1 p
2.	Point card files assembly including I/O cards and terminal panels	2 p
3.	Cabinet with power supply and battery back-up unit	1 p
4.	Local Batch operator station /LBOS/ including electronic package monochrome monitor functional keyboard printer	1 p
5.	C - link redundant communication system	1

TOTAL PRICE IN U.S.DOLLARS ABOUT 120 000$

Like secong stage implementation of Uninterrupted automatic control system /UAC/ - 1 p . for about 60 000 U.S.Dollars is also recommended.

In this case it is clear that the return of the investments is expected to be less than 6 mounths.

Further justification arouses from fact that a pessimistic savings estimate was taken as a basis for evaluation of the economic benefits.

4.3.ETHYLENE GLYCOL Plant

The existing units have been examined and some areas where possible benefits can be obtained by implementing advanced control strategies have been identified.

4.3.1. Background information

(a) Analysis of the existing automation system.

It consists of mainly pneumatic instruments , manufactured in India under Taylor license, and a few electronic ones. The temperatures for indication are measured by thermocouples, which are conected to electronic temperature indicators and recorders.These are also some analytical instruments including:

- Oxygen analyzers;
- IR analyzers;
- Gas chromatographs;

Most of the control loops are centralized in a control room. The instruments are in a good condition. However, the pneumatic instrumentation due to its inherent limitation does not allow an application of advanced control.The information from the gas chromatographe is not suitable to be included in the control loops because the existing chromatographs have not dedicated outputs proportional to the concentration of each component.

(b) Input/output and functional characteristics.

The input/output and functional characteristics are as follows:

Analog inputs, high level (It is assumed that a chromatograph control unit will be implemented)	112
Analog inputs, low level	55
Digital output	100
Analog output	100
Basic control loops	100

4.3.2.Recommendations

(a) Suggestion for application of advanced control systems.

The description of the proposed advanced control strategies is given below and the PID - in ANNEX 3, where the existing basic controls are represented by rings and the advanced comtrol in rectangles.

<u>SECTION 1</u>: Ethylene oxide reaction and scrubbing. The main unit in this section, and also in the whole plant, from the control point of view is the sector D-110, where the oxidation of ethylene and oxygen to ethylene oxide takes place along with a side reaction to carbon dioxide and water.

The main characteristics of the reactor are its selectionity and conversion, which are influenced by catalist status, ethylene, oxygen, carbon dioxide concentrations, presence of catalyst poisons, temperature, inhibitor concentration, etc.

<u>Objective</u>: To maximize the ethylene oxide yield per unit of ethylene

In order to achieve this objective it is not enough to implement only advanced control scheme but due to the complexity of the relations of selectionity and conversion degree with the aforementioned parameters, a mathematical model and an optimizer procedure need to be implemented.

The elements of the proposed advanced and optimal control strategy are as follows (Fig. 1 in ANNEX 3):

-Mass flow control of oxygen feed in ratio with the mass flow of ethylene feed in order to ensure tight ratio of these main reactions.

-Optimal control of reactor temperature and inhibitor addition based on estimated current selectionity and conversion degree.

The estimator parameters are periodically updated utilizing laboratory and on-line data.

<u>Section 2</u>: Carbon dioxide removal system Advanced control strategy not identified.

<u>Section 3</u>: ETO stripping and reabsorption system. Advanced control strategy not identified.

<u>Section 4</u>: Ethylene oxide purification. The main unit is the posification column D410.

Objective: To stabilize the operation of the column and minimize . kg/cm2 steam consumption

Advanced control elements:

- Internal seflux control;

- Heat flow calculation and control of stripping system

- Mass flow calculation of feed;
- Dynamic feedforward compensation from feed to bottoms heat input;

- Product quality control by temperature trim of distillate flow controller.

Section 5: Glycol reaction and evaporation.

Objective: To minimize 12 kg/cm2 steam consumption in the evaporation train.

Advanced control elements:

- Heat flow calculation and control of primary steam input;

- Dynamic feedforward compensation from feed corrected for density.

Section 6: To stabilize the operation of D610 and D 620 columns, tighten the quality control of MEG product and minimizing steam consumption in reboilers.

Elements of advanced control:

- Mass flow calculation and control of the feed of D610.

- Heat flow calculation and control of steam input to reboilers E 610 and E 620;

- Dynamic feedforward compensation from feed mass flow to input heat flow controller of D 610;

- Dynamic feedforward compensation from feed flow and reboiler recirculation flow to input steam flow to reboiler of D 620.

- Temperature trim of reflux controller.

(b) Specification of a DCS .

The specification of the modules of a TDC 3000 system, which is described in ANNEX B of the present report, necessary for the implementation of the functions, listed in Table 4.3.2 optimal control,proposed in point (a) , is given in Table 1, ANNEX 3. The system configuration is given on Fig 6 in ANNEX 3.The proposed optimal and advanced control functions, will be implemented in the sequence control section of the multifunctional controller by use of SOPL programming language.

(c) Specifications of additional field instruments.

For the implementation of the proposed advanced control strategies only a few additional field measuring instruments are required as follows:

Section 5: A temperature and a pressure transmitter for steam to E 531;

A density analyzer for the feed of D 531;

Section 6: A pressure transmitter for steam to E 610;

A temperature transmitter for the feed of D610

A flow transmitter for recirculation bottoms of D 620;

A temperature and a pressure transmitter for steam to E 620;

(d) Recirculation for the field instrumentation.

In order to interface the digital control system to the field measuring instruments two approach can be feasible:

1. Reavmp all the existing pneumatic instruments with electronic ones;

2. Install P/I converters;

From technical point of view the first approach is superior, because the electronic transmitters, especially the new so-called "smart" ones, have much better accuracy(up to 0-1% of full scale), no drift, better temperature and pressure compensation, higher reliability , easier maintenance and they interface directly with the digital control system (without a converter).

In this plant the second approach is also feasible, because the pneumatic transmitters are in a good condition and it would be cheaper one.

If the first approach is addopted we recommend to use Honeywell series ST 3000 smart transmitter for pressure, differential pressure and level measurement.

A description of these transmitters is given in APPENDIX A.

The existing analyzers thermocouples and switches can be interfaced directly to the digital system using signal conditioning cards.

Interfacing the existing process chromatographs present a problem , because their output is in the form of chromatograms, therefore quite a lot of signals must be interconnected between the computer system and the chromatograph and also a lot of calculations for determining the composition of each component from the chromatograms.

The preferrable is to implement a dedicated control unit which will perform the aforementioned function for both chromatographs, to control them and to interface with the digital control system by providing a dedicated sample and hold circuit for each component is analyzed , a simple and hold circuit stores the value of that component. The output of each circuit is a 4 to 20 mA signal, that is connected to an analog input in the control system. A nunmber of such unit are available on the market.

(e) Possible benefits and savings.

Section 1: Ethylene oxide reaction and scrubbing.

1/ Due to catalyst aging /i.e.decrease of catalyst activity/ it is required to change the operating conditions/ for compensation usually once a month.During the transition period, which is from 40 to 60 hours,off spec products are produced. The duration of the transition period can be prolonged due to improper actions of the operators. From the data collected on-site the hourly production of ethylene oxide at steady state is around 2t, and ethylene consumption is around 1.8t.

The suggested optimal control strategy acts as follows:

- The signals from field transmitters/analysers are processed by process variable algorithms(transmitter limits check, filtering, resification, converting to engineering units, alarm generation) and stored in a data base;

- The values of the signals are fed in the estimator/which includes a mathematical process model/It outputs the estimated values of current catalyst selectivity and conversion degree to the optimizer/includes the optimization program, constraints and objective function/The optimizer calculates and outputs to the regulatory controllers the optimal /for this particular unit/ setpoints. In this way it is made possible for the control system to track the moving optimum thus realizing continuous optimal operation of the unit and eliminating or considerably reducing the transitional periods, during which of spec products are produced.The extent to which these periods could be reduced can be estimated with greater accuracy if an economic feasibility study is done, but from our experience and other case studies we can presume that at least 60% of the time can be saved for normal operation , resulting in increased monthly etnylene oxide production of 48t. The amount benefit will be:

48t x 12 (month) x 180($/t) = 103680$/y

2/ The main characteristics of the catalyst strongly influenced the process, is its selectivity.Along with the main reaction of oxidizing ethylene to ethylene oxide a side reaction of full oxidation of ethylene to water and carbon dioxide also taken place.Selectivity is the ratio of moles ethylene oxide to moles ethylene.As the selectivity depends not only on the nature of catalyst but also on working conditions, partuculary temperature, it is feasible to maximize the selectivity using the optimal control strategy thus saving ethylene- about 2-3% from our experience, or annually:

1.8t/h etylene x 8000 h/year x 0.02 x $100 = 28 800$/y

Total expected saving for Section 1: 132480 US$/y.

<u>Section 2</u>: Carbon dioxide removal system savings not identified.

<u>Section 3</u>: ETO stripping and reabsorption savings not identified.

<u>Section 4</u>: Ethylene oxide purification.

The advanced control strategy of the purification column D 410 is aimed at stabilizing the internal flow of liquid and vapour thus tightening product quality control and minimizing steam consumption.

The heat input to the column bottom is controlled by a series of modules, which form a feedforward loop from mass flow of column feed to a heat flow controller, manipulating the set point of steam flow controller. This structure is intended to maintain a contstant ratio between feed flow and heat input flow, thus ensuring stabilization of the internal flow by also keeping a constant temperature in the bottom of the column. This approach is better than the existing basic control structure , because the real heat transfer is controlled and it is compensated dynamically for variations in feed flow.Using it typical savings of 5-10% of steam can be achieved, or annually:

1520t/h x 8000h/y x $10 x 0.05 = 6080$/y

The control structure also stabilizes the internal reflux and product quality, however no hard savings can be estimated.

Total expected savings for Section 4 : 6080 $/y.

<u>Section 5</u>:Glycol reaction and evaporation.

The advanced control structure minimizes 12 kg/cm2 steam consumption in the evaporation train utilizing a feedforward dynamic compensation from I effect evaporator D531. feed flow (adjusted for its density) to heat flow controller, resetting the set point of the steam flow controller.

Thus due to the feedforward signal the control system reacts quickly to change both in feed flow and density with an accuracy within +/- 2%,m necessiting a feedback signal from TRC 530.

The savings from implementing this advanced control structure, apart from stabilization of product quality, that is water concentration can be calculated from an estimated 3-5% decrease in steam consumption, or annually:

6t/h x 8000h/y x 0.03 x $12 = 17280$/y

<u>Section 6</u>: Glycol drying and refining

The advanced control structure comprises:

1. Advanced control of heat input to reboilers of D 610 and D 620.

The heat input of the columns is controlled similary to that of D 410, already described above.

Savings in steam consumption in the 5 to 10% range can be expected yielding:

(1.26+2.85)t/h x 8000h/y x 0.05 x 12$/t = 21456$/y total savings for both columns.

2.Implementation of a temperature controller in D 620 and a distillate flow controller for tightening product quality control, which will stabilize MEG quality. However hard savings cannot be estimated. Total expected savings for Section 6: 21460 US$/y

Summary of annual expected savings in US dollars

Section 1 :	132480
Section 2 :	not identified
Section 3 :	not identified
Section 4 :	6080
Section 5 :	17280
Section 6 :	21460
Total:	177300

(f) Options

It is clear that most of the savings are expected to come from ethylene oxide reactor optimal control, so the alternative to total plant control system revamping is to revamp the instrumentation only of the ethylene oxide reactor unit. In this case the necessary TDC 3000 system configuration will be:

MC package - 1
Including:
electronic package
system firmware
data base
terminal panel

I/O point card files - 2
Including:
I/O cards
terminal panel
cabinet with power supply
and battery back-up unit.

Local batch operating station - 1
Including:
electronic package
monochrome monitor
functional keyboard
printer

C-link redundant communication(UAC) system - 1

Total: 120000 US dollars

Uninterruptable automatic control system - 1

(Optional) 60000 US dollars

A control unit with dedicated analog outputs for each component's concentration will also be necessary for the chromatographs, its price beeing about 40000US dollars

In this case it is clear that without the UAC option the return of investment is expected to be less than a year and a half(having in mind the price of P/I converters) and with UAC option-less than 2 years and therefore the implementation of the suggested control strategy is economically justified for the ethylene oxide reactor unit.

Further justification aroses from the fact that a pessimistic savings estimate was taken as a basic due to insufficient time for a economic feasibility study and therefore it is reasonable to expect greater savings.

4.4. LAB Plant

4.4.1. Background information.

The existing control automation system is centralized analog one.It consists of pneumatic instruments(field and control room mounted) , which are produced in India.There are also some electronic and very few one channel digital units, mounted in the control room (Fisher & Porter, Foxboro and Honeywell) The control valves applied in this plant are produced by Gulde. There are implemented two moisture analysers (Du Pont and Parametric), two pH analysers (Foxboro) and one dielectric constant analyser (Druxel - Brook) also.

The condition of the existing control system is good. Nevertheless it has been used from 1978 , so one can expect maintenance problems in near future.Another point is operator personnel's difficulties due to a large space of panels in control room, especially when some installation is going to be started,or in alarm situations.

Obviously it is not possible to apply any advanced control schemes or modern strategies on the basis of pneumatic instruments,due to such constraints as signal legs,limited lines lenght, impossibility to create some functional blocks, tuning parameters ,interrelations, etc.

The scope of the system, or so called input/output functional characteristics, is listed below:

- Analog inputs - 341 pes
- Analog outputs - 180 pes
- Digital inputs - 120 pes
- Total parameters - 641 pes
- Closed control loops - 180 pes

Note: In total parameters quantity the interlocking system signal are not included.

4.4.2.Recommendations

4.4.2.1.Advanced control implementation.

The existing units in LAB plant have been reviewed and some areas where possible benefits can be obtained by implementing advanced control strategies have been indentified as follows:

A. Heat transfer advanced control.

In controlling a distillation tower, the major problem is to stabilized the internal flow of liquid and vapour.This is achieved by regulating the reflux valve and the heat input valve.

The reboiler heat input is controlled by the series of modules, which form a feedforward loop from the mass flow of the column feed.The structure is intended to maintain a constant ratio between feed flow and rebioler heat duty, thus ensuring stabilization of the internal flow by keeping a constant temperature in the bottom section of the column.

Heat transfer can be calculated directly by using temperature and hot oil flow measurements.This function is performed by the modules of the hot oil flow control loop.This approach is much better than simple hot oil flow stabilization because of the real heat transfer quantity control.

The corresponding control system structure is shown on Fig.1, ANNEX 4.

This scheme is applied for the following aparatus:

Hydrogen process unit

- stripper column H-V9
- stabilizer column H-V11

Molex unit

- desorbent splitter column Mo-V6
- extract column Mo-V9

Pacol unit

- stripper column Pa-V2

Alkylation unit

- benzene column A-V9

- parafin column A-V11

B/ Internal reflux advanced control

The aim is to stabilize the internal reflux which depends not only on the external reflux but also on the subcooling of the condensate on the return plate.Flactuations in the condensate termometer will cause variations in vapour condensation at the top of the column therefore the net flow of vapour out of the column will change.

The suggested control system is shown on Fig.2,ANNEX 4.

The heat duty advanced control is arranged in already described manner.

If the heat duty set point changes while the feed remains constant (e.g. to change bottom product quality) the module automatically compensates the reflux set point.The objective is to balance the internal flow liquid/vapour to keep a stable column profile.

The next module is a feedforward loop which maintains the internal reflux in a constant ratio with the feed.The column mass feed is calculated and ratioed to the reflux with proper lead/lag dead time compensation to account for tower dynamics.

This scheme is applicable for the following aparatus:

Molex unit

- Raffinate column Mo-V5

- Product splitter column Mo-V16

Observing the product splitter column Mo-V16 note that the liquid circuit between column base and the reboiler forms a "V" tube.Liquid contained in a "V" tube has a capability of resonant oscillation.A restriction in the liquid line sometimes can correct this problem by providing damping.

C/.Rerun column A-V12 advanced control.

The proposed structure for rerun column control is shown on Fig.3,ANNEX 4.

The temperature in the bottom part of the column is stabilized by the heat flow controller,arranged in already described manner.It's cascaded with bottom level controller thus ensuring the necessary heat for evaporation when the feed flow is constant.To eliminate the load disturbance in case of flow changes, a feedforward signal is added to level controller output and the result is used as a set point of the heat flow controller.

The same signal which is actually the column mass feed calculated, is ratioed to the reflux with proper compensation for column dynamics.

By this way the column mode is stabilized, ensurung relatively constant product quality for a minimum heat required.

D/ Combustion advanced control

The principal disturbances which can affect the charge heaters operation stability are the possible variations of fuels and air physical characteristics:

- Temperature of fuel oil

- Pressure, temperature and combustion calorific power of fuel gas.

- Temperature and pressure of combustion air.

The air flow usually is forced in the heater in proportion to the fuel. Otherwise it is impossible to achieve satisfactory combustion control efficiency. Therefore this can be the main recommendation for the charge heaters and hot oil heaters in LAB plant.

Possible thermal content variations of the fluid to be heated by the furnace are another source of instability.

Variations of feed flow and temperature determine oscillations of the outlet temperature.

Therefore the objective of the advanced control strategies will be to compensate the above described disturbances in order to reduce the consequent oscillations.

Usually, in a furnace where two fuels are burnt, one tends to maximize the combustion of the fuel which cost results to be the lower or the one which is the most available.

In this case the combustion of fuel gas results to be more advantageous .Hence, the task of the advanced control strategy is also, for a given request of combustion heat, to maximize the utilization of fuel gas and minimize consequently the fuel oil one.

When an acceptable level of stability is reached, it will be possible to calculate "on-line" the furnace efficiency and search, with an iterative program, the excess of air value to which the maximal operative efficiency corresponds.

The general objective is to supply to the furnace a total fuel quantity which can produce a combustion heat correspondent to the one requested by the furnace outlet temperature controller.

In order to achieve it, the oil and gas combustion heat flow are calculated and summed to obtain the total heat flow, then the set points of the oil and gas volumetric flow controllers are corrected to equalize the total supplied combustion heat to the requested one.

The advanced control modules are represented by the rectangles, the basic control ones by circles.

There are many interactions between basic modules and hierarchical superior modules; therefore the implementation of such strategy is possible only if a microprocessor based system is utilized.

The mosaic is set in such a way that , at any moment, the operator can activate/disactivate the superior modules without provoking process perturbations.

The control strategy in Fig 4,ANNEX 4 allows to compensate the disturbances determined dy the combustion air physical characteristics variations.

Its objective is to supply a quantity of combustion air corresponding to the one requested by the stack oxygen percent controller which determines the ratio between the combustion heat flow and the comburent oxygen flow.

This way to solve the problem is approximate.In effect the excess of air should not be kept in constant ratio with the combustion heat flow, but it should be determined also in function of the composition of the total fuel supplied.

If sudden fuel composition variations occur, mainly in the fuel gas, this calculation and the consequent compensation could bring additional benefits to the process stabilization.

Otherwise, slow composition variations can be easily compensated in feedback by the [illegible] oxygen percent controller.

In any case the result of this control strategy is to avoid that combustion air temperature and pressure variations can destabilize the control functions. This frequently occurs during weather perturbations.

This scheme also allows to maximize the fuel gas consumption because in this case it results to be cheaper than fuel oil.

The advanced control module gradually increments the gas flow.This increase is compensate by a consequent decrease of the fuel oil flow, because the control strategy described keeps the combustion heat flow equal to the requested one.

The gradual increase operation ends when the following constraints are encountered:

- General fuel gas header minimum pressure

- Minimum fuel oil flow accepted by the furnace burners.

The structure described allows to compensate the disturbances determined by the fluid to be heated.

It is evident that feed temperature or flow variations will require correspondent combustion heat flow variations to keep constant the furnace outlet temperature.

Therefore this strategy acts directly on the fuel demand and increments/decrements it proportionally to the requested heat variation occured.

This avoids violent feedback correction by the outlet temperature controller , which certainly cause oscillation in the system, and therefore assures a higher outlet temperature stability.

After having implemented the advanced control strategies, and consequently a high level of furnace operational stability has been reached , it is advisable to proceed to calculate the efficiency of the operative unit.

The necessary data are all present in the above described schemes.

The calculation methods depend on the chosen standard: heat loss or input/output efficiency.Sometimes both calculation are implemented and confronted before being utilized in an optimization strategy.

In any case, as this values is obtained by means of a relatively simple calculation, with an iterative search program the oxygen percent controller set point is incremented or decremented until the furnace efficiency will be driven to its maximum.

This structure is applicable for the following apparatus, after the corresponding technological changes:

Hydrogen process unit

- Charge heater H-H1

Pacol unit

- Charge heater Pa-H1

Hot oil section

- Hot oil heater H-H2

E/.Internal coils passes outlet temperature equalization.

The advanced control strategy of Fig.5,ANNEX4,allows to aviod the unbalancement of the outlet temperatures of several furnaces internal coils.It is demonstrated that if the fluids at different temperature are mixed in order to reach a final predefined temperature the combustion heat transfer of the furnace will be lower.

Furthermore the highest temperature coil is subjected to a higher risk of reaching a level of temperature which would damage it.

In petrochemical furnaces the yield depends on the reached temperature level, hence there is a tendency to operate the unit at a temperature value very closed to the maximum admittable.

Consequently, the better the parallel coils temperature equalization, the higher the final temperature reachable at the merging point, and the higher the yield.

The pass balance algorithm reads the outlet temperature of the coils and acts on the single coils' flow controllers to equalize them.

An increase of flow in one pass determines a decrease of its outlet temperature.

The algorithms must also takes in account that every flow variation in one pass is reflected in opposite manner on the others, that the total furnace feed flow must remain constant and the flow distribution on the various passes has to occur with such control valves' opening that the pressure drop is minimized (at least one valve completely open)

Maximum coils' skin temperature and minimum single pass flows limits must also condition the execution of such algorithm.

F/. Hot oil distribution advanced control.

As it was shown in heat transfer advanced control structures a given rate of heat transfer into a fluid may be achieved by an infinite number of sets of flow and temperature rise (or fall) But the minimum loss in available work is coincident with the minimum temperature difference and therefore the maximum flow. Higher flow rates are also helpful in reducting resistance to heat transfer and fouling of surfaces. An optimim flow exists where the improvements are balanced by increased pumping cost.

In LAB plant , hot oil and fuel oil section, the hot oil fired heater is transfering heat to the oil, used by several reboilers. A low oil flow requires a higher reboiler inlet temperature to transfer a given flow of heat to the column.And the higher inlet temperature in turn requires a higher flue-gas temperature in the heater and tehrefore causes higher stack losses.Maximum efficiency will be realized when oil flow is always maximum and oil temperature is minimum.

The control system should be arranged so that the reboiler demanding the most heat will receive full oil flow, with oil temperature set to deliver that heat.Flow to the other reboilers may then be throttled to match their requirements. Such a system is shown on Fig6,ANNEX 4.Each column has its own heat input controls manipulating hot oil flow.The valve-position signals are compared in the higher selector, and the highest is sent to the valve-position controller. This device adjusts oil temperature until the highest valve signal is at or near full opening.Then the oil temperature will be at its minimum acceptable value, as will the hydraulic power loss through the control valves. The valves are free to be manipulated by the individual column controls for fast response in the short term while the slower-acting valve- position controller minimizes energy loss in the long term.

Conventional heater controls include a bypass, recirculation hot oil back to the cold-oil line , to protect against loss in flow through the heater.The bypass valve is manipulated to control pressure in hot-oil line. As less heater is required by the reboilers, the bypass valve opens to maintain a constant flow through the heater. However bypass hot oil represents a loss in available work since it is blended with the cold oil. In the schemes shown in Fig 6 bypassing is not normally required since one of the lead control valves is always nearly full open. However , a bypass valve that fails open should still be used to protect the heater against a control failure.

4.4.2.2. Digital system implementation.

The appropriate microprocessor based distributed control system TDC 3000 is described in APPENDIX B. The corresponding configuration for LAB plant , which is able to perform not only the function specified in Item 4.4.1., but also these related to the advanced control strategy, optimization and necessary software is presented in Table 1, ANNEX 4.The structure of the system suggested is shown on Fig 7, ANNEX 4. The operator center, which is assumed to be located in the existing control room, consists of three CRT operators stations with the corresponding keyboards, arranged as follows:

The first one is assigned to alarm handling. It reports immediately the current status of the alarms and is connected to a printer for alarm documentation.

The second station is assigned to basic control and monitoring of the process. It is connected to three 3-pen analog recorders, which are capable of recording continuously for a period of time every eight process variables, chosen by the operator.

The third one is provided for displaying process graphics with live data. It is connected to a printer, currently reporting the process status.

The operator stations are fully independent and in the same time the data base is the same for all. In this way if one of the operator stations malfunctions, the others remain in operating mode.

The main modules of the system proposed as multifunctional controller, communication system and power supply are fully reserved, thus ensuring very high degree of reliability.

4.4.2.3. Field instrumentation implementation.

For implementation of advanced control schemes, proposed above, it is necessary to provide some additional field instruments, mainly measuring devices. A specification of the mantioned equipment, with prefferable suppliers, is listed in Table 2 ANNEX 4. Note, that all temperatures are connected via signal cards, integrated in the digital control system, therefore no temperature converters are required.

Interfacing digital control system with process via existing field instrumentation is not possible. Moreover, it can't be recommended because of the relatively low accurasy, repeatability, dynamic responsibility,etc.

Solving the task one can consider three possible approaches:

a/ Full replacement of existing pneumatic instruments, with electronic ones;

b/ Install P/I converters, using the existing measuring devices as sensors;

c/ Combination of the above two approaches,

Taking a decision it is necessary to keep in mind, that most of the pneumatic equiplent which is to be used implementing alternative (b) are mounted in the control room. Therefore digital control modules have to be also installed there. In this case no intrinsic safety hardware is necessary, but the accurasy of the system is limited to that of pneumatic instruments.

First alternative (a) is the best one, removing all problems connected with the pneumatics.The digital control system can be distributed through the plant, by this way reducing cable costs. Significantly unfortunately this one is not expensive.

We think the third one to be the most appropriate. All closed control loops and measurements, related with advanced control solutions proposed have to be electronically equipped. The rest part of the information can be collected from the existing instrumentation using P/I converters. Thus automatic control information shall be high accurate and reliable and fully resurved.The information used only for indication shall be left on the pneumatic instruments, but it can't affect the main system functions in case of a eventual failure.

The necessity of oxygen analizers for charge heaters and hot oil heater was already discussed in part D, 4.4.2. By our opinion it is desirable to be installed one more moisture analyser for benzene column A-V9 bottoms, because the end product quality is related with this feed.

4.4.2.4. Possible credits.

In item 3.3. an approach for evaluation of economic and social benefits was described. For LAB plant such data was not available, so the possible savings here below are estimated on a basis of similar process advanced control implementation.

The total equivalent fuel oil consumption of Pacol charge heater Pa-H1 is about 120kg/h. For hot oil heater H-H2 this value is equal to 300 kg/h. From many applications of combustion advanced control, described in part D 4.4.2.1.it is estimated fuel oil saving 5-18%.Bigger savings were obtained changing natural with forced air furnace. For similar heaters it is about 7-15%. So the annual benefits is expected to be :

3.12t/h x 280$/t X 8000h/y x 0.07 = 489216$/y

Next point is reboilers heat transfer advanced control, described in part A. The savings estimated in similar chemical complexes or total for corresponding reboilers annually:

1.9t/h x 280 $/t x 8000h/y x 0.03 = 127680$/y

Note, that Hydrogen process unit charge heater H-H1 is not considered calculating savings.

Total for LAB plant, implementing advanced control strategies with digital control system we expect annually savings 616896 $/y. These are fully demonstrated hard credits. In ANNEX 4, Table 1 the necessary investments are defined.

From corresponding columns in LAB plant additional savings will come, mainly improving product quality.

These are difficult to be calculated ,so obviously they are soft credits mentioned above (Item 3.3.)

4.5. Industrial Utilities.

4.5.1. Background information

The functional scheme of the existing in IPCL - BARODA Complex industrial utilities is presented in Fig.4.5.1.

The Industrial Utilities Plant -

a/ Supplies:

1. Steam with pressure 42,14,5,and 3 kg/cm2	to 13 plants
2. Nitrogen	to 13 plants
3. Oxygen	to EG plant
4. Process air	to 13 plants
5. Instrument air	to 13 plants
6. Cooling water	to 13 plants
7. Raw water	to 13 plants
8. Demineralized water	to 13 plants
9. Process water	to ACN plant
10.Refrigeration	to AF & PBR plant
11.Electricity	to 13 plants

b/ Receives:

1. Condensate /steam/	from 13 plants
2. Electricity	from Baroda network

At present 3 medium size boilers with steam production capacity of 80 t/hour each are in operation These boilers are operated from local control panels equipped with pneumatic instruments.

Two additional Mitsui greater size boilers are in operation also. Each of this boiler produce 130 t/hour steam and each is

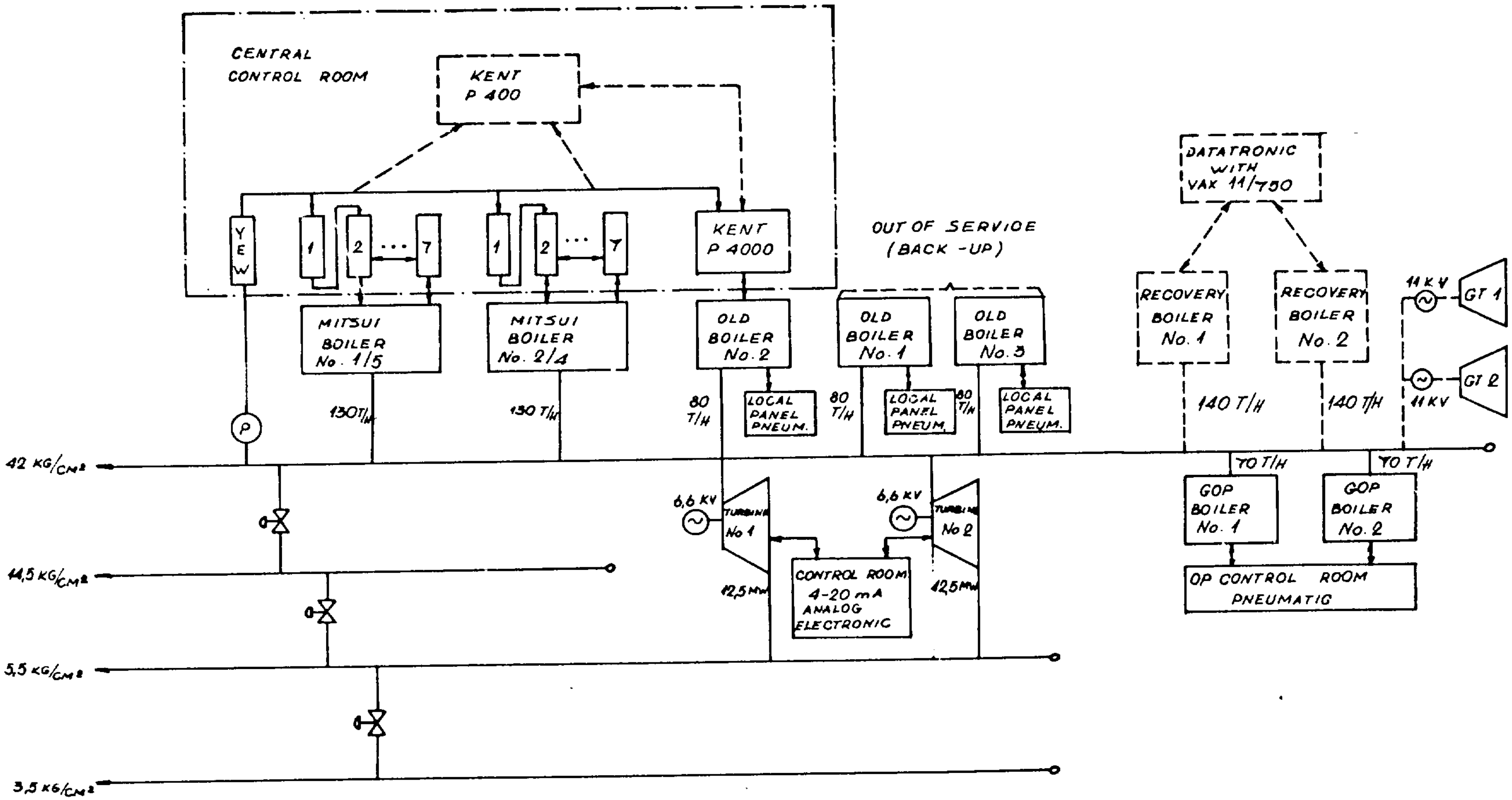


FIG. 4.5.1. INDUSTRIAL UTILITES PLANT
(FUNCTIONAL SCHEME)

equipped with four dual fuel burners.
The control system implemented for Mitsui boilers is with multiloop electronic type controllers, produced by Yokogawa from Japan.
The combustion control schemes for the Mitsui boilers includes the following main control elements:

- Master pressure controller;

- Air cross limiting;

- Gas fuel flow is in BTU; pressure, temperature and density compensated;

- Oil fuel flow is in BTU, pressure and temperature compensated;

- Air measurement is provided;

- Oxygen in the gases is measured by Westinghouse analyzer;

- CO in the flue gases is measured and indicated, but is not included in the closed loop control, because there are common variation of the fuel quality and of the boiler's loading.

The burners control and interlocking system is realized by implementation of Programmable Logic Controller (PLC)-Network 90, produced by Control Bailey.

The steady state contents of oxygen in the outlet gases is about 1,8 to 2,0 %.

After detailed examination and provided discussions with IPCL specialists the following future projects, which at present are in different realization stages, for industrial utilities expansion and modernization are described.

1. Project for installation of two new Recovery boilers with 140 t/hour steam production capacity each. This boilers will be equipped with digital electronic control system- General Electronic Data Tronic Information and Control system which includes a VAX 11/750 supervisory level computer with operator CRT colour consoles.

2. Project for modernization of the control system of the old boiler No 3 with implementation of microprocessor based system of G.KENT P400 and P200 with CRT based operator console in the existing control room.

3. After commissioning of the new recovery boilers it is planned to use the old boilers No 1 and No 3 like a back-up and those both boilers will be normally out of service.

The main objective is to recommend an integrated digital system supporting management of utilities in IPCL and to optimize the energy production and consumption after the realization of the abovementioned revamping of the boilers and turbines . The concrete objective is to suggest a software system for optimal control of the boiler loading and steam distribution systems with balancing capabilities for energy utilities after revamping of the boiler units.

In this case our suggestions concerne the supervizory integrated computer control system which can be implemented after modernization of the control systems for each particular utility unit with implementation of microprocessor based electronic systems.

4.5.2. Recommendations for industrial utilities management system implementation.

After detailed examination of different possibilities for realization of the abovementioned tasks we suggest to modify and apply the existing Industrial Utilities Management System /IUMS/software application modules, designed by HONEYWELL, for use with the VAX 11/750 computer after realization of the hardware connections between this computer and the old boiler No 2,the two Mitsui boilers, all turbines, electricity distribution subsystems and other utilities devices and units.

IUMS modules may be applied to utilities and powerhouse equipment and system including the following:

- Equipment

  Boilers - Conventional and Recovery

  Turbogenerators

  Gas Turbines

  Air and Gas Compression

  Refrigeration

  Auxilliaries - Deareators, Pumps,Fans, Heaters

- Systems

  Steam System

  Electronic Power System

  Fuel System

  Coolong Water and Other Support Systems

IUMS modules do use the base system for fundamental system operations such as data base building, data acquisition, control operator interface and reporting. Table 1 outlines IUMS modules, the base system and the remaining site specific application work.

IUMS is made up of standard application modules, which are tailored and applied to the needs of a specific site by configuration, rather than programming. Configuration involves building points that use the built-in utilities calculation and control functions. This permits the user to select the calculations and functions that are needed for specific units and systems, and to specify sources of input data, scheduling information and other information.

There are twelve IUM modules, each of which includes functions that apply to a specific portion of utility plant. The modules are designed so that each may be optionally selected, based on user requirements. The Support Calculations Module is generally required on all IUM systems, except for those systems that use only the Power Emergency load Shed Module. The optimal modules may be grouped into four main categories.

1. Equipment Monitor and Control
2. System Control
3. System Optimization
4. Emission Compliance Monitoring and Reporting

Equipment Monitoring and Control Modules develop efficiency data for steam and power generating equipment, refrigeration equipment and compressors. The data are used to track degradation in equipment performance and for input to optimization modules.

System Control modules provide:

1. Adjustment in the amounts of generated and purchased power to minimize the purchased power peak demand and cost.

2. Shedding of electrical loads to compensate for partial loss of the power supply.

3. Shedding of non-critical steam loads in the event of a sudden reduction in steam supply.

The System Optimization Modules provide optimal operation of single boilers, multiple boilers, turbines and header systems and of the overall system. The three types of optimization are:

1. Evolutionary Optimization (EVOP), determines optimal air feed rate of an individual boiler by varying its stack O2 setpoint. It requires no system model and may be used alone or in conjunction with the other optimization modules.

2. Configuration Optimization - Configurable Optimization optimizes multiple suppliers of a common demand (e.g. several boilers supplying steam to the same header) and also a system of steam turbines and headers. The optimization model is built by configuration of data base points.

3. General Optimization - The General Optimization Support Module used in conjunction with On-line Process Optimization (OPO) optimizes the overall utilities system. It uses a model program that is a mathematical representation of the utilities system to determine the optimum values of independent costs while meeting all constraints and demands.

General Optimization and Configurable Optimization may both be used with EVOP; however it should be noted that they are not used together. General Optimization is used in place of Configurable Optimization when Configurable Optimization does not adequately treat a complex utilities plant.

The Emission Compliance Monitoring and Reporting Module retrieves and outputs selected emission related data averages in a special report format that is tailored to the Environmental Protection Agency's reporting requirements.

A listing of the individual modules appears in Table 2. Table 3 shows the interdependencies between them. Figure 1 is a block diagram depicting the relationship between the modules at a general level.

BASE SYSTEM	IUM MODULE	APPLICATION WORK
Process Interface Calculation Algorithms Control Algorithms Optimization Operator Interface BPL Programming FORTRAN - Scheduling - Data Base - Interface Data Base Build Display Build History Reporting On-line Process Optimization (OPO) (optional) Freetime (background processing) Files Management	Utilities Algorithms - Monitoring - Control Configurable Optimization - Economic Load - Allocation - Turbine/Header - Optimization Evolutionary Optimization General Optimization Support EPA Reporting	Data Base - Process I/O - IUMS Calculated - Information - IUMS Control Displays Tailoring Optimization Model(optional) Minimal Programming

Table 1 - IUMS Modules, The Base System, And Application Work

BASE	
110	Support Calculations
EQUIPMENT MONITOR AND CONTROL	
210	Boiler and Steam System Monitoring and Control
220	Generation and Power Monitoring
230	Compressor Monitoring
240	Refrigeration Monitoring
SYSTEM CONTROL	
310	Tie Line Control (see Note 1)
320	Power Emergency Load Shed (see Note 1)
330	Steam Emergency Load Shed (see Note 2)
SYSTEM OPTIMIZATION	
410	Evolutionary Optimization (see Note 3)
420	Configurable Optimization (see Note 3)
430	General Optimization Support (see Note 3 and Note 4)
PLANTWIDE MONITOR AND AUDIT	
510	Emission Compliance Monitoring and Reporting

NOTES

1. The Tie Line Control module and the Power Emergency Load Shed module both require the Generation and Power Monitoring module.
2. The Steam Emergency Load Shed module may require the Boiler and Steam module to provide calculations of unmeasured flows into or out of the steam balances.
3. The optimization modules require any other modules pertaining to equipment or systems included in the optimization.
4. The General Optimization Module requires FORTRAN and OPO and is therefore available on TOTAL or Custom base systems only.

Table 2 - IUMS Modules

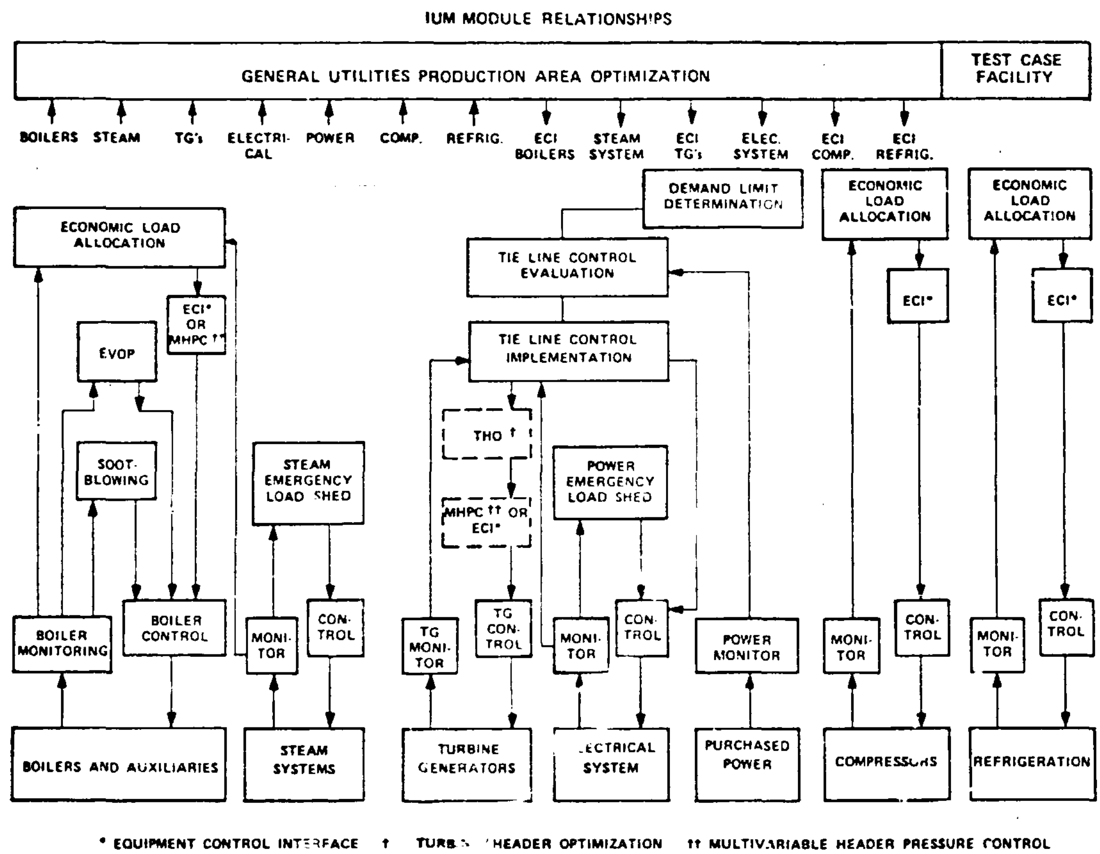


Figure 1 - General IUM Module Relationships

MODULE NAME & NUMBER	OTHER MODULES REQUIRED	SPECIFIC BASE SYSTEM REQUIRED
110 SUPPORT CALCULATIONS	NO	NO
210 BOILER & STEAM SYSTEM MONITORING & CONTROL	MODULE 110	NO
220 COMPRESSOR & POWER MONITORING	MODULE 110	NO
230 COMPRESSOR MONITORING	MODULE 110	NO
240 REFRIGERATION MONITORING	MODULE 110	NO
310 TIE LINE CONTROL	MODULES 110 & 220	NO
320 POWER EMERGENCY LOAD SHED	NO	NO
330 STEAM EMERGENCY LOAD SHED	MODULES 110 & 210	NO
410 EVOLUTIONARY OPTIMIZA-TION	MODULES 110&210 TROUGH 330 AS APPROPRIATE	NO
420 CONFIGURABLE OPTIMIZATION	MODULES 110 & 210 TROUGH 330 AS APPROPRIATE	NO
430 GENERAL OPTIMIZATION SUPPORT	MODULES 110 & 210 TROUGH 330 AS APPROPRIATE	NO
510 EMISSION COMPLIANCE MONITORING AND REPORTING	NO	NO

Table 3 -IUMS Module Interdependencies

A given IUMS module may consist of any or all of the following general components: algorithms, non-algorithm software and application documentation.All IUMS functions are accomplised using these components. Figure 2 shows the relationship of the modules to the general components.

Algorithms

A major portion of IUMS is implemented as Process Variable (PV) Algorithms and Control Algorithms, which are added to the complement of algorithms that exist as part of the base system. These algorithms are preengineered calculations that support operating information, advanced control and optimization . In general, the IUMS algorithms carry out functions that are larger in scope than those of the base system algorithms. For this reason, they have greater numbers of inputs and outputs.However parameter limits and other point structure constraints defined by the base system are not exceeded. This enables the base system file builder to be used for building IUMS points in the same fashion as base system points. It also enables the base system point display to be used for IUMS points.

IUMS algorithms are used analogously to the base system algorithms.Variable input data from other parts of the process data base is accessed from PVC (input) pointers. Constant input data is accessed from PVC words (coefficients for Control Algorithms). The Major output value is returned as the PV for PV Algorithms, or as the output of Control Algorithms. Secondary outputs are returned in PVC or AGC words.

IPT pointers in the base systems specify the point and parameter from which tha data is to be input. Some IUMS algorithms extend the use of input pointers to allow the algorithm to access or store data in a number of referenced point attributes (defined for the particular algorithm). This is in addition to the one attribute explicitly specified by the IPT pointer.

Non-algorithm software.

This component in IUMS includes special points, BPL or FORTRAN programs that are supplied pre-built for a specific purpose. Some may require modification for a specific purpose. Some may require modification to match them to a specific plant or user requirement. Non-algorithm software is used chiefly in General Optimization Support.

Some IUMS installations may include other application BPL and/or FORTRAN programming in addition to the algorithms and non-algorithm software to augment the standard IUMS features.

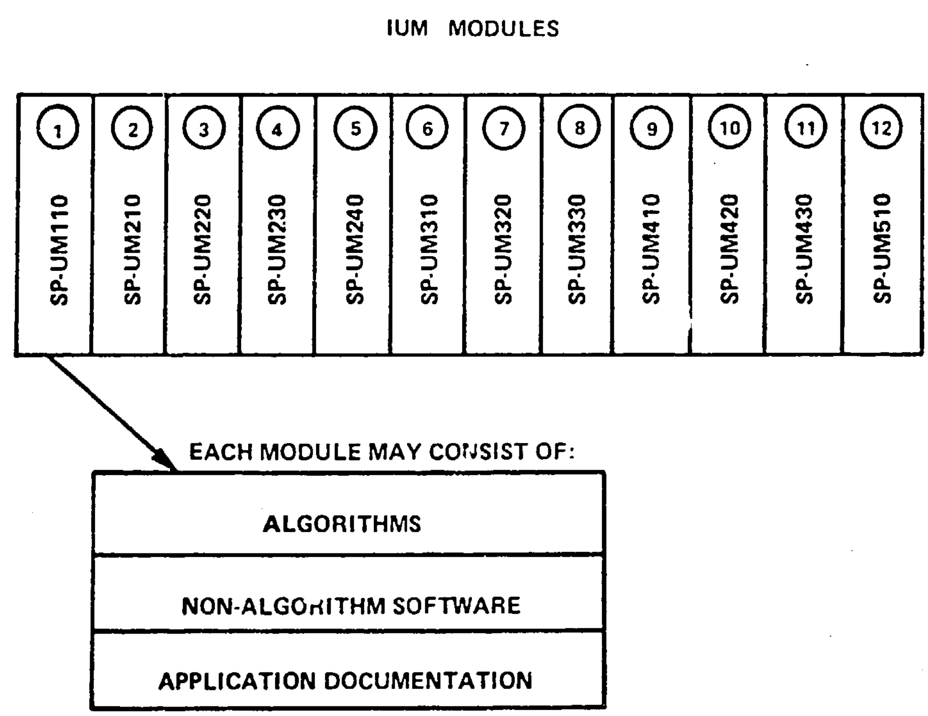


Figure 2 - IUM Modules and General Components

More details about IUMS are given in APPENDIX C of the present report.

4.5.3. Possibilities for Technical Realization of IUMS

There are two general possibilities for technical realization of the hardware and software connection and communication between the supervisory level VAX 11/750 computer, which will be implemented with the installation of the new recovery boilers, and the existing control system of the Mitsui boilers-electronic multiloop controllers from Yokogawa, and the planned to be installed electronic system for old boiler No 2 from G.Kent:

1. To expand the input/output capabilities of the Speed-Tronic Mark IV DAS system, which is part of the GE Datatronic Information and Control system, for abovementioned boilers control.

2. To implement the so-called High Level Process Interface Unit /HL-PIU/, produced by Honeywell like part of TDC-3000 system, with analog and digital input/output boards. To use additionally a Hiway Traffic Director /HTD/and General Purpose Computer Interface /GPCI/ devices from TDC -3000 system and on second stage to realize a connection to the supervizory level PMX H-45000 computer, where the IUMS modules can be used.

If the first approach will be chosen the hardware problems for realization of the IUMS integrating all utilities in IPCL will be less but the software work for adaptation of the IUMS modules for use with the VAX computer will be much greater.

However in both cases the payback time according to the international and own experience will be less than 2 years.

4.5.4. Possible benefits from IUMS implementation

The benefits can be achieved at all levels of operation:

- power operators can examine all information needed to maximize equipment operating efficiency;

- Maintenance personnel can obtain heat transfer and efficiency trend reports to help establish priorities for overhaul and cleaning.

- engineers can study proposed changes using the computer in background mode for simulations and other calculations, while the system is performing normal energy management duties;

- management can use the risk of shutting down spare equipment, and if overtime can be justified to return efficient equipment to service.

5.INTEGRATED COMPUTER CONTROL SYSTEM

5.1.Functional schemes of ICCS.

Distributed microprocessor based system for data acquisition and process control have many attractive features - relatively low cost,graceful degradation, modularity and simple real-time software requirements. Those base-level digital systems can be easily integrated in the computerized process control systems, where the process computers allow supervisory setpoint control and sophisticated control schemes to be implemented. They allow process models to be used as a part of advanced control strategies as well as more process data to be available to operative personnel and to process engineers in formats that can be easily understood such as process graphics or well designed reports.

The successful fulfilment of all these functions for the suggested modernization of the control system IPCL production plant can be shared between the proposed in the previous section 4 (of the present report) distributed process control system and the supervizory level computer system which will be described in this section 5.

On the basis of dataway communication and application of digital process control and process monitoring functions with process optimization, production planning and control as well as energy and total plant management functions. The control hierarchy in an ICCS of a process industry production complex can be devided into the following functional levels:

Level 1 : - Instrumentation and Process Control

Level 2 : - Monitoring and Process Optimization

Level 3 : - Production Planning and Energy Management

Level 4 : - Economic Planning and Total Plant Management

The functions of an Integrated Computer Control System for total plant management are shown schematically on Fig.5.1.Fig 5.2. represents the capability of the TDC -3000 system for practical realozation of Integrated information data base from the sensor to the boardroom of a great production complex.

The Functional scheme of an ICCS for DMT Plant, EG Plant,ACN Plant, LAB and Industrial Utilities Plant of IPCL Complex in BARODA, which is recommended for implementation on the second stage of the process control system revamp for overmentioned plants, is shown on Fig.5.3. On this figure the TDC - 3000

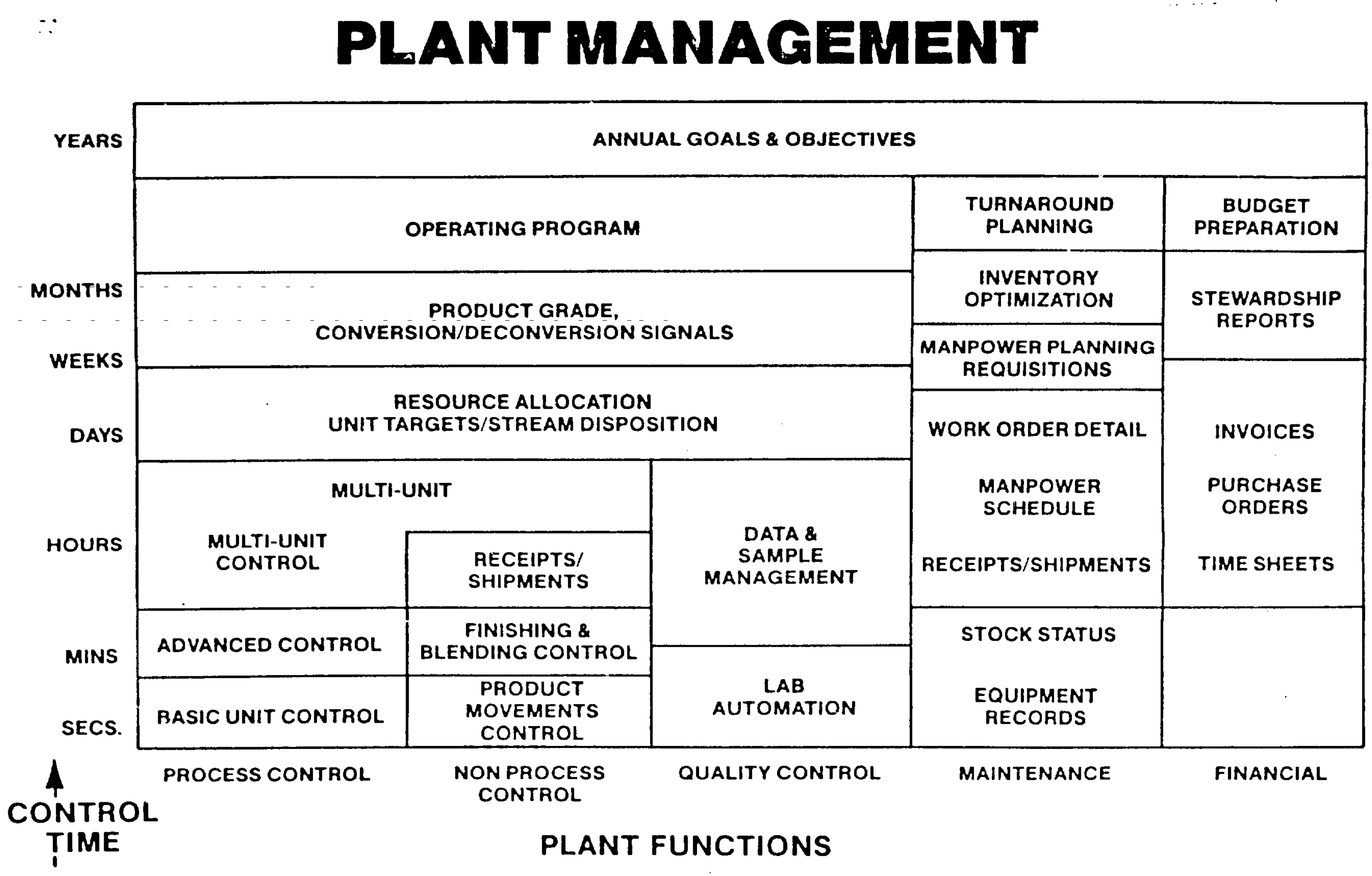


Fig. 5.1. FUNCTIONS OF an ICCS

INTEGRATED INFORMATION
FROM THE SENSOR TO THE BOARDROOM

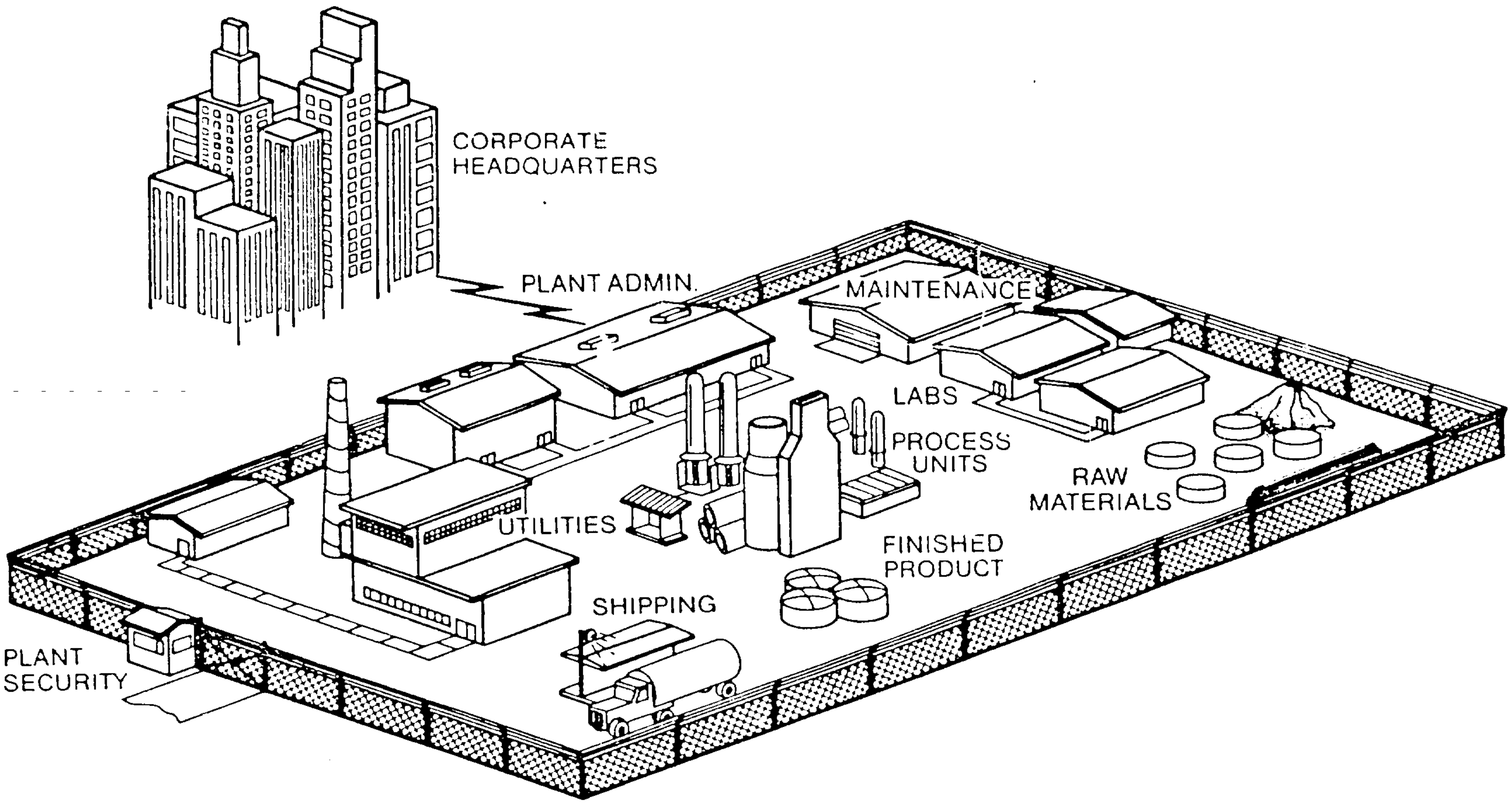


Fig. 5.2.

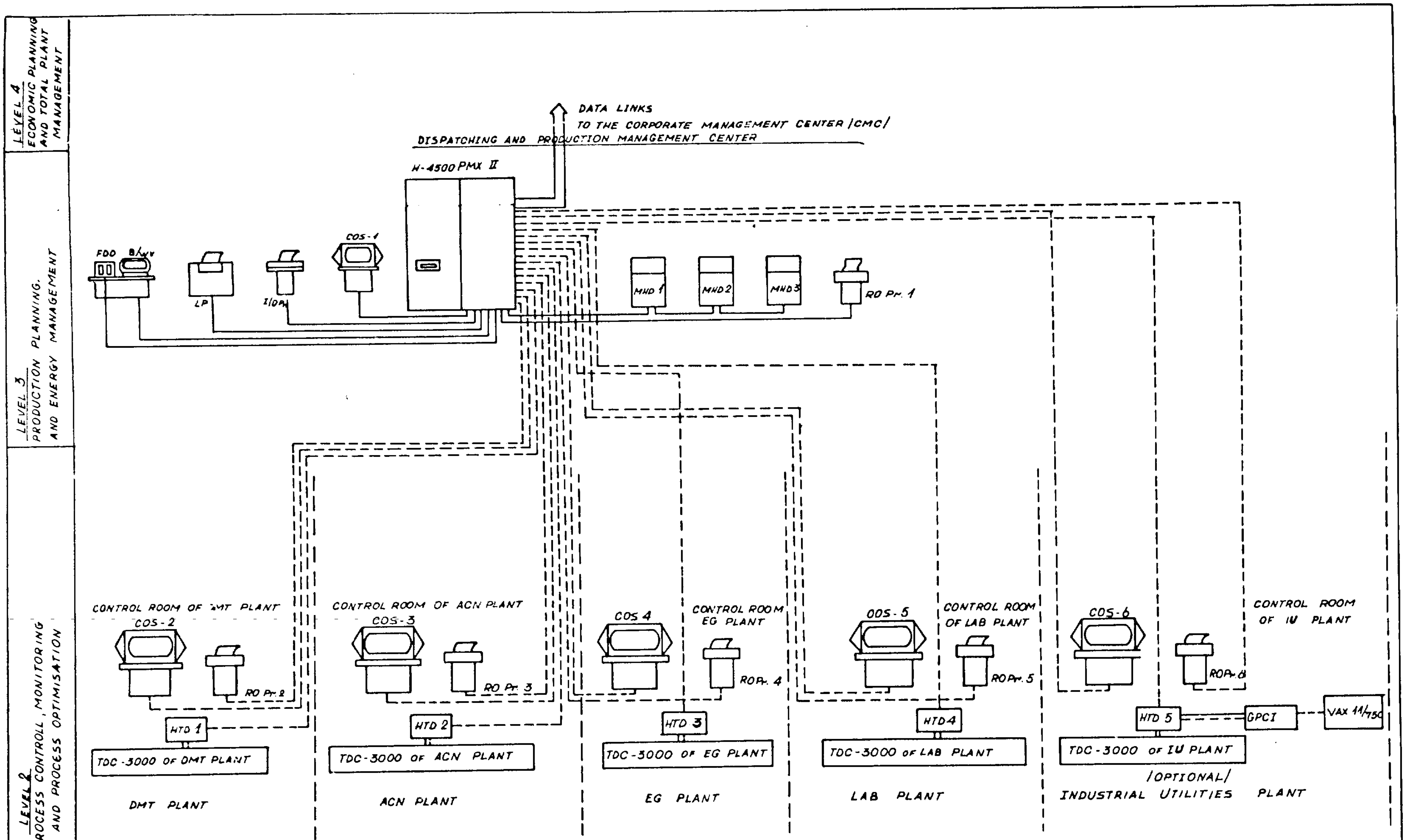


FIG. 5.5. TDC-3000 COMPUTER CONFIGURATION

system configuration for realization of process control monitoring and partly optimization functions, which have been described in section 4 represented only on the form of rectangles.

5.2. Specification of additional system modules for realization of ICCS.

Integration of the digital distributed process control and monitoring system with the supervizory process computer rezults in some specific design of the computer system and its application software in oprder to take advantages from the new control philosophies and concepts, from functional capabilities and new requirements offered by digital systems. The main topics duting the design of the suggested hierarchical computer control system for IPCL complex are the realization of the following functional capabilities:

- Stabilizing process control in the limits of given technological requirements;

- Design of advanced control schemes and sophisticated control strategies and their practical realization;

- Identification, control and registration of process dynamics;

- Implementation of process optimization routines;

- Reporting of aggregated production and economic indexes needed for higher levels of the plant management;

Different production units (plants) in IPCL complex , monitored and controlled by corresponding control rooms with suggested new operator station centers /see configuration of these centers in the section 4/ are interconnected by means of direct connections /with or without transportation legs/ or intermediate storages. Due to the interconnections between different types of inputs and outputs of production units, the rational utilization of external and internal resources, which have accumulative (such as raw-materials intermediate and final products) and now-accumulative (like energy, labour, etc) nature is an overall problem of prime importance.

The effective solution of this problem must be based on the particular process control strategies of the production units and can be realized by the corresponding control room, equipped with computer operator stations (COS) and report and alarm printers, connected to the supervizory process control computer.

Consequently, the production control problem in the IPCL complex can be defined, as an overall operative planning and dispatching of the production rates of the processing units and the correspoding level and consumptioin of accumulative and non-accumulative resources available in different production modes so that by utilizing the production and buffering capability in the complex the goals specified by the corporate management are achieved at minimum costs. However, the energy management with associated costs control, orders and maintenance scheduling constitutes the basis for the effective control of the complex as a whole.

All functions considered above can be realized by so-called Dispatching and Production Management Center (DPMC), connected with the process computer and with the control (host) computer, installed in IPCL computer center, via data links.

The architecture suggested for implementation ICCS ,shown on Fig.5.3. is based on TDC -3000 PMX II Release 2 Computer system. It includes the following supervizory level units and devices which shall be distributed in different location in IPCL Complex- BARODA:

1. Dispatching and production management center, which can be equipped in some of the existing centrally situated building after some reconstruction and modernization:

- H-4500 Central Processor Unit with appropriate computer pheripheral controllers and devices, generate with PMX II Release 2 operating software.

- Computer control station No 1.

- Moving head disk drives with necessary disk packs,

- Main console printer and programming video console;

- Report and Line printer devices;

2. Satelite computer control centers;
Each of the suggested in section 4 operator centers for DMT,EG,LAB,ACN and IU Plant, which shall be mounted in the existing control rooms will be additionally equipped with:

- Computer control station -PMX type

- Alarm and report printer

The communication between the TDC - 3000 Multifunctional system configuration and the PMX H-4500 computer can be realized through the connection of the each Hiway Traffic Director /HTD/ with the supervizory computer.

3. Connection with the planned to be installed in Industrial Utilities PLant VAX computer can be realized with implementation of so-called General Purpose Computer Interface /GPCI/ .

Detailed description of the computer functional capabilities and characteristics is attached in APPENDIX B of the present report.

5.3. Hardware and software solutions

A/ The suggested hardware configuration of a supervizory level PMX H-4500 computer for realization of Integrated Computer Control System for DMT, EGH, LAB, ACN and IU Plants in IPCL - BARODA is shown on Fig. 5.3. This hardware configuration can be described as follows:

1. Computer connected stations in the main control rooms of DMT,ACN,LAB,EG and IU Plants.

Each of these satellite supervizory computer connected operator centers includes:

- PMX Operator station with trend channel capabilities;

- Additional Engineering Keyboard for each station;

- Report and alarm printing device;

- Data hiway interface connection with the H-4500 PMX computer which can be mounted on a distance about 1000m from each of these satellite computer console centers. This connection will be realized via outdoor mounted double cables between each Hiway Traffic Director /HTD/ and the Data Hiway Interface module of the Central Processor mounted in the so-called Dispaching and Production Management Center /DPMC/.

2. Dispaching and Production Management Center /DPMC/.

This center will be based on a PMX H-4500 Computer consisting of:

- Central Processor Unit /CPU/ with 1024 KB operative memory, five Data Hiway Interface units and pheripheral controllers for the following devices:

- Two moving head disk drives with one additional disc drive for on-line copying capabilities. These disk drive will use 16 MWords disc packs with generated PMX II Release 2 software system;

- One Computer colour operator station with engineering keyboard and trend channel capabilities;

- One Input/Output printer for system messages and operator's communication with the PMX system software;

- One report and alarm printing device;

- One line printer for user's additional programme development;

- One Black & White video console for user's programming;

- One dual drive Floppy disc device for Hardware testing of the system

3. Communication with the Corporate Management Center /CMC/ of IPCL - BARODA, which can be estimated in the existing Computer Center of the petrochemical complex is suggested to be realized by implementation of two Asynchronous Data Links with appropriate software. The technical specification software for realization of Integrated data base of ICCS is given in Table 5.1.,ANNEX 5.

B/ TDC -3000 PMX computer is supplied by the

following standard software for realization of Integrated data base of the whole five production units.

- Real - Time Multiprogramming Operating System /RTMOS/ ;

- Freetime IV software system for realizatrion of background mode application software programs,

- File's builder software Files IV;

- Additional PMX application oricuted software.

This computer system offers real-time capabilities for Fortran IV user's programming.

All suggested in section 4 advanced and optimal control strategies can be realized by shared data base and application software between the multifunctional controller systems suggested and the supervisory level PMX computer. In addition the production planning and energy management oriented software can be added to the computer H-4500 PMX system.

The Industrial Utilities Management System /IUMS/, described in APPENDIX C, can be realized with the PMX computer also. The detailed technical specification of the system modules, which can be implemented for utilities management, is given in Table 5.2. ANNEX 5.

C/ Suggestions for functions of Corporate Management Center /CMC/

Economic planning is one of the Functions of the management. The practical realization of any long-term plant is closed interconnected with medium and short-term (year, quarter, mounth) plans as well as with the monitoring and control of quantitative and qualitative aspects of managrial, coordinational and operational activities of the complex as a whole . In order to utilize all the resources available in the complex the large integrated data base must accept all the levels of ICCS suggested for successful realization of management decisions. This integrated management system approach depends upon the hierarchical levels of ICCS, the architecture of which is shown in Fig. 5.3.

The highest level of ICCS is so called Corporate Management Center -CMC, which can be established in the Computing Center of IPCL Complex. The CMC is suggested to integrate the production planning and energy management activities of DPMC with the economic planning and coordination activities of the total complex management.

One of the main characteristics of the CMC will be the degree of realization of the ICCS consepts at all functional levels and through the following modularity structured subsystems:

- Corporate Planning;
- Marketing and Distribution
- Research, Development and Engineering
- Production and Energy Management
- Finance and Accounting
- Personnel and Labour.

The practical realization of all these functions will take about 10 man/years of software development work.

SECTION 6

6. GENERAL RECOMMENDATIONS

6.1.DMT Plant.

The analysis of the DMT Plant, the suggestions for optimal and advanced control strategies using a distributed computer control system and the estimation of expected savings, made in item 4.1. of the present report show that the implementation of the proposed system is acceptable from economic point of view, considering only the expected hard credits.

However, large savings can be expected from application of advanced control to hot oil heaters, which are not calculated.

Further credits, which can be classified as soft, are expected to come from advanced control of distillation column and stabilization of intermediate products composition.

In view of the expected reconstruction and enlargement of the DMT Plant and relatively good economic justification our recommendation is to revamp the whole DMT plant instrumentation with the proposed instrumentation and distributed computer control system, implementing optimal and advanced control strategy, as suggested in item 4.1.

6.2. ACN Plant

The analysis of the ACN Plant, the recommendations for advanced control strategies implementation for the main distillation columns of the plant and for optimal control of the reactor unit implementing a distrinbuted microprocessor based process control system, and the estimation of the possible savings and benefits show that the application of the suggested system is economically justified.

Our general suggestion is to start the modernization of the existing pneumatic control system of the ACN Plant with the implementation of a microprocessor based digital control system for the acrylo reactor unit, as proposed in Option (F) of the p.4.2.2. of the present report. After this implementation we recommend to extend the capacity of the digital sistem with revamping the instrumentation of the other production section of the ACN Plant step by step. This approach will be possible because of the modular structure of the suggested distributed system and will give a possibility for distribution of the investments in time.

In parallel with the implementation of the digital process control system the modernization of the existing pneumatic instrumentation will be done with as much as possible application of the new electronic type field instruments.

Finally the payback time for this process control system and instrumentation revamping can be estimated at about 2 years.

6.3. Ethylene Glycol Plant

From the analysis of the plant and of the proposed advanced control strategies, made in item 4.3. of the present report, it is clear that the implementation of an advanced control strategy using a distributed digital control system at the present moment is economically justifiable only for Section 1: Ethylene oxide reaction and scrubbing.

Our general recommendation is to start the modernization of the control system of the EG Plant with the implementation of a distributed computer control system for the ethylene oxide reaction unit, as proposed, and later proceed gradually with revamping the instrumentation of the other sections step by step.

This approacn is possible, because of the modular nature of the recommended distributed computer control system. That makes easy the expansion of the system.

Besides the distribution of the necessary investments in time this approach has got other merits:

- The process personnel will become aquainted with the system in a more gradual way;

- The system will prove its advantages and savings from its implementation before it is installed on a large scale, that is for controlling the whole plant.

6.4.LAB Plant

The advanced control schemes which are to be recommended for LAB Plant are three types, mainly : combustion control of charge heaters, heat transfer control of reboilers and feedforward control of distillation columns. Expected benefits implementing those schemes are fixed.

Three approaches for appropriate field equipment modernization are compared.It is clarified , that suggested advanced control schemes and closed control loops are to be interfaced with digital control system through new electronic equipment. For the rest part of the process information the existing pneumatic instrumentation with P/I converters can be used. On the base of this approach still remains the possibility of replacing the pneumatic instrumentation in the future. In the same time these exesive P/I converters can be utiliesed for the other installations modernization.

The main tool for realizing the recommended control strategies is a digital distributed control system.An appropriate hardware configuration is described and the recommended supplier for this system is Honeywell.

As a general recommendation it is proposed to change the type of the existing industial furnaces from natural to forced air supplying.

This is the most important point and can be performed after consultations with UOP,USA, possibly without replacing the furnaces themselves.

The necessity of oxigen analyzers for outlet gases is proven and prefered supplier Westinghouse is recommended.It is pointed the advantage of Agar Instruments moisture analyzer implementation in Alkylation unit.

Finally it can be recommended to start LAB control system modernization with realization of combustion advanced control as a first stage.There is a possibility (depending on available investments) to perform only this stage of modernization.

6.5. Industrial utilities.

From the recommendations for implementation of Industrial Utilities Management System /IUMS/made in p.4.5.2 of the present report is clear that the realization of integrated boiler control and steam distribution control in IPCL Industrial Utilities Plant on the first stage is economically acceptable. On the next stage of the practical implementation of IUMS the described software modules can be realized on step by step basis.

Generally we suggest the following sequence for implementation of the IUMS software modules:

1. Support Calculations and Boiler and System Mounting and Control
2. Emission Compliance, Mounting and Reporting
3. Steam Emergency Load Shed
4. Generation and Power Monitoring
5. Power Emergency Load Shed
6. Compressor and Refrigeration Monitoring
7. Evalutionary and Configurable Optimization
8. General Optimization Support

The final success of the practical realization of the suggested IUMS depends mainly on the unification of the hardware and software suppliers. The integration between the basic digital systems from different producers can cause great technical problems.

6.6.Recommended stage of implementation

The basic approaches when implementing digital control systems in an industrial complex is to start either with the revamp of the first in the stream plant or with this one where the return of investments is shortest.

In the case of IPCL BARODA Complex we suggest to follow a workable compromise of aforementioned approahes and procede with the implementation of a istributed computer control system as follows:

<u>I Stage</u> : Olefins plant and Industrial Utilities

Althought not studied in the present report it is obvious from international experience that biggest savings and improvement of downstream plant operation will result from implementation of advanced and optimal control to this plant. These are a number of commercially available optimization packages, such as SETCONT by Setpoint Inc.(marketed also by HONEYWELL).

What is more, there is a large experience and confirmed economic benefits from implementation of a digital control system in oletin plants.

Simultaneously it is advisable to proceed with the implementation of an industrial utilities software management system which is expected to require small capital investments and to yield major savings in the energy bill.

<u>II Stage</u> : DMT Plant.

Based on the recommendation to revamp the whole plant instrumentation and the expected good economic results it is advisable to proceed with the DMT Plant control system modernization on the second stage.

<u>III Stage</u> : Ethylene Glycol Plant, Linear Alkyl benzene Plant and Acrylonitrile PLant.

At the present the economic expectation from a modernization of the instrumenation in these plants are not so high and generally it is recommended to implement modern control systems only in some units of the plants, it is advisable to have their control systems modernization for the last stage.

6.7.Recommendations for preferrable instrumentation company suppliers

The recommendation for revamping of field instruments are given in item 4 for each investigated plant.

Generally for flow, pressure and level measurements pressure and differential pressure electronic silicon-type transmitters are suggested, whenever they are applicable.

Preferred vendors for these instruments tank on our experience are HONEYWELL, ROSEMOUNT, HARTMAN & BROUN, FISHER& PORTER and FOXBORO. The functional characteristics of the instruments of these vendors are comparable and among the best. The final choice should be made considering other points, such as financial conditions, service availability, etc., which are different in each case into consideration.

One noteable exception is the series ST 30000 smart transmitter by HONEYWELL , which is functionally on a high level than standard conventional transmitters, but is also about 30% more expensive.

For high vescosity flow measurements rotameters by KROHNE and FISHER & PORTER are recommended. The functional capabilities of both vendors' choice should be based on other, non-technical consideration, for which no general recommendation can be given.

For positive displacement meters, based on our experience, only one vendor, BOPP & RENTER is recommended, because the quality of its production is quite higher, thah that of its competitors.

For temperature measurements RTD's are generally recommended. Only two vendors HONEYWELL and DEGUSSA are indicated, known for their wide range and good quality of these products. However there are also quite a number of other vendors, offering RTD's with comparable characteristics. Final choice of these instruments should be determined mainly by the financial conditions in each case.

Five vendors are recommended for control and cutoff valves - HONEYWELL, GULDE, MASONEILAN, FISHER and NELES. The valves of these vendors are characterized by high quality standards, easy service, long life, durability and good references. HONEYWELL, MASONEILAN and FISHER have wide range of products for a lot of applications, while GULDE and NELES products are more or less specialized, suitable for essential or unique applications. NELES ball valves are especially suitable for hasardous conditions.

The List of preferrable instrumenation vendors is attached in Table 6.7. below.

TABLE 6.7

LIST OF PREFERRABLE INSTRUMENTATION VENDORS FOR PETROCHEMICAL INDUSTRY

A. P AND DP ELECTRONIC TRANSMITTERS

1. HONEYWELL /USA/
2. ROSEMOUNT /USA/
3. H & B /FRG/
4. F & P /USA/
5. FOXBORO /USA/

B. ROTAMETERS

1. KROHNE /FRG/
2. F & P /USA/

C. POSITIVE DISPLACEMENT METERS : BOPP & REUTER

D. CONTROL AND CUTOFF VALVES

1. HONEYWELL
2. GULDE
3. MASONEILAN
4. FISHER
5. NELES /FINLAND/

E. RTD

1. HONEYWELL
2. DEGUSSA

6.8.Comparative evaluation of distributed digital control system for ppetrochemical industries - general recommendation

6.8.1. Introduction

In the second half of the 1970's several digital automation system with new decentralized control and centralized monitoring were introduced. On the basis of digital techniques and dataway communication it is now possible to integrate the automatic control monitoring with production planning and management.

The generalized architecture of an integrated production management system typical of the petrochemical industry is given in Fig.1.It consists of five hierarchical levels, namely:

1. Measurements

2. Stabilizing control

3. Set-point optimization

4. Operative planning or scheduling

5. Strategic planning.

From the organizational aspect these levels are usually specified as follows:

1. Field or loop

2. Group or aggregate

3. Process or department

4. Mill or plant

5. Integrated mill or corporate

The time factor, especially the decision making period and the time of consideration (planning horizon), plays an important role in formulating the goals and problems at each of the levels mentioned above. At the measurmewnt level the decision periods are usually given in s or min., at the stabilizing control level in min or h, at the set-point optimization level in h, shifts or days and operative planning level in months, quarters or years.

The control hierarchy of integrated process industry mills usually be devided into four functional levels or so-called strata , as follows: (cf. Fig. 1)

1. Instrumentation and process control

2. Monitoring and optimization

3. Production control

4. Corporate management

The main purpose of the unified approach described is to analyze the functional capabilities and capacity parameters of the different units and systems, as well as the interrelationships which are realized by what is called dataway (bus) communication of information exchange. The functional capabilities and capacity parameters determined later are based on certain known experience with digital distributed control system used in petrochemical industry. Therefore the approach suggested is and can be used in analysing different systems available at present on the market.The technology presented can be used successfully at the first stage of the design of integrated production management systems in different petrochemical industry mills.

6.8.2.Instrumentation and Process Control

6.8.2.1.Measurement and field instrumentations

The notations and symbols used at the "MEASUREMENTS" level are given in Fig. 1 .

The following conventions for Analog (A), Binary (B) and Frequency (F) Inputs (I) are used in this section:

1. Number (N) of Analog Inputs - NAI

2. Number of Binary Inputs - NBI

3. Number of Frequency Inputs - NFI

Analog process vartiables measured in process industry mills can (without major loss of generality) be classified into five main groups, namely:

1. Temperatures - T, of which by Termo Couples - TC and Termo Resistors - TR.

2. Pressure - P

3. Flows - F

4. Level - L

5. Concentrtations and analysing measurements - C

The analog inputs can be further specified using these symbols. For instance, NAIT of which NAITC and NAIP, etc.

MUltipleXers- MUX are usually of two types, Analog-AMUX and Digital- DMUX.

The total Number - N of MUltipleXors is logically denoted by NMUX which usually represents the sum of the Number of AMUX(s) and the Number of DMUX (s), i.e. NMUX = NAMUX + NDMUX.

Each AMUX is specified by the Number of Analog Inputs/Outputs- NDI/O (cf.Fig.1)

Each AMUX can further be specified by the process variabls symbols - T,P,F,L nad C in the manner presented above.

The Maximal Disyance at which a giver Multiplexing Process Interface Unit - MPIU is dentoed by MDMUX.

The multiplexors and all the other units presented in Fig. 1 can be ordered and numbered in the familiar manner - MUX 1, MUX 2, ..., MUXN or MUXI, I = 1,2,...,N.

Every production management system (no matter how simple or complex) depends on the accuracy and reliability of its field instrumentation. Therefore the specification of the main

- primary sensors
- transmitters
- analysers
- actuators, positioners and transducers
- control valves

is also necessary in studying the critical aspects of system reliability as a whole.

6.8.2.2.Controller Station - CS.

The basic unit at stabilizing control level is a group or aggregate Controller(s) Station - (CS), the total number of which is denoted by NCS. Every CS can be specified by the following 20 capacity parameters and functional capabilities:

1.Number of Analog Inputs - NAI

2. Number of Binary Inputs - NBI

3. Number of Frequency Inputs - NFI

4. Number of Analog Outputs - NAO

5. Number of Binary Outputs - NBO

6. Number of Frequency Outputs - NBO

7. Number of Control Algorithms - NCAL

8. Number of Alarms - NALA

9. LINearization capabilities - LIN

10.DIgital FLittering capabilities - DIFI

11.NOnLInear functions - NOLI

12.CAscade and RAtio control capabilities - cara

13.Number of Histories - NH, with different time of consideration - T (min,h) and discrete (averages) intervals - DT (s,min,h)

14. REVERsing action capabilities - REVE

15. TYpe of PRocessor TYPR

16. SElf DIagnostics - SEDI

17. MEMory Security -MEMS

18. Flexible Configuration and Programming Capability - FCPC

19. Automatic Balancing and Self Tuning - ABST

20. Local Operator Interface - LOI with Number of Supervized Control Loops - NSCL

6.8.2.3.Redundant Copntrol Unit - RCU

The Redundant Control Unit (RCU) is basically used for most important loops so that the reliability of the system is increased at the desired level. The number of RCUs is logically denoted by NRCU. Normally, an RCU consists of a Transfer Unit- TU and a Reverse Controller - RC, (cf.Fig.) which have the same specific features as CS.Therefore the RCU is characterized only by the following 5 parameters.

1. Number of Redundant Analog Loops - NRAL

2. Number of Redundant Binary Loops - NRBL

3. Number of Redundant Frequency Loops - NRFL

4. Transmitter Time - TT

5. TYpe of PRocessor - TYPR.

6.8.2.4. Process Interface Unit - PIU

In general, the Process Interface Unit(s), the total Number of which is NPIU,consist(s) of 4 types of various process interface units, namely

1. Multiplexing PIU - MPIU

2. Standard PIU - SPIU

3. General PIU - GPIU

4. Analog PIU - APIU

The purpose of each of the types of process interface unit mentioned above can easily be understood from the graphic symbols used in Fig.1 and presented in Table 1.

It is logical to assume that all these process interface units contain LINewarization ,FIltration and NOnLInearization capabilities, which are not especially indicated in Fig.1.

In addition to the notations and conventions explained above, the following symbols are used in different types of process interface units:

1. Number of connected MUltipleXors - NMUX

2. Number of Sub-Unit in given PIU - NSU

3. Number of Analog Inputs/Outputs - NAI/O

4. Number of Binary Inputs/Outputs - NBI/O

5. Number of Frequency Inputs/Outputs - NFI/O

6. Measuring Speed in Points /s - MS

7. Scanning Speed in s/point - SS

8. SIgnal Testing capabilities - SITE

9. ALArming capability - ALA

10. Direct Digital Control capability - DDC

11. SELf DIagnostic - SEDI

12. TYpe of PRocessor - TYPR

Each PIU can be further specified with respect to the type of process variables and corresponding measuring devices, in the manner shown above.

6.8.2.5.Programmable Logic Unit - PLU

The main functional capabilities of a PLU the total number of which is NPLU, can be defined as follows:

- Group (aggregate) logical operations
- Direct control
- Automatic start-up of the process
- Automatic shut-down of the process
- Numerical program control

Therefore this unit can be characterized by the following capacity parameters:

1. Number of Analog Inputs - NAI

2. Number of Discrete Inputs - NDI

3. Number of Analog Outputs - NAO

4. Number of Discrete Outputs - ndo

5. Time of consideration - T (s,min,h)

6. Discrete interval - DT (s,min,h)

7. Number of Programmable Operations - NPO

8. Number Bolean Variables - NBV

9. Number of Position Controls - NPC

Furhter specification of the measured and controlled process variables can also be indicated in the PLU block represented in Fig 1.

6.8.3.Monitoring and Process Optimization

6.8.3.1.Dataway and Data Traffic Controller - DTC

The dataway plays the key role in each digital automation system with decentralized architecture. This simple and effective communication link integrates the Stabilization control level and process Set-points optimization level in the strata called Monitoring and process optimization (cf.fig.1) The reliability and other advantages of decentralized control can at the same time be preserved.

The information traffic (exchange), consisting of data messages of various lenght is basically controlled by Dataway Traffic Controller(s) - DTC, the total Number of which is denoted by NDTC. Each , DTC can be characterized by the following capacity parameters and functional capabilities:

1. Number of Data Ways (Branches) - NDW
2. Redundancy of Data Way - RDW
3. Maximal Distance of Data Way - MDW
4. Number of units (devices) on one Data Way (Branch) - NUD
5. Number of Units Controlled by one DTC-NUC
6. Number of Priority levels - NPL
7. Number of Synchronous Channels - NSC
8. Number of Asynchronous Channels - NAC
9. Communication Speed - CS (K baud/s)
10. STANdard used in communication - STAN
11. PROTocol Type - PROT
12. MESSage Type - MEST
13. CODe Type (for error detection) - CODT
14. CHEcking Type- CHET
15. ADress DIagnosticd - ADDI
16. Data EcHO capability - DEHO
17. Over-Flow PRotection - OFPR
18. EXtension CApability - EXCA

6.8.3.2.Control Room - CR2

This hybrid solution combining digital and analog automation is typical of most of the system implemented in several process industries during the late 1970's. The analog part usually varies between 5 and 15% of all control loops and togetherwith the Redundant Control Units - RCUs forms the desired back up (respectively increased reliability) of the process control system. In such a way CR2 utilizes the advantages of dataway communications and of the new forms of operator interface, concentrating all information to monitor, control and optimize the processes in a physical space within easyreach of a seated operator.

Two (sometimew only one) redundant Monitoring Stations - MS with corresponding Analog Recorders - AR and TYPEr(s) (but without displaying process schematics and computing advanced control strategies) represents the most frequently used at present solution for installing digital automation systems in existing process industry mills.

Therefore, CR 2 can be characterized by the following capacity parameters and functional capabilities of only one Monitoring Station - MS.

1. Maximal Capacity of thew MS - MCMS in loops or points per one MS

2. Redundancy of MS - RMS in %

3. TYpe of CRT, with Number f COLors - NCOL, RAY Frequency - RAYF etc.

4. Number of Controlled Loops in the GEneral display - NCLGE

5. Number of Groups (Control Stations-CS) which can be Displayed - NGRD (NCSD)

6. Number of Controlled Loops in one GRoup - NCLGR.

7. Number of PArameters in a Control Loop display - NPACL

8. Number of Histories with different Times of consideration (horizones) - T in min,h) and Discrete (Averages) intervals -DT (AT) in (s,min,h)

9. Number of Trends - NT with corresponding T and DT.

10. Number of Control Loops recorded by one Analog Recorder -NCLAR the total number of which is NAR.

11. Number of Alarms in GEneral display NALGE

12. Number of Alarms in Group display - NALGR

13. Number of Alarms in SUmmary display - NALSU

14. Number of Loops and/or Cards in Diagnostics display - NLCD

15. Number of Specialized REports - NSRE

16. TYpe of PRocessor (CPU) TYPR(TYCPU)

17. TYpe of Auxiliary Storage Devices - TYASD

18. PROgramming Capabilities and Languages - PROCL

19. Number of Special and Advanced Control Algorithms - NSACA

20. Process Optimization Capabilities - POC

6.8.3.3.Control Room M - CRM

This solution in principal utilitizes all the additional advantages of modern digital automation systems, including:

1. Process schematics by graphic displays.

2. Scheduled and prescheduled logic operations

3. Extended processing and operating capabilities, such as monitoring capacity, number of histories, trends, dataway,etc.

4. Advanced calculations in order to permit the integration of different processes into a complex process or production unit

5. Optimization (on-or off-line) of the processes and/or the production unit.

6. Production scheduling and control as an element of production and energy management.

7. Realization of advanced strategies.

In addition to both Monitoring Stations - MS ,Analog Recorders -AR and TYPEr(s), the Control Room M - CRM, normally includes a so-called Dispatching Station - DS , realizing some or all of the additional advantages mentioned above. Therefore the DS can be partly characterized by the same 20 capacity parameters as have used for MS. It is however, necessary to include as positions 21-30 of the block CRM in Fig. 1 the following additional 10 characteristics:

1. Number of Data Ways (connected with the DS) -NDS

2. Number of Process Schematics (graphically displayed by the DS) - NPS

3. Production Scheduling and Control - PSC

4. Energy Scheduling and Control - ESC

5. Intermediate Storage Control- ISC

6. Scheduled Logs Capability - SLC

7. OPtimal control capability - OPC

8. ADaptive Control capability - ADC

9. Management Information System Fu.ctions - MISF.

10. Advanced Control Strategies - ACS

The block CRM can be further specified by the parameters given in the block CR1.

6.8.4. Practical Example

In the example, the functional capabilities and capacity parameters in nine units at the level "Stabilization control" (Fig.1) for illustrative purposes are compared:

1. Controller Station (CS), described in 2.2.

2. Redundant Control Unit (RCU) - 2.3.

3. General PIU (GPIU) - 2.4

4. Analog PIU - (APIU) -2.4

5. Multiplexing PIU - (MPIU) - 2.4

6. Standard PIU - (SPIU) - 2.4

7. Programmable Logic Unit (PLU) - 2.5

8. Dataway and Data Traffic Controller (DTC) - 3.1

9. Monitoring Station (MS) - 3.3

On the basis of sertain known experience with digital automation system implemented in different petrochemical units and complexes, the following seven distributed control systems are selected:

1. TDC - 3000 by HONEYWELL - HON

2. TELEPERM - M by SIEMENS - SIE

3. CONTRONIC by HARTMAN & BROUN - H&B

4. SPECTRUM by FOXBORO - FOX

5. DAMATIC by VALMET - VAL

6. PROVOX by FISHER - FIS

7. MOD - 300 by TAYLOR - TAY

This selection is made to give a maximal possible cover of different architectures satisfying the requirements of petrochemical industrial complexes.

All information about the units and systems, described above. which is based on the technical documentation of different producers, is summarized in a suggested useful form given in Table 1 (In this table only the results of CS,DTS,MS and Summary table are shown).

The abbrevations of the system in Table 1 are shown on the line "Producers". Symbols of capacity parameters and functional capabilities are listed in the column "Code".Values of parameters for every system are grouped into two columns:

1. Absolute Values - ABS

2. Relative Values - REL

In the column "ABS" values are in engineering units, whereas in the column "REL" these absolute values are related to the best values of the parameter considered.

For instance, let us discuss the unit Controller Station (CS) .For No 1 with "Code" NAI (NUMBER OF ANALOG INPUTS) the best value is 64 analog inputs for the system "DAMATIC", produced by VALMET. In such a way "ABS" is 64 and "REL"is 1. For TDC 3000(HON) ,number of analog inputs of the CS is 32 or "ABS" and "REL" is 32/64+0,5.

The ideal system then will have for all parameters relative values (REL) equal to 1 and the abstract ideal CS will have the absolute value (ABS) equal to 21. "ABS" for the CS of TDC 3000 (HON) system is 16,13 and then "REL" is 16,13 /21 = 0,77. By this solution the total absolute value of ideal abstract system is "ABS" = 118 (see Summary table from Table 1) and for example, the total "ABS"for TDC 3000 (HON) is 77,21. The "REL" value is equal to 77,21/118 = 0,68, which means that this system satisfies about 68 percent of all specific requirements to the ideal distributed control system suitable for petrochemical industrial complexes.

6.8.5. General Recommendations for the Distributed Control System

The recommended in the present report technical solutions for implementation of distributed microprocessor based control system can be successfully realized with the everyone of the compared above systems of different producers.

Whereas the wordwide tendency for implementation of digital system integrated with higher level computer functions "SYSTEMATICS" suggests to IPCL a modernization of the existing pneumatic instrumentation and control system using the devices and systems from one supplier. On the basis of the worldwide refferences for different petrochemical plants and complexes our suggestion for the system is as follow:

1. TDC 3000 - PRODUCED BY HONEYWELL

2. SPECTRUM - PRODUCED BY FOXBORO

3. DAMATIC - PRODUCED BY VALMET

When the final technical solutions mus be done we recommnded to have in mind technical, financial, support and etc. considerations.

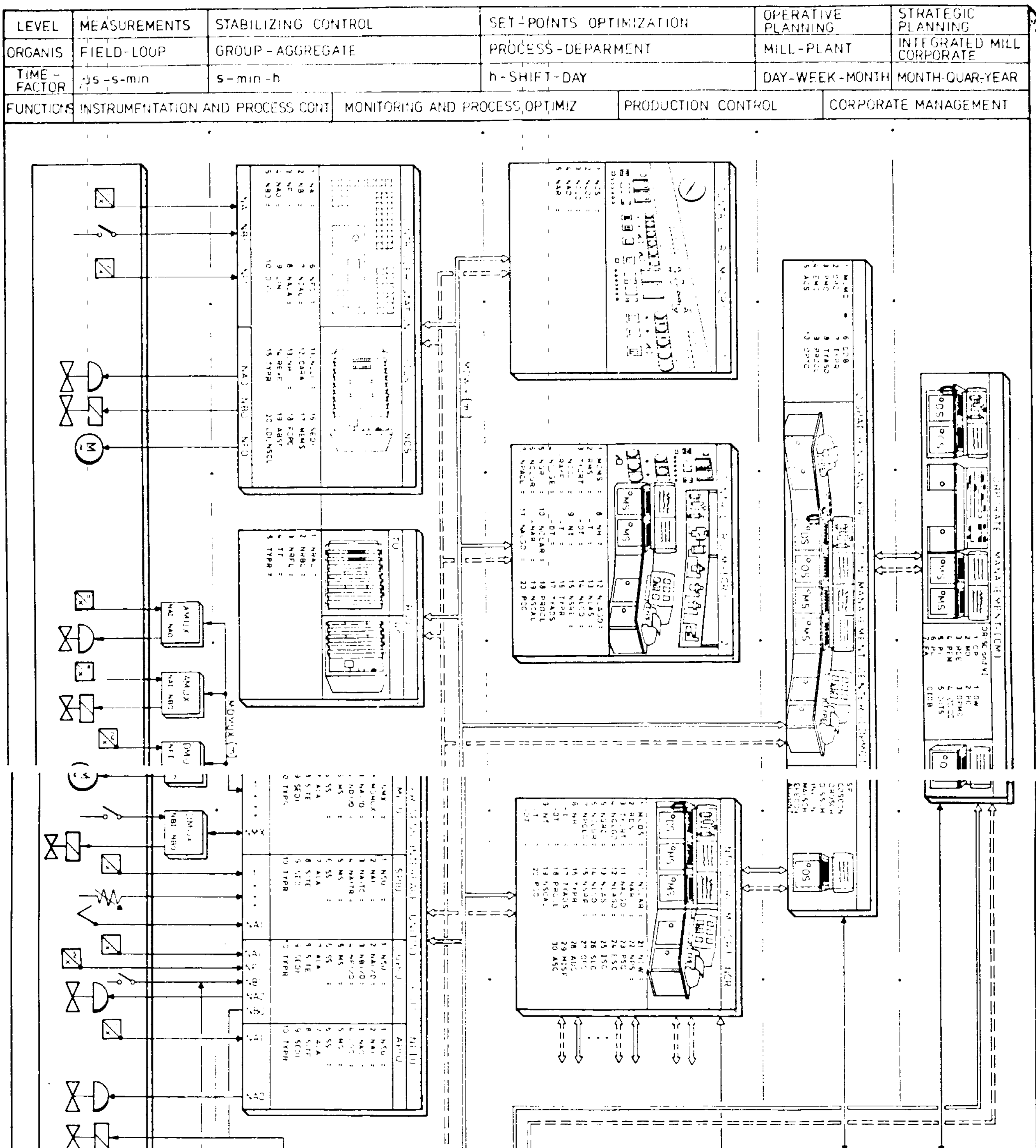

LEVEL
MEASUREMENTS
STABILIZING CONTROL
SET-POINTS OPTIMIZATION
OPERATIVE PLANNING
STRATEGIC PLANNING
ORGANIS
FIELD-LOUP
GROUP-AGGREGATE
PROCESS-DEPARMENT
MILL-PLANT
INTEGRATED MILL CORPORATE
TIME-FACTOR
s-min-h
h-SHIFT-DAY
DAY-WEEK-MONTH
MONTH-QUAR-YEAR
FUNCTIONS
INSTRUMENTATION AND PROCESS CONT
MONITORING AND PROCESS OPTIMIZ
PRODUCTION CONTROL
CORPORATE MANAGEMENT

Fig 1

ANALYSIS OF PRODUCTION MANAGEMENT SYSTEM

UNIT		CONTROLLER STATION - CS													
Producers		HON		SIE		H&B		FOX		VAL		FIS		TAY	
Parameters															
No	CODE	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL
1.	NAI	32	0,5	30	0,46	16	0,25	60	0,94	64	1	8	0,13	1	0,016
2.	NBI	128	0,13	45	0,19	128	0,53	240	1	128	0,53	0	0	1	0,004
3.	NFI	64	0,53	0	0	0	0	120	1	64	0,53	0	0	0	0
4.	NAO	16	0,27	30	0,5	12	0,2	60	1	24	0,38	8	0,13	1	0,016
5.	NBO	64	0,26	45	0,19	128	0,53	240	1	64	0,26	0	0	1	0,004
6.	NFO	0	0	0	0	0	0	120	1	64	0	0	0	0	0
7.	NCAL	64	1	40	0,57	46	0,66	30	0,43	48	0,7	23	0,33	10	0,14
8.	NALA	64	0,53	60	0,5	16	0,13	120	0,94	128	1	16	0,13	2	0,06
9.	LIN	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
10.	DIFI	YES	1	NO	0	YES	1	YES	1	YES	1	NO	0	NO	0
11.	NOLI	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
12.	CARA	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES-NO	0,5
13.	NH	8X16	1	variable	1	3x16	0,37	NO	0	4	0,03	NO	0	NO	0
14.	REVE	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
15.	TYPR	16 bit	1	16 bit	1	16 bit	1	16 bit	1	8bit	0 ,5	8bit	0,5	16 bit	1
16.	SEDI	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
17.	MEMS	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
18.	FCPCD	YES	1	YES	1	YES-NO	0,5	YES-NO	0,5	YES	1	NO	0	NO	0
19.	ABCT	YES	1	YES	1	YES	1	YES	1	YES	1	NO	1	NO	1
20.	LOI	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
21.	NSCL	128	0,64	94	0,47	200	1	60	0,3	32	0,16	1	0,005	1	0,005
TOTAL		21	16,13	21	13,88	21	14,17	21	18,11	21	15,62	21	8,10	21	7,70

TABLE 1a

ANALYSIS OF PRODUCTION MANAGEMENT SYSTEM															
UNIT		DATAWAY AND DATA TRAFFIC CONTROLLER - DTC													
Producers		HON		SIE		H&B		FOX		VAL		FIS		TAY	
Parameters															
No	CODE	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL
1.	NDW	3	0,1	1	0,03	1	0,03	10	0,3	7	0,22	1	0,3	1	0,03
2.	RDW	100%	1	100	1	100%	1	100%	1	100%	1	100	1	100	1
3.	MDW	2X4500	0,75	4000	0,67	4500	0,75	4500	0,75	2000	0,33	1500	0,75	305	0,05
4.	NVD	28	0,22	100	0,79	127	1	10	0,8	30	0,22	30	0,24	20	0,16
5.	NVC	63	0,2	100	0,3	127	0,39	100	0,3	98	0,3	30	0,1	20	0,06
6.	NPL	3	1	2	0,66	3	1	3	1	2	0,67	2	0,66	2	0,67
7.	NSC	3	0,1	0	0	0	0	10	0,3	7	0,22	1	0,03	0	0
8.	NAC	0	0	32	1	32	1	0	0	32	1	0	0	19	0,6
9.	CS	250kbud	0,25	250	0,25	1000	1	1000	1	250	0,25	200	0,2	9,6	0,02
10.	STAN	NONSTAN		RS-422C		RS-422C		RS 232		RS 232		NON STAN		RS 242	
1.	PROT	HDLC		HDLC		HDLC		SOLC		HDLC		HDLC		ASCII	
2.	MEST	31 bit		128b		16bit		VAR		16bit		16bit		16bit	
3.	CODT	BCH 5bit		LRC		CRC, LRC		CRC		CRC		LRC		LRC	
4.	CHEC	1984		4X9bit		1X64		32X100		7X16, 38x16		30X16		8X16bit	
5.	ADDI	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
6.	DEHO	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1	YES	1
7.	OFPR	NO	0	NO	0	NO	0	YES	1	NO	0	YES	1	NO	0
8.	EXCA	NO	0	NO	0	NO	0	NO	0	YES	1	YES	1	NO	0
TOTAL		13	5,62	13	6,70	13	8,17	13	7,73	13	6,22	13	6,51	13	4,58

Table 1b

ANALYSIS OF PRODUCTION MANAGEMENT SYSTEM

UNIT		MONITORING STATION - MS													
Producers		HON		SIE		H&B		FOX		VAL		FIS		TAY	
Parameters															
No	CODE	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL
1.	MCMS	1500	0,49	3072	1	900	0,29	1080	0,33	2048	0,66	1200	0,39	1024	0,33
2.	KMS	200%	1	200%	1	200%	1	0%	0	200%	1	200%	1	100%	0,5
3.	TYCRT	col	1	col	1	col	1	col	1	col	1	col	1	black	0,5
3a	NCOL	4	1	4	1	4	1	4	1	4	1	4	1	1	0,25
4.	NCLGE	288	0,5	256	0,44	576	1	64	0,11	64	0,11	240	0,42	238	0,41
5.	NGRD	150	0,39	384	1	150	0,39	150	0,39	150	0,39	100	0,26	32	0,08
6.	NCLGR	8	0,27	8	0,27	6	0,2	8	0,27	8	0,27	12	0,4	30	1
7.	NPACL	40	1	25	0,62	20	0,5	8	0,2	12	0,3	23	0,57	10	0,25
8.	NHT	960	1	672	0,7	300	0,31	64	0,07	384	0,4	720	0,75	256	0,27
9.	NT	9	1	9	1	7	0,78	8	0,89	5	0,55	5	0,55	7	0,78
9a	DT	3	1	0	0	0	0	0	0	3	1	0	0	0	0
10.	NCLAR	3	0,05	3	0,05	3	0,05	6	0,06	3	0,05	0	0	64	1
10a.	NAR	100	0,16	56	0,09	640	1	200	0,31	96	0,15	0	0	640	1
11.	NALGE	1152	1	512	0,44	1152	1	128	0,11	384	0,33	960	0,83	0	0
12.	NAGLR	100	1	32	0,32	64	0,64	32	0,32	96	0,96	0	0	0	0
13.	NALSU	600	0,71	512	0,85	576	0,68	64	0,08	576	0,68	800	0,95	512	0,61
14.	NLCD	640	1	0	0	0	0	0	0	8	0,01	0	0	0	0
15.	NSRE	9	0,9	10	1	10	1	6	0,6	10	1	8	0,8	4	0,4
16.	TYPR	16bit	1	16bit	1	16bit	1	16bit	1	8bit	0,5	16bit	1	16bit	1
17.	TYASD	cass flop.	1	flop-pyd.	1	cass	1	disket	1	flop-pyd.	1	flop-pyd.	1	flop-	1
18.	PROCL	YES	1	YES	1	NO	0	YES	1	NO	0	NO	0	NO	0
19.	NSACA	YES,6	1	YES,6	1	NO	0	NO	0	YES	1	NO	0	NO	0
20.	POC	YES	1	YES	1	NO	0	YES	1	NO	0	NO	0	NO	0
TOTAL		23	18,47	23	15,78	23	12,48	23	9,77	23	12,35	23	7,56	23	8,88

Table 1c

ANALYSIS OF PRODUCTION MANAGEMENT SYSTEMS SUMMARY TABLE															
Producers		HON		SIE		H&B		FOX		VAL		FIS		TAY	
Parameters															
No	CODE	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL	ABS	REL
1.	CS	16,13	0,47	13,88	0,66	14,17	0,68	18.11	0,87	15,62	0,74	8,10	0,39	7,70	0,37
2.	RCT	2,75	0,55	-	0	2,87	0,57	3,94	0,79	1,97	0,39	-	0	-	0
3.	PLT	8,10	0,90	2,76	0,31	3,39	0,38	-	0	4,06	0,45	-	0	-	0
4.	CRIT	8,95	0,69	9,71	0,75	8,37	0,64	8,36	0,64	8,00	0,61	6,48	0,50	7,25	0,56
5.	ARIT	7,60	0,76	8,75	0,87	9,11	0,91	8,35	0,83	8,43	0,84	-	0	-	0
6.	MRIT	8,11	0,62	8,32	0,64	-	0	-	0	-	0	1,20	0,55	8,61	0,66
7.	SPIT	6,54	0,59	7,63	0,69	7,08	0,64	9,25	0,84	6,21	0,56	-	0	-	0
8.	DTC	5,62	0,43	6,70	0,51	8,17	0,69	7,73	0,59	6,22	0,48	6,51	0,50	4,58	0,35
9.	MS	18,47	0,80	15,78	0,64	12,84	0,56	9,77	0,34	12,35	0,54	10,92	0,47	8,88	0,39
TOTAL		77,21	0,68	73,53	0,61	66,00	0,55	65,51	0,55	62,86	0,53	39,21	0,33	37,02	0,31

Table 1 d

6.9. Recommendation to the on-line analyzers

The presented list of implemented in IPCL on-line analyzers shows their quite wide variety by types and manufactorers. We suggest to decrease the type numbers especially for the universal instruments (excluding the specific type or analyzers like Melt Index Analyser, Catalyst Activity Analyzer, etc.).
In this field a good achievement is the orientation to such manifacturer as Beckman, which univerasl serie 6750 could become a basic one for all units under revamping. For the chromatographs in the Ethylene plant we suggest Siemens and Yokogawa products.
The unification is not so important for some simple analyzers (such as pH-meters).International practice confirms Electrofact, Foxboro and Kent products.

A question of present interest is the application of oxygenmeters with zirconic probes for flue gasses. We have good experience with WESTINGHOUSE type oxygen analyzers, because of the precise measurement and high reliability.

In any other cases of oxygen measurement in the gasses we are suggesting to use analyzers operating on paramagnetic principle and the common application of Magnos 2t and 5T of Hartman & Braun /FRG/

For measuring the microconcentration of the oxygen in the gasses we are recommending to apply electro-chemical analyzers of Hersh - England U.K.

The last series analyzers have common application in measuring tha gas humidity in reformer units, nitrogen, H2, instrument-air, etc.

We suggest wide application of infrared analyzers especially where some components could be separated from the mixture by means of optical filters. In this case a frequent calculation is required.The main supplier can be Hartman & Broun - Uras model.

The measurements requirement for creating the relevant balance or for two-component mixtures and pure products are often made by densitymeters supplied from ADAR-USA; FOXBORO-USA, and POLUX- FRG, especially for gasses.

Liquid density meters based on stream-lined balance should be avoided because of hard particles settlement on the measuring tube. The same problem stays for float-type densitymeters.

Considerably less problems create the analyzers determining the quality of the demineralized water feeded to the power station or applied for chemical purpose. In this case the important property to follow is the salt content (hardness, conductivity, oxygen, etc) . Most often we suggest the application of German or Soviet Union analyzers, which are easily maintaned and have a steady operation.

The most common analyzers of the first group are the one of "PHASE SEPARATION"-UK and SERES - FRANCE determining the content of total organics in the waste waters: phenol, copper, amonia and other dangerous reagents, influencing the biological mass in the water treatement units. The equipment supplied for that purpose is controlled by aothorized department. We recommend all plants in IPCL Complex to be with systems for explosive gas indication on the most dangerous points.According to standard requirements the alarm should appear at 20% from the lower explosion limit as well as preliminary alarm at 15%. The most popular application have the analyzers supplied by SIEGER - FRG and by STI & IKAR - FRANCE.

The second company is the supplier of the analyzers giving an alarm at given ACN,H2S,SO2 etc.concentration, dangerous for humans health.

Our recommendation is to apply the Model 58 Automatic Silica Analyzer produced by EIL ANALITICAL INSTRUMENTS OF KENT INDUSTRIAL MEASUREMENTS GROUP - U.K.

The main features of this analyzer are :

- high sensitivity
- automatic zeroing
- aprobation issued by GEGB for Model 5827

Multi-stream versions are also available. In boiler system where sodium phosphate is used for pH regulation of presence of phosphate is possible.The use of Model 5835 silica analyzers of EIL is recommended.There is an additional reagent to inhibit the effects of phosphate.

The operating principle of analyzers 58 series is based on the molybdenum blue method as in Model 8065 Silica Monitc .
The difference between 8065 and 5827 is in the elimination of the participation of molybdic acid and the gas or air bublles in the cuvette as well as the presence of automatic compensating unit. The accuracy and repeatability are higher.

The problems concerning the correct choise of a process (on-line) analyzers, their installation and commissionning and especially their maintenance, require special care and consideration.Plant process engineers have the leading role in this because the proper operation of each analyzer depends in reasonable instance by their exactingness and help.

6.10. On-line conversion implementation of a distributed digital control system

On-line conversion means to carry-out the transfer from existing pneumatic analog control system in the plant to the new digital distributed control system without interruption of the plant operations. This is critical job and should be well planned and organized before starting the actual implementation work.

The following steps explain the procedure to be followed to carry-out this work in a safe and reliable way.

1. Planning and Organization

It is essential to define the responsibilities of each party sharing the implementation of conversion. Construction, Maintenance (Mechanical, Electrical and Instrumentation) Operations and engineering services Department should be directly involved and coordination in between these departmens is needed. A conversion group should be formed from all these departments to be responsible for conversion work and to follow a predetermined procedure.

2. Conversion Group

This group shall be responsible to carry-out the following work:

- Set-up a detailed conversion schedule.
- Finalize installation task list
- Carry-out calibration and test of the equipment
- Prepate detail drawings verify connections;
- prepare all installation materials;

During Conversion:
- Izolate the loops;
- Install all new electronic type field equipment (e.g. transmitter field cables, mount P/I and I/P converters etc)
- connect field wiring in junction boxes and termination panels of the new control system;
- carry-out calibration and function tests;
- carry out final acceptance tests;
- up-date drawings as built

In order to carry out the above task the conversion group should be well organized and coordinated.

Typical e..ample of the organization chart for this group is shown on Fig 6.8.1.

Speed of conversion

During conversion part of the plant process units shall be controlled from the old panels/old control room/while other part shall be controlled from the new distributed computer/digital/control system.This position create operational inconvenience and is not recommended to last for a long period. It is therefore necessary to carry-out the conversion in the shortes possible time. It is estimated that on-line conversion for the whole process units/sections/ can be carried-out in a period of 60 to 90 days / 2-3 months/

Implementation Procedure

a/ The conversion team hold daily meeting to be converted and define the importance of the loops and if they are critical and at the same time check about the results of conversion for the previous day.

b/Preparation in field of necessary installation material, drawing portable communication radio units and sound power telephones;

c/Check with existing control room panel operator and new control system console operator if they are ready to start the work.

d/Change the control loop to manual mode and take it out of service and start wiring change in the field junction boxes or on the trermination panels of the new system;

e/ Functional test of loop by new digital system to be done by service engineering and resetting the variables to be realized

f/Give the loop to new control console operator who starts to work with the loop after conversion

g/Update loop drawings as built and approve the change over by instrument engineer, electrical engineer and operations section head;

h/Review the drawings by project engineer and modify the original drawingd for records.

ORGANIZATION OF CONVERSION GROUP

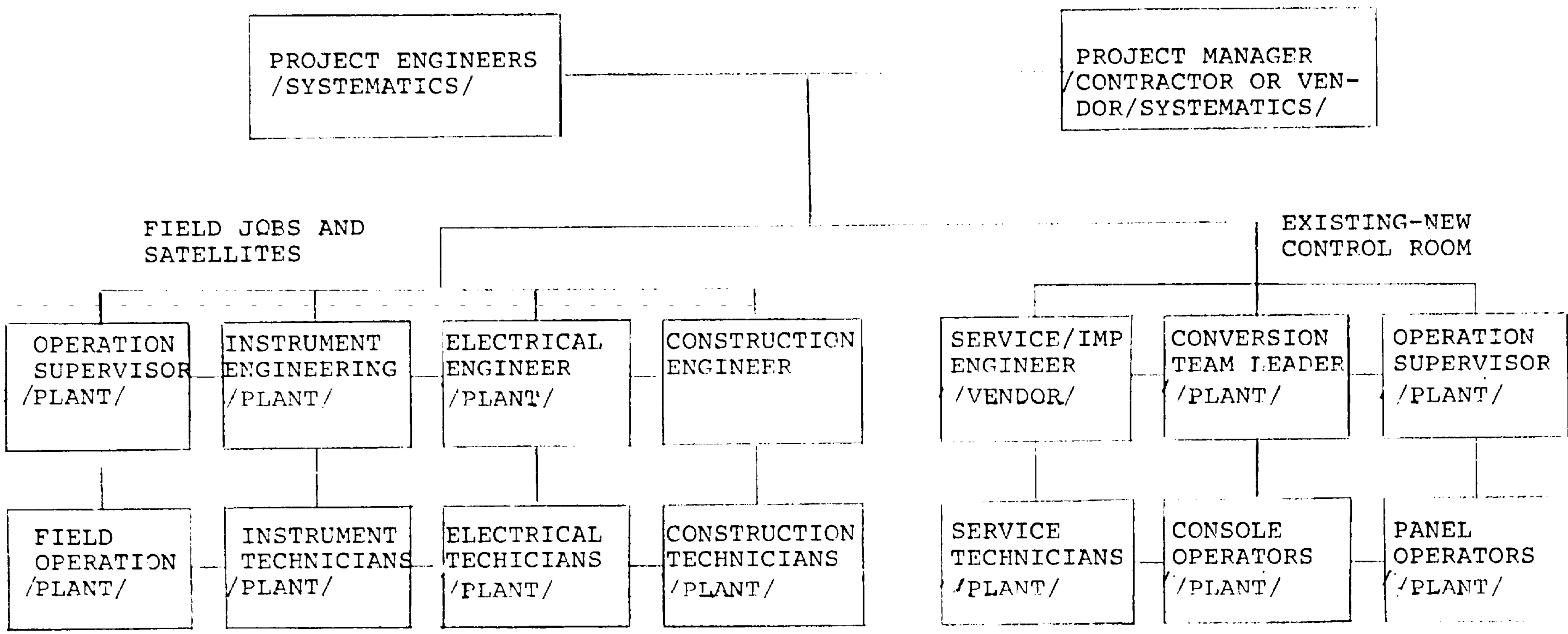


Fig 6.8.1.