# CHEMICAL ENGINEERING DESIGN PROJECT
A Case Study Approach

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CHEMICAL ENGINEERING DESIGN PROJECT
A Case Study Approach

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Introduction to the Series

The subject matter of chemical engineering covers a very wide spectrum of learning and the number of subject areas encompassed in both undergraduate and graduate courses is inevitably increasing each year. This wide variety of subjects makes it difficult to cover the whole subject matter of chemical engineering in a single book. The present series is therefore planned as a number of books covering areas of chemical engineering which, although important, are not treated at any length in graduate and postgraduate standard texts. Additionally, the series will incorporate recent research material which has reached the stage where an overall survey is appropriate, and where sufficient information is available to merit publication in book form for the benefit of the profession as a whole.

Inevitably, with a series such as this, constant revision is necessary if the value of the texts for both teaching and research purposes is to be maintained. I would be grateful to individuals for criticisms and for suggestions for future editions.

R. HUGHES

Acknowledgements

I About This Book — The Case Study Approach

This book provides a case study approach for the teaching and appreciation of the work involved in a chemical engineering design project. Ail undergraduate chemical engineering students are required to perform a design project, usually in the final year of the course. It may be the last piece of work that a student completes (after all other subjects have been examined) prior to graduation, carried out over a period of between 6 to 10 weeks (depending upon departmental policy). Alternatively, the design project may be performed during the entire final year of study. No doubt, variations on these alternatives occur in certain   faculties.

Courses that are accredited by the *Institution of Chemical Engineers (IChemE)*   UK, must include a design project unit conforming to their specifications (see Section IV). All UK chemical engineering degree courses are accredited by the IChemE; courses in territories having strong historical links with Great Britain, e.g. Africa, Australia, West Indies, etc., also usually aim for IChemE recognition.

In the United States, most engineering courses are accredited by the *Accreditation Board of Engineering and Technology (ABET),* of which the *American Institute of Chemical Engineers (AIChE)* was a founding organisation. The requirements of the AIChE regarding the teaching of chemical engineering design and the design project are different from those laid down by the IChemE, although all US accredited courses are expected to include some form of design project work to be performed by their students. Only graduates from courses accredited by the IChemE are admitted to professional membership of that institution (or graduates from non-accredited courses who can subsequently fultil the IChemE   requirements).

This book is intended to provide guidance specifically to those students who are enrolled in IChemE accredited courses, and are about to commence the design project. Those same students will also find this book useful when they are studying earlier units in Plant and Process Design; reference to this text will illustrate how certain topics are to be applied during the design project. However, other students in courses not accredited by the IChemE (specifically in the USA) should also find this text useful when studying similar course units.

The approach adopted here is to provide brief notes and references for a wide range of topics to be considered in the design project. Case study material concerning *The Manufacture of Nitric Acid* is presented, and

illustrates what is required in the design project. The case study material is adapted from the design project performed by D.W. Johnston at Cur-tin University of Technology, Perth, Western Australia, in 1986. This project was awarded the CHEMECA Design Prize for the best Australian university design project in 1986, and the CHEMECA medal was presented at the fifteenth Australian Chemical Engineering Conference. The Curtin University chemical engineering course is accredited by the IChemE and the design projects performed at the university conform to the Institution requirements.

A coherent view of the design project requirements is obtained by using one typical design study to provide all the case study material for the text. Some appendices relating to background information and the documentation of detailed calculations, e.g. mass and heat balances, have been omitted in order to limit this book to a reasonable size. The basis of all calculations are included and students should be able to check the validity of the stated results if so desired. The authors would be grateful for details of any errors (of calculation or logic) which the reader may discover. Design projects are seldom (if ever) perfect and this book, and the case study material, is no exception.

It was decided that a realistic appreciation of the stages in a design project, and the sequence of tasks that the student performs, would be obtained by including the descriptive notes in "Times" typeface 'between' the case study material, which appears in "sans serif" typeface. This was in preference to presenting all the notes followed by the 'typical' student design project.

The aspects of the design that were considered in this project are more comprehensive than those required by the IChemE in their design problem for external students (see IV The Scope of the Design Project and Appendix K). Topics such as market appraisal, site selection, plant layout, etc., are considered here. The detailed requirements and particular emphasis on certain topics, e.g. control and instrumentation, economic analysis, HAZOP, etc., often depends upon the experience and philosophy of the supervisor and departmental policy. However, we feel that the aspects of design presented in this book cover a wide and comprehensive range of possible topics, although it is expected that most lecturers would prefer a more detailed coverage in certain areas. Ultimately this book is intended to provide guidance *to the student,* not to be a complete text on all aspects of plant design or an alternative to Perry's Handbook.

II Advice to the Student

As a student faced with a chemical engineering design project, you probably have two immediate feelings. First, excitement at finally beginning the project that has been talked about so often in your department, This excitement is enhanced by finally being able to undertake a piece of work that is both challenging and satisfying, and which will enable you to contribute your own ideas. After so much formal teaching it provides the opportunity to create a finished product that is truly your own work.

The second feeling will probably be apprehension about how this daunting task is to be achieved. Will you be able to do what is required? Will you be told what is expected? Do you already possess the necessary knowledge to complete the project? Other similar questions probably come into your mind. The simple answer is that design projects have been performed by students in your department since the course began, very few students fail this unit and most produce at least a satisfactory project, and often a better than expected report. Previous students have started the project with the same basic knowledge that you possess and, by asking the same questions, they have completed it using the same resources available to you.

Information, assistance and advice should be provided by the project supervisor. Do not stand in awe of this person, ask what you want/need to know, ask for guidance, and persist until you know what is expected. However, understand that a supervisor only provides guidance, and will not (and should not) perform major parts of the design project for you. This is the time for you to show initiative, and to impress the lecturers with your knowledge of chemical engineering and your own ability to solve problems.

My main advice to the student undertaking a chemical engineering design project is: 'don't work in a vacuum!'. By this I mean obtain information and help from as many sources as you can find. Do not assume that you alone can, or should, complete this project unaided. Talk to the project supervisor, other lecturers in your department, lecturers in other departments and at other universities and colleges, other students, technicians, librarians, professional engineers, research students, officers of the professional institutions, etc. Some of these people may not be able to help, or may not want to; however, it is usually possible to find some helpful and sympathetic persons who can offer

advice. The most obvious people to approach are the design project supervisor, other lecturers in the same department, and other chemical engineering students (your peers and research students).

Valuable information can often be obtained from chemical/chemical engineering companies (at home and abroad). The information provided may range from descriptive promotional material, press releases, published technical papers, patents and company data sheets, to detailed advice and information from company employees. Some of this information, especially the latter, may be provided on a confidential basis. A company may refuse to disclose any information, particularly for new products or processes benetitting from recent technological advances. The older processes used to produce 'traditional' chemicals are usually well documented in the technical literature. Information concerning new project proposals may have been deposited with government departments, particularly concerning environmental impact regulations. Some of this information may be available to the public and can provide valuable data for feasibility studies. It is usually necessary to plan well in advance to obtain company information, particularly from overseas.

The completed project should be a testimonial to the student's abilities as a chemical engineer, soon to be employed in industry and eventually to become a recognised professional engineer. The work should demonstrate a breadth of knowledge relating to chemical engineering in general, and an appropriate depth of knowledge in relation to particular chemical engineering design problems that have been tackled. The project should be the student's own work, and must represent an achievement in terms of the application of chemical engineering principles.

In my experience, the 'best' projects are usually produced by those students who are widely read and are interested not only in chemical engineering but also in a wide range of subjects. In this case, 'best' means a competent or satisfactory design and a project that includes consideration of a wide range of relevant factors, not only the technical aspects of equipment design. However, I find that most students, even those with a previously poor academic record, are inspired by the prospect of being able to work on a reasonably open problem with the opportunity to produce work that is truly their own. Students in general tend to rise to the challenge rather than merely engage in 'going through the motions'.

III To the Lecturer

This book is not intended to be a recommended text for a taught unit in Plant and Process Design: there are several books which already satisfy those requirements, although it provides useful background reading for that subject. This textbook helps the student performing the chemical engineering design project. It provides only essential notes for a range of associated topics, and the case study material (taken from an actual student design) provides a detailed example of the contents and format of the project report.

Many students are overwhelmed, apprehensive and unsure how to proceed when faced with the design project. It is unlike any assignment they have previously been given and represents a true test of their abilities and initiative. However, too often students spend this initial phase wondering what is actually required and viewing past students' projects, which serve merely to emphasise the enormity of the task ahead rather than provide a detailed analysis of what is needed and a plan of action. This book should satisfy the students' need for guidance, and provide a useful case study example as the project proceeds through each stage.

The case study included is just one particular example of the way in which the project can be performed and presented. Each department (and supervisor) will define their own requirements, but our approach and presentation should not be too different. The emphasis in our course at Curtin University is for effective communications. In the design project report this means presenting only essential information for immediate attention and confining all additional information and numerical calculations to appendices. Summaries are required at the beginning of each sub-section and as an introduction to each of the two major parts of the report.

In this book we also present the design project in two parts. Part I describes the Preliminary Design related to aspects of the Technical and Economic Feasibility Study of the project. During this stage of the project it is still possible to change the earlier major decisions such as production rate, process route, etc., if certain factors indicate particularly adverse conditions or a more economic alternative. The feasibility study should make recommendations such that the detailed equipment design can be performed in Part II. Students sometimes assume that the design is (almost) wholly concerned with the design of

equipment (i.e. Part II). However, without a thorough feasibility study to precede these designs, the project becomes more of an academic teaching (rather than learning) exercise.

Part II contains the design of a major item of equipment (in this case study, it is a sieve-tray absorption column (Chapter 9)), including the mechanical design, fabrication, materials specification, detailed engineering drawing, HAZOP study, control scheme and associated instrumentation. In summary, as complete and professional a design as possible is presented within the time available, while recognising the student's experience and abilities. The design of a second unit is presented in less detail, shown as a steam superheater (Chapter 10) in this case study. Part II also includes the full specification for a particular pump or compressor within the plant, including selection of an actual pump, from a manufacturer's catalogue, corresponding to the required operating characteristics. Part II concludes with the design of a pressure vessel, such as a storage tank, reactor shell, etc., to be designed in accordance with an appropriate pressure vessel code or standard. The pressure vessel design may be included in the design of one of the two specified items of equipment. The pump specification and selection and the pressure vessel design are included because they are common tasks given to young graduate engineers in industry, and they emphasise a practical dimension of the project.

In summary, the design project consists of a detailed technical and economic feasibility study for the process, followed by the detailed design of selected plant items and associated equipment. The production rate, selling price, etc., are determined from a detailed market analysis and economic forecasts. An appropriate process route is selected, followed by the site selection, plant layout, mass and energy balances, etc., as outlined in the contents list for Part I.

In some cases, the student is required to accept decisions made by the supervisor, e.g. selection of an older process for which design data is readily available in preference to a newer (secret) process, or choice of production capacity assuming future export markets in order to design a plant of significant size which is economically feasible. Although these choices may mean that the design no longer represents the optimum or 'best' design possible, the experience obtained by performing the project should not be diminished.

IV The Scope of Design Projects

Each year, the IChemE set a design project for external candidates. A copy of the detailed regulations is available, and also Notes for *the Presentation of Drawings* with an accompanying example of a process flow diagram. A list of the design projects set by the Institution from 1959 to 1986 is included in Appendix K. Full details are also given for the production of nitric acid problem set in 1980. More details of selected projects can be found in Coulson and Richardson (eds): *Chemical Engineering,* Volume 6 (1983; Appendix G, pp.795~820). Copies of the information provided with particular projects can be obtained from the IChemE (for a small charge).

Students at Curtin University are provided with a set of guidelines for the design project, including requirements for the oral presentations, and a booklet: *Presentation of Literature Surveys.* Interested parties can obtain copies of this material directly from Dr. Martyn Ray at Curtin University.

V Effective Communications

Written communications need to be effective, i.e. convey the intended message in a clear and concise manner. In order to achieve this objective, it is necessary to consider both the audience that will receive the information (and act upon it) and the nature of the information itself. In some situations a formal, fully detailed report is required; however, quite often a condensed form of communication (e.g. memorandum) is satisfactory.

Peters and Waterman (1982) identified several factors that were common to successful American companies. One of these factors was the implementation of a system of effective communications within an organisation. Two of the most successful companies, United Technologies and Procter & Gamble, required that all communications were in the form of a 'mini-memo' of one page *maximum* length.

In some chemical engineering departments, the length of student design projects tends to increase each year or to have stabilized at a rather voluminous 'norm'. Students refer to previous projects and usually assume that their length is acceptable *and* required. Quite often student projects are unnecessarily lengthy and much of the 'extra' information is attributable to other sources, e.g. Perry (1984), Kirk-Othmer (1978-84) etc., and could be replaced by an appropriate reference.

We believe that all student projects, including the design project, should contain only *necessary* information. Extensive background information for a project should be reviewed, summarised and referenced, whereas only new mathematical developments and relevant design equations should be included and referenced to the original source. Essential information should be included in the main body of the report and all additional information, data, calculations, etc., presented in appendices. The design project report should be presented so that it can be assessed by someone with a background in chemical engineering, but without any particular knowledge of the chosen process. The following features should be included in the written report to facilitate an assessment of the proposed chemical plant design.

(a) A one page summary at the beginning of the project detailing the project specification, the work performed, major decisions, conclusions, etc. This summary includes both Parts I and II.

(b) A one page summary for each of Parts I and II, to be included at the beginning of the relevant part of the report.

(c) A summary at the beginning of each chapter or major section of the report (or for a particularly significant topic).

(d) Brief conclusions at the end of each chapter or major section.

(e) Information that is not essential for an assessment of the project (but which provides useful/necessary background data) is included in appendices. Company literature, materials specifications, trade statistics, etc., are all presented in appendices, whereas conclusions drawn from this information are presented and discussed in the report itself. Calculations relating to the mass and energy balances are also detailed in an appendix, but the basis of all calculations and the results of these balances are presented as ledger balances within the report.

(f) Reference rather than reproduce ─ the use of appropriate referencing rather than reproducing large sections of readily available information.

(g) Guidelines should be given for the expected length of the report and for the design sections contained in Part II. These guidelines should refer only to the main body of the report; appendices can be as long as is required (within reason!).

The important principle is for clear and concise presentation of the design project report. This approach should make the marking and assessment as easy as possible, and the report should truly reflect the student's own work.

Reference:

Peters, T.J., and Waterman, R.H., *In Search of Excellence: Lessonsfrom America's Best-Run Companies,* Harper and Row, New York (1982).

VI Comments on the Case Study Approach

Chemical engineering students usually undertake a major study concerned with the design of a chemical plant in the final year of their undergraduate course. Such a study requires not only a thorough knowledge and understanding of all the chemical engineering subjects taught previously in the course, but also a wider appreciation of the restraints that are placed upon an industrial design, e.g. time, economics, safety, etc.

Although design can be taught by a traditional lecturing approach like any other topic, the graduating engineer will only become a 'good' designer if he/she:

(a) can apply the basic knowledge of chemical engineering;

(b) understands the broad constraints placed on chemical plant design, e.g. economics, environmental, social, etc.;

(c) is widely read, thinks about the ideas encountered, and uses the knowledge and ideas in a design study.

In terms of personal qualities, the student should be:

(i)   enthusiastic;
(ii) positive;
(iii) realistic;
(iv) self-motivated;
(v)   a problem-solver;
(vi) an accurate, careful and logical worker;
(vii) superhuman!

Using these notes and the case study material, and the books and papers published on plant design, the student and the engineer must *apply* what is known in order to produce a good design. Consider the notes included in each section of this book as a useful reference source only (or a bibliography of essential reading), not as a condensed version of everything there is to know or study!

The Case Study — Summary for the Completed Project

The results of the design project for the commercial production of nitric acid are presented. The project has been performed in two stages. The first part concerns the feasibility of the project, and the second part presents the detailed equipment designs.

From the investigation into project feasibility, it is proposed to construct a plant that will deliver 280 tonnes per day of 60%(wt) nitric acid. This capacity is based on 8000 hours of operation per year, i.e. 330 days. It is envisaged that this nitric acid production facility will be centred within a larger chemical complex to be located in the Bunbury region of Western Australia. Other plants on this site will include an ammonia plant and an ammonium nitrate plant. Approximately 70% of the product acid will be consumed *in situ* for the production of crystalline ammonium nitrate. The remaining acid will be available to exploit the neighbouring South-east Asian export market.

The process chosen for the nitric acid plant is the 'single-pressure' process based on the technology developed by C & I Girdler.

Part II of the project concerns the design of two main plant units (the NO, absorption column and the steam superheater), a pump to deliver the 'red' product acid from the absorption column to the bleaching column, and finally a product storage tank.

Absorption of the nitrogen oxide components (NO,) in the process gas stream is conducted in a sieve tray-type absorption column. This tower contains 59 sieve trays, of which the top 45 trays contain herringbone-type cooling coils to remove heat of reaction/dilution and maintain low absorption temperatures.

The steam superheater is a clamp ring-type, internal floating- head, shell-and-tube heat exchanger. It can produce up to 5775 kg/h of steam at 300°C and 4000 kPa.

A single-stage, single-suction, centrifugal pump is recommended to deliver 'red' product nitric acid from the base of the absorption tower to the product bleaching column.

The proposed nitric acid storage tank will provide product storage capacity for one week in the event of a plant shutdown in the adjacent ammonium nitrate facility. The tank has a capacity of 1950 m3 (representing 1450 tonnes of product acid).

The project was performed and written in two distinct parts. Some minor changes to the first part were necessary as more detailed design information became available. A summary of these amendments is presented at the beginning of Part II.

PART I

Preliminary Design —
Technical and Economic Feasibility

CHAPTER 1
The Design Problem

1.1 Initial Considerations and Specification

A CHEMICAL engineering design project does not follow a set of standard steps like many undergraduate teaching problems, nor does it have a single 'correct' solution. The considerations in a design project are many and varied. The solution that is finally accepted is (usually) the 'better' solution (often based upon economic considerations) from several alternatives. The important feature of a design study is that decisions must be made at every stage, and compromises are frequently required.

The following six steps in the design of a chemical process have been identified (see Ulrich, 1984):

1. Conception and definition
2. Flowsheet development
3. Design of equipment
4. Economic analysis
5. Optimization
6. Reporting

The student would benefit by reading the introductory chapters in one of the popular texts on plant and process design, (e.g. Peters and Timmerhaus, 1980; Ulrich, 1984; Baasel, 1976).

The first step in a design project is to identify all the relevant information that is available. The second step is to identify the information required, decided after initial consideration of the problem. Two possibilities exist:

(a) not enough information is available — a search is required (see Appendix L for ideas regarding sources of information);

(b) too much information is available — the task is to assess the reliability of conflicting information.

The possibility that all the required information is available (and sufficient) is remote (at least outside university teaching departments!). *Remember:* Published information is not necessarily correct!

This assessment and definition stage is often either rushed, overlooked or postponed by eager students, such an approach usually leads to wasted time and effort later in the design. Make it a rule to know what you have and where you are going, rather than simply *thinking* that you know.

Undergraduate design projects are often well defined, although this need not be the case. The design project as set by the *IChemE* (UK) is for the design of a process for the production of a particular chemical at a given production rate, and the design of particular items of equipment. In universities, the chemical(s) to be produced is(are) specified and the following information may be given (depending upon the philosophy of the supervisor):

   (a)   production rate (plant throughput);
   (b)   purity of final product;
   (c)   raw materials to be used/available;
   (d)   utilities available;
   (e)   site location;
   (f)   expected markets;
   (g)   the process route.

Some or all of this information may be expected to be obtained by the student as part of the design project. Industrial process designs are usually (although not always) clearly defined   'this is what we want, this is what we know, now decide how to achieve it (and make a profit)'.

Action.   *Identify useful sources of information and their location. Obtain personal copies of essential/useful information. Initiate a filing and reference system for useful information (preferably on computer disk). Prepare a complete design spectjication. Identifi the essential information that is available. Identify the information that is required. Obtain all necessary information.*

The Case Study — Summary for Part I

Feasibility Study and Initial Design Considerations

It is proposed to construct a plant that will deliver 280 tonnes per day

of 60%(wt) nitric acid. This capacity is based on 8000 hours of operation per year, i.e. 330 days. It is envisaged that this nitric acid production facility will be accommodated within a larger chemical complex to be located in the Bunbury region of Western Australia. Other plants on this site include an ammonia plant and an ammonium nitrate plant. Approximately 70% of the product acid will be consumed *in* situ for the production of crystalline ammonium nitrate. The remaining acid will be available to exploit the neighbouring South-East Asian export market.

The process chosen for the nitric acid plant is the 'single pressure' process based on the technology developed by C & I Girdler.

Part I of this project covers the major aspects of project feasibility and process description. Major design work is contained in Part II.

1.2 Case Study- Defining the Problem and Background Information

Summary

Nitric acid is a strong acid and a powerful oxidizing agent with enormous possibilities for applications in the chemical processing industry. It has commercial uses as a nitrating agent, oxidizing agent, solvent, activating agent, catalyst and hydrolyzing agent. In relation to world production, approximately 65% of all nitric acid produced is used for the production of ammonium nitrate (specifically for fertilizer manufacture).

Nitric acid is now produced commercially using the stepwise, catalytic oxidation of ammonia with air, to obtain nitrogen monoxide and nitrogen dioxide. These nitrogen oxides are subsequently absorbed in water to yield between 50% and 68% strength nitric acid by weight. For applications requiring higher strengths, several methods of concentrating the acid are used. The traditional methods are by:

(a)   extractive distillation with dehydrating agents such as sulphuric acid or magnesium nitrate;

(b)   reaction with additional nitrogen oxides.

The latter technique has the greatest application in industry.

The chemistry of ammonia oxidation is remarkably simple with only six main reactions that need to be considered. There are also

seven side reactions appropriate to this process, but for most design purposes they have little bearing on the results.

References Used

General Introduction: G6.
Process Technology (PT): 1,4,5,6,12,13.

1.2.1   Introduction

The initial design problem is to determine whether: 'It is both *economically and technically feasible to establish a facility to produce nitric acid in Western Australia'.*
   This is a diverse and complex undertaking that necessitates a full investigation into the uses, properties, market, process technology, and production economics, associated with this particular chemical. Having considered these aspects and several others, an appropriate plant to fulfil the assessed market requirements is sized and specified accordingly.

1.2.2 Properties and Uses

Nitric acid is a colourless   liquid at room temperature and atmospheric pressure. It is soluble in water in all proportions and there is a release of heat of solution upon dilution. This solubility has tended to shape the process methods for commercial nitric acid manufacture. It is a strong acid that almost completely ionizes when in dilute solution. It is also a powerful oxidizing agent with the ability to passivate some metals such as iron and aluminium. A compilation of many of the physical and chemical properties of nitric acid is presented in Table A.1 of Appendix A. Arguably the most important physical property of nitric acid is its azeotropic point, this influences the techniques associated with strong acid production. The constant-boiling mixture occurs at 121.9°C, for a concentration of 68.4%(wt) acid at atmospheric pressure.
   Nitric acid has enormously diverse applications in the chemical industry. It has commercial uses as a nitrating agent, oxidizing agent,

solvent, activating agent, catalyst and hydrolyzing agent. The most important use is undoubtedly in the production of ammonium nitrate for the fertilizer and explosives industries, which accounts for approximately 65% of the world production of nitric acid.

Nitric acid has a number of other industrial applications. It is used for pickling stainless steels, steel refining, and in the manufacture of dyes, plastics and synthetic fibres. Most of the methods used for the recovery of uranium, such **as** ion exchange and solvent extraction, use nitric acid. A full breakdown of the uses and applications of nitric acid is included in Ref. G6.

An important point is that for most uses concerned with chemical production, the acid must be concentrated above its azeotropic point to greater than 95%(wt). Conversely, the commercial manufacture of ammonium nitrate uses nitric acid below its azeotropic point in the range 50-65%(wt). If the stronger chemical grade is to be produced, additional process equipment appropriate to super-azeotropic distillation is required.

There is a potential health hazard when handling, and operating with, nitric acid. Nitric acid is a corrosive liquid that penetrates and destroys the skin and internal tissues. Contact can cause severe burns. The acid is a potential hazard, the various nitrogen oxides present as product intermediates in the process are also toxic. An assessment of the health risk must be fundamental to the design of any process. Further consideration and recommendations for the operating health risk and environmental impact of the plant are presented in Section 5.4.

1.2.3 The Evolution of Nitric Acid Production Processes

Until the beginning of the 20th century, nitric acid was prepared commercially by reacting sulphuric acid with either potassium nitrate (saltpetre) or with sodium nitrate (Chile saltpetre or nitre). Up to four tonnes of the two ingredients were placed into large retorts and heated over a furnace. The volatile product vapourized and was collected for distillation. An acid of between 93-95%(wt) was produced.

In 1903 the electric-arc furnace superseded this primitive original technique. In the arc process, nitric acid was produced directly from nitrogen and oxygen by passing air through an electric-arc furnace.

Although the process benefitted from an inexhaustible supply of free feed material (air), the power consumption for the arc furnace soon became cost prohibitive.

Researchers returned to the oxidation of ammonia in air, (recorded as early as 1798) in an effort to improve production economics. In 1901 Wilhelm Ostwald had first achieved the catalytic oxidation of ammonia over a platinum catalyst. The gaseous nitrogen oxides produced could be easily cooled and dissolved in water to produce a solution of nitric acid. This achievement began the search for an economic process route. By 1908 the first commercial facility for production of nitric acid, using this new catalytic oxidation process, was commissioned near Bochum in Germany. The Haber-Bosch ammonia synthesis process came into operation in 1913, leading to the continued development and assured future of the ammonia oxidation process for the production of nitric acid.

During World War 1, the intense demand for explosives and synthetic dyestuffs created an expansion of the nitric acid industry. Many new plants were constructed, all of which employed the ammonia oxidation process. This increased demand served as the impetus for several breakthroughs in process technology. These included:

(a) The development of chrome-steel alloys for tower construction, replacing the heavy stoneware and acid-proof bricks. This enabled process pressures above atmospheric levels to be used.

(b) The improved design of feed preheaters enabled higher process temperatures to be attained. Higher temperatures improved the yields and capacities, and also reduced equipment requirements.

(c) Early developments in automatic process control improved process performance and reduced labour requirements.

All of these factors helped to improve the process efficiency. The increasing availability of ammonia reduced processing costs still further.

In the late 1920's the development of stainless steels enabled manufacturers to use higher operating pressures. The increase in yield and lower capital requirements easily justified the use of high-pressure operation despite increased ammonia consumption.

The introduction of higher pressure processes resulted in a divergence of operating technique within the industry. The United States producers opted for a high-pressure system, using a constant

high pressure throughout the process. The European manufacturers opted for a split-pressure system. This latter system entails operating the ammonia oxidation section at atmospheric pressure, while the absorption unit is operated at higher pressures (capitalising on improved absorption rates).

Recent developments in the ammonia oxidation process have included efforts to reduce catalyst losses in the process. Platinum recovery filters have been installed at various stages in the process. Gold/palladium gauze filter pads have been added on the exit side of the catalyst bed, inside the reactor/converter units. These filters have reportedly ensured a platinum recovery of 80% (Ref. PT4). Another trend has been for the use of additional filters in the downstream units. These filters are of alumino-silicate construction.

Perhaps the greatest progress in nitric acid production technology has been in the manufacture of strong nitric acid (>90% by weight). Advances in the areas of super-azeotropic distillation and in high-pressure absorption are most significant. For further information consult References PT4 to PTl 1, PT14 and PT15.

Research work is continually being performed in an effort to reduce nitrogen oxide emissions from nitric acid plants. The Humphreys and Glasgow/Bolme nitric acid process is just one example of a new philosophy being applied to the absorption systems of weak nitric acid plants (50-68% by weight). Nitrogen oxide emissions have been reduced from 2000-5000 ppm to less than 1000 ppm (Ref. PT12, PT13). For the production of stronger nitric acid, tail gases are now being treated by selective or non-selective catalytic combustion systems. These innovative units have reduced the nitrogen oxide emissions to below 400 ppm (Ref. PT5, PT6).

1.2.4 Ammonia Oxidation Chemistry

Notably, all commercial nitric acid production methods used today are centred around the oxidation of ammonia. It is therefore appropriate to investigate the chemistry of this process, in the knowledge that it is directly applicable to any of the production processes available.

The chemistry of the oxidation of ammonia is surprisingly simple. It begins with a single pure compound, plus air and water, and ends

with another pure compound in aqueous solution, with essentially
no by-products. The process may be described by just six major
reactions as shown in Table 1 .I .

With reference to Table 1 .I, Reaction 1 is the overall reaction for
the process. This net result is achieved from three separate, and
distinct, chemical steps. The first is the oxidation of ammonia to
nitrogen monoxide (Reaction 2). The second is the further oxidation
of nitrogen monoxide to nitrogen dioxide (Reaction 3), then nitrogen
dioxide to nitrogen tetroxide (Reaction 4). The third and final stage
involves the absorption of these nitrogen-based oxides into water to
form the nitric acid product (Reactions 5 and 6). In most commercial
processes, each of these three stages are conducted in separate
process units.

The first step in the process is the heterogeneous, highly
exothermic, gas-phase catalytic reaction of ammonia with oxygen
(Reaction 2). The primary oxidation of ammonia to nitric acid (over a
catalyst gauze of 9:I platinum/rhodium alloy) proceeds rapidly at
process temperatures between 900-970°C.

The second step in the process involves two reactions (Reactions 3
and 4). These are the oxidations of nitrogen monoxide to the dioxide
and tetroxide forms. The equilibrium mixture is loosely referred to as
nitrogen peroxide. Both reactions are homogenous, moderately
exothermic, gas-phase catalytic reactions.

TABLE I.I

Chemical reactions for the oxidation of ammonia

Main reactions	*Heat Of reaction* $(AH,; kJ/mole)$
1. $NH_3 (g) + 2O_2 (g) \rightleftharpoons HNO_3 (aq) + H_2O (l)$	- 4 3 6 . 9 1 8
2. $4NH_3 (g) + 5O_2 (g) \rightleftharpoons 4NO (g) + 6H_2O(l)$	▬ 226.523
3. $2NO (g) + O_2 (g) \rightleftharpoons 2NO_2 (g)$	– 57.108
4. $2NO_2 (g) \rightleftharpoons N_2O_4 (g)$	-28.617
5. $3N_2O_4 (g) + 2H_2O (l) \rightleftharpoons 4HNO_3 (aq) + 2NO (g)$	– 15.747
6. $3NO_2 (g) + H_2O (l) \rightleftharpoons 2HNO_3 (aq) + NO (g)$	58.672

All reactions shown are highly exothermic. The known side reactions and their heats of
reaction are presented in Appendix A (Table A.2). The source of this reaction data is Ref.
PT1 (p.174).

The third step in the process involves cooling the reaction gases below their dew point, so that a liquid phase of weak nitric acid is formed. This step effectively promotes the state of oxidation and dimerization (Reactions 3 and 4), and removes water from the gas phase. This in turn increases the partial pressure of the nitrogen peroxide component.

Finally, nitric acid is formed by the reaction of dissolved nitrogen peroxide with water (Reactions 5 and 6). The formation of two moles of acid is accompanied by the formation of one mole of nitrogen monoxide gas. This nitrogen monoxide must be recycled within the process.

CHAPTER 2
Feasibility Study and Literature Survey

2.1 Initial Feasibility Study

THE FEASIBILITY study for a chemical process design investigates both the technical and economic feasibility of the proposed project. This feasibility study is only an introductory assessment, at this stage the process route has not yet been finalised although a preferred route may be apparent. Part of the work in preparing the feasibility study is to obtain information regarding the alternative process routes, and to provide an assessment of the suitability of these routes for the particular project.

The technical part of the feasibility study considers the alternative processes, and the equipment that constitutes the chemical plant in each case. At this stage it is necessary to identify any items of equipment that pose particular or unusual design problems, or which are very expensive or hazardous. The feasibility study should determine whether it is possible to design and build a chemical plant for a particular manufacturing process. Any external factors that may influence the operation of the plant should be noted, e.g. discharge levels, stability of raw materials supply, etc.

The economic feasibility of the process should be established at this stage. Again, this is only an introductory assessment performed more to establish that the plant is not definitely a loss-maker, rather than deciding that it is a particularly attractive proposition. A full and detailed economic evaluation of the plant and process is performed later in the design study (see Chapter 6) after the process route has been finalised, a detailed equipment listing prepared, and preliminary equipment designs have been performed. The following steps need to be performed to establish the economic feasibility of the process:

(a)   Determine the cost per tonne (or kg) of *all* raw materials used in

the process, these should be ***delivered*** prices to the anticipated plant location including shipping, handling, etc.

(b) Determine the current price paid per tonne (or kg) for the proposed chemical at the plant location, or at the market location if these are very different (in this case estimate transportation costs).

(c) Estimate the quantities of raw materials required to produce (say) one tonne of product. Determine whether the cost per tonne of raw materials is ***more*** than the cost of purchased product for the production process. If it is, why bother to build the plant?

(d) Establish the market for the chemical, both locally and for exports. Determine whether these are contracting or growth markets. Establish the ***minimum*** sales level that would be expected. Local or national sales are usually more predictable and stable than overseas markets. Compare this expected sales figure with estimates of the economic throughput for such a plant — based on available data for similar plants.

(e) Obtain total capital cost data for an existing plant producing the same chemical, preferably a recently commissioned plant of approximately the same required production rate. However, any data are (usually?) better than none, so if no data for such a plant is available try to obtain information for a plant producing a similar type of chemical with comparable plant equipment and throughput. Use one of the factored cost estimate methods (see Chapter 6) to estimate the present capital cost of the proposed facility.

Remember that this is only an estimate ($\pm$ 30%!), an accurate capital cost figure requires detailed designs of the plant equipment to be performed and a large amount of time to be spent. All that is required at this stage is a 'ball-park figure' to establish the feasibility of the project. Note any major differences between the existing plant and the proposed facility, e.g. cost of land, local availability of raw materials, existing transportation networks, etc. In particular, be careful about the choice of exchange rates. If the exchange rate has fluctuated widely over the previous 12 months, it will be preferable to use the highest and lowest rates to obtain a range for the projected capital cost estimate, rather than the current or average exchange rate.

(f) Estimate the number of years of possible plant operation (longer for a 'basic' chemical commodity) and determine the pay-back period for the plant capital cost. Determine the rate of interest charged (almost as difficult as estimating exchange rates!). Determine the yearly capital

cost repayments, and also the repayment cost per tonne of product. Estimate (very approximately) the operating costs, e.g. labour, utilities, maintenance, etc., for the plant (is 10% of the capital cost a suitable 'first' estimate?).

(g) Itemise all costs either annually or per tonne of chemical produced, including raw materials, capital cost repayments, operating costs, etc. Compare this final total figure with the expected (maximum and minimum) revenue from sales of the chemical.

Does the plant appear to be economically feasible at this stage of the project?

(h) Present the findings of the feasibility study in a clear and concise written form. Identify ***all*** assumptions made and any factors that significantly affect this study. Make recommendations.

Should the design project proceed to the next stage?

Note Students sometimes complain that a project is not feasible at this stage and should be abandoned. However, the design team may not be fully aware of external developments, e.g. anticipated oil price rises, shortages of key commodities, etc. If a project does not ***appear*** feasible, it may be necessary to complete the design study in readiness for possible market changes and more favourable economic conditions. In this situation, it is preferable to perform the detailed economic evaluation (see Chapter 6) in terms of the changes in raw materials costs, increase in product price, etc., necessary to make the project profitable and feasible, rather than as a detailed statement of loss per annum.

Action: ***Decide whether the design project should proceed to the next stage. If no, list all the reasons (in order of priority) and indicate what changes (if any are possible) would reverse this decision. If yes, list all the assumptions/restrictions upon which this decision is based (in priority order).***

References

Baasel, Chapter 3 (1976); Peters and Timmerhaus, Chapter 2 (1980); Ulrich, Chapter 2 (1984); (see Appendix L here). Extensive reference lists are included at the end of the quoted chapters in Baasel(1976) and Peters and Timmerhaus (1980).

2.2   Presentation of Literature Surveys for Projects

The following notes are intended to provide ideas and suggestions —
they do not provide the 'correct' presentation or the only way of
presenting   literature   surveys.

Summary

The literature survey is a review of the important information available
which is *relevant* to the particular project topic. It may include
information available in books, encyclopedias, lecture notes, journal
articles, reports (both government and company), doctoral theses,
standards, patents, and personal communications. It is important to
identify and include relevant information only, and to state at the start
of the survey what kind of information is required.

1. Ensure that a literature survey is appropriate for:
(a) the amount and type of information available;
(b) the intended audience (find out what the supervisor/marker wants!).
The literature survey must conform to any formal requirements
provided by a company or institution (e.g. for the presentation of a
graduate thesis). Find out if any requirements exist, and study previous
reports.

2. The literature survey of a report is *not* an essay. It must provide clear
and concise bibliographical information about the publications
themselves, *and* summarise the relevant information they contain.

3. The literature survey *must* include an introductory statement
making clear what kind of information is required for the particular
project.

4. The presentation of the literature survey will depend upon the type
of project being undertaken, e.g. laboratory based or a theoretical study.
The presentation will also depend upon the number of publications to be
discussed, i.e. a few papers can be reviewed in detail whereas a large
number of publications (usually) require more selective discussion.

5. Suppose that you have a 'reasonable' number of worthy references
(say 20-30). These are often discussed in chronological order, perhaps
from the most recent back to the earliest (the most recent publications
are often the most relevant -but not necessarily so!). **A** chronological

survey is usually presented if general background information is required.

6. Alternatively, the 'best' (i.e. containing the most useful/relevant information) 5 or 10 publications could be discussed first, and then the others possibly in less detail.

7. Sometimes one author is prominent in a particular field and these publications can be reviewed separately as a group.

8. In certain situations it is appropriate to review the available literature under relevant headings, e.g. historical developments, mathematical models, experimental data, review papers, general textbooks, etc. This would usually be done when a large amount of information is available.

9. If specific factual or numerical material is required, it is useful to organise the discussion of available literature around the specific points or items in question.

10. *What should the literature review review?*

The literature review is an *evaluation* of the published material in the field of interest. It is *not* merely a list of references-although that is also needed.

Select material for your literature review such that it includes:

(a) sufficient background information;

(b) the essential publications in the area (i.e. the 'definitive works');

(c) the major fields of investigation that are relevant to the particular project.

The review should refer to what is of importance/interest for the particular project. Each evaluation of a publication should detail what the publication contains that makes it worthy of inclusion, e.g. a theoretical development, experimental data, details of equipment design. It may be appropriate to add a subjective judgement of the publication, e.g. good/poor; useful/background; overview/detailed study; etc.

11. A summary of what has been revealed by the literature survey must be included. This may take the form of an extraction of the best data, a summary of pertinent opinion or a judgement on the state of knowledge in the field. This summary must concur with the objective stated at the outset (see point 3 above).

12. The literature review *must* be supported by a complete and accurate reference list. Each reference (in the reference list) must include full publication details. Suggestions regarding the presentation of

reference lists and references to material in a report, are included in point 19 below.

13.   Publications that are not referred to in the literature review (or in the report itself), but which provide useful background information, should be detailed in a separate list entitled 'Further Reading' or 'Additional Bibliography' or similar title.

14. Do not include full reference details in the review itself, or as footnotes — use a separate and complete reference list.

Reference to a publication within the literature review (and in the project report) is usually by means of *either:*

   (a) a reference number (in ascending order throughout the report) and a supporting numbered reference list; *or*

   (b) reference to the first and second authors names and the year of publication, e.g. Ray (1981) or Ray and Smith (1981) — use Ray et al. (1983) for three or more authors. If an author has published several papers in one year, it is necessary to distinguish between them, e.g. Ray (1980, A).

The supporting reference list is prepared in alphabetical order, do not use the abbreviation 'et al.' here — include full details of all authors (including all initials). Refer to the suggestions for preparation of references in point 19 to follow.

15.   The best advice is to adopt a referencing method, and to present a literature review and the reference list, as the supervisor wants them to be — go and ask what is required!

16. Examples of literature reviews/surveys are not included here (except in the Case Study material of Section 2.4). Students often assume that by using a particular example as a blue-print, they can produce a near perfect review. This is seldom true, *all* literature reviews can be improved. Always prepare a literature review so that it is appropriate for the information to be presented, and acceptable to the person who will mark the report!

17.   Only say what is important! Be clear and concise, do not copy or quote extensively from the publications. Excess verbiage often loses marks.

18.   Present your report in an impersonal and technical manner, pay attention to your spelling and English 'style'. Avoid using 'humorous' comments, the report should be interesting because of the information it contains and the way it is presented. Let someone read and criticise your first draft, and then make corrections. Write neatly, legibly and correctly. It may seem unfair, but reviews that are typed correctly

(without numerous mistakes) are usually initially better received. Aim the survey at the level of the graduate student. Avoid excessive use of jargon.

19.   Additional detailed notes *for* **the presentation** *of* **References and Bibliographies**

The Reference List (sometimes called Literature Cited or Bibliography) should only contain references that are cited in the text. A list of support literature, background or further reading, contains literature not cited in the text that is also relevant to the project.

Any reference listed must be so specific that the reader could easily locate the information. References to books must include: all authors and all initials; the full title and edition; the volume (if more than one); publisher, place and date of publication; any pages cited (either given in the text or included in the full reference). The **ZSBN** is optional but sometimes useful. References to journal articles must include: all authors and all initials; the full title of the article, the journal name (or accepted standard abbreviation); volume number (underlined or in italics); part number (in brackets); inclusive page numbers for the article; year of publication.

References to lecture notes should include the course and unit titles and designation numbers, the lecturer, the particular topic, the date, and any other relevant information, References to anonymous authors, personal communications, or where incomplete information is available, should be clearly stated.

References in the reference list should be presented as follows:

Konstantinov, L., Joosten, J., and Neboyan, V., 'Nuclear Power and the Electronics Revolution', **IAEA Bulletin, 27 (3),** pp. 3-6, 1985.

> **Note** Authors initials may be written before the surname; the abbreviation **'et al.** ' should not be used in the reference list — all names and initials should be included.

Ray, M.S., **'Elements** *of* **Engineering Design: An Integrated Approach'.** Prentice-Hall International, Hemel Hempstead, UK (1985).

> **Note** In this case, the relevant pages or chapter must be cited in the text as they are not included with the reference details; author's initials may be written first, i.e. MS. Ray; this is the first edition.

It is essential to be able to clearly identify specific information referred

to in the review. For short publications, usually journal articles where information is presented in a concise form, it may not be necessary to refer to specific pages or sections. For books and other lengthy publications, it is *essential* to quote the particular pages, chapters, figures, appendices, etc. This may be achieved by inclusion either in the text or in the Reference List. The former method is illustrated in the three examples given below.

Reference to a specific section of a publication in the Reference List may be achieved in the following ways:

In the review: refer to Ref. 8.
In the Reference List for Ref. 8: include all publication details and pp. 3946, or see Chapter 6, etc.

If a separate part of this publication (i.e. Ref. 8) is referred to later in the survey, it is then assigned an appropriate (new) number, say Ref. 15, and the Reference List includes:

15. Refer to pp. 51-55 in Ref. 8 above.

Full publication details should *not* be repeated within the
Reference   List.

References within the report may be cited in one of two ways. The method chosen should be adhered to throughout the report. In the first method, references are numbered in the order in which they are cited throughout the report. In the text, use the following reference notation as   preferred:

— as demonstrated by Ray[8] (pp. 3946).
— as demonstrated by Ray(8), Chapter 6.
— previously demonstrated (Ref. 1; Figure 3).

For this system, the Reference List is presented in numerical order with the   appropriate   number   before   the   first   author's   name.   The disadvantage of this referencing system is that any reference included in the report at a late stage requires a renumbering of all subsequent references, and the reference list.

In the second method, the reference may be cited by using the first author's name; any second author is also included, and *'et al.'* for three or more authors; the date of publication is included, and a suitable designation if the author has published several times. Specific pages, etc., are quoted in the text as shown below. The notation in the text may be:

— as supported by Ray (198 1: 18-24).

or

— as supported by Ray and Jones (1986: pp. 19-23).

or

— as supported by Ray et al. (1985, ref. B: Chapter 5).

An alphabetical reference list is prepared and included. For the examples given above, the list would contain a reference by Ray and Jones in 1986 and other years, or by Ray et al. with two references (at least) in 1985, designated Ref. A and Ref. B in the list. The disadvantage of alphabetical referencing is the extra 'bulk' required in the text compared with use of a number system. However, inclusion of an additional reference does not disrupt the numbering system (although the additional reference must be inserted in the list). The choice of referencing method may be a personal decision, or it may be specified by the company funding the work.

20. Final comments

Prepare and present material in a manner that is acceptable to the supervisor — go and ask what is required!

Only say what needs to be said — be brief, concise and informative.

Do not quote or reproduce extensive amounts of published material.

Present the review in a clear, structured and logical manner.

Your work will be judged by its presentation, i.e. spelling, legibility and English 'style'.

Obtain a critical appraisal of your written work and always aim to improve upon previous reports.

The literature review is an important part of a Written project. It contains useful information and is a valuable aid to future readers — if it is correctly structured and presented.

The notes presented in this section are intended to provide guidance to the student preparing a literature review for a project report. They should not be considered as a definitive set of rules to be followed in all situations. These notes should be applicable to the presentation of any type of report, e.g. design project, laboratory investigation, theoretical study, etc.

Action: Consult a range of project reports and read the literature review

sections. Identify their good and bad points. Prepare a draft literature review for your project. Consider carefully the objectives and the use of this review. Let someone who is suitably qualified(?) provide a critical assessment of your review. Re-write the draft review.

2.3 Case Study — Feasibility Study (Market Assessment)

Summary

The Australian domestic nitric acid market was found to suffer from cyclical variations, with seasonal highs and lows each year. This fluctuation is attributed to the major acid consumers (fertilizer and explosives manufacturers) being susceptible to seasonal variations in demand, and to the level of world commodity prices. However, the overall trend has been for 3% annual growth. Current Australian production is 200 000 tonnes each year (100% acid basis). Exports from and imports into Australia are virtually non-existent. A protective barrier in the form of high shipping costs, has in the past effectively closed the domestic market.

Worldwide annual production of nitric acid is at present approximately 34 million tonnes. The USA, UK, Poland and France are the largest producers. The trend in the last decade has been for growth by the larger producers, very much at the expense of the smaller ones. The global scene is a much more stable market. This can be attributed historically to consumption being more broadly based with a sizable consumption in chemical production processes.

An annual local market (Western Australia) of 65 000 tonnes for a 60%(wt) nitric acid product is envisaged for this plant. This market assessment is based upon anticipated demand from a new 50 000 tonne/year ammonium nitrate production facility. There is also an opportunity to exploit the potential South-East Asian market for up to 60 000 tonnes per year. This market may well be viable if low-cost shipping can be guaranteed through an arrangement with Alcoa of Australia Ltd. Alcoa imports caustic soda into WA, and the ships leave empty. It is proposed to use one of these ships to export up to 27 000 tonnes of 60% nitric acid into the foreign market.

Therefore, the total annual plant capacity proposed is 92 000 tonnes of 60% nitric acid. The plant should operate for 330 days/year (8000 hours) with production running at 280 tonnes per day.

References Used

Market Data (MD): 1,3,4,5.
General Introduction: G5.

2.3.1 Introduction

In order to establish the feasibility of the project, a study of the history of the nitric acid market over the last decade is presented, both in a domestic and global perspective. The size and nature of the market is studied, including the determination of general trends in industry and potential growth areas.

The limitations of looking at the history of any industry are recognized, however, the value of such an exercise should not be underestimated. This background, and a knowledge of the broader spectrum of the prevailing domestic economic climate, improves the potential for more reliable decision making.

The potential market size and its location, and the required product quality are all determined from this study.

2.3.2 The Domestic Scene

Australia currently produces approximately 200 000 tonnes of nitric acid (100% basis) per year (Ref. MD5). It is a cyclical market that responds directly to the performance of the agricultural and mining sectors. This occurs because Australian nitric acid is used almost exclusively for the production of ammonium nitrate (a nitrogen-based fertilizer and a mining explosive). The nitric acid industry has grown from a production capacity of 32 000 tonnes in 1967 (Ref. MD1). During the last decade, large deviations in production levels have occurred (Refs. MD3, MD4, and MD5). The overall trend has been for a 3% increase each year.

A breakdown of the areas of application within Australia are shown below:

Australian Market Distribution
Fertilizer Manufacture (NH_4NO_3) 45%
Explosives Manufacture (NH_4NO_3) 45%
Chemical Industry   10%

Western Australian Market Distribution
Fertilizer Manufacture (NH_4NO_3) 10%
Explosives Manufacture (NH_4NO_3) 90%

Australia is highly dependent upon the fertilizer and explosives industries. The Western Australian market is unique in its links with the explosives industry. Urea has largely replaced ammonium nitrate as a compound fertilizer.

Other graphs and tables which compliment the domestic market analysis are presented in Appendix B. They relate to almost all areas of activity for the local nitric acid industry. A list of these tables and figures is given below:

Figure B.l Australian production trend for nitric acid (1976-86).
Table B.l Australian annual production data for nitric acid (1976-86).
Table 8.2 Monthly Australian production for nitric acid (1985-86).
Table B.3 Australian export figures for nitric acid (year ending June 1986).
Table B.4 Australian import figures for nitric acid (year ending June 1986).

There are several manufacturers of nitric acid in Australia. They produce mainly for the fertilizer industry, which uses the acid for the production of ammonium nitrate and assorted nitrogenous compound fertilizers. A comprehensive list of those Australian producers is shown in Table 8.5 of Appendix B.

The activities of the sole Western Australian producer, Kwinana Nitrogen Company Pty Ltd (KNC), are relevant to this project. A brief summary of KNC operations is presented in Section B.l of Appendix B, including details of the process used and particular features, production capacity and product market. The most important disclosures made in researching KNC activities were the revelations regarding the cyclical acid demand and also the strong relationship

between the price of oil and production process economics.

   The implications of the data presented are discussed in more detail in Section 2.3.4 in relation to the world market.

2.3.3 The Global Market

Current world production is approximately 34 million tonnes per annum (Ref. MD4), and over 30% is produced in the United States. Of the remaining production, about 60% is based in Europe. The USSR (6 million tonnes), United Kingdom (3.3 million tonnes), Poland (2.4 million tonnes) and France (1.5 million tonnes) are the main producers. The world figures for the production of nitric acid over the last decade are shown in Table B.6 and plotted in Figure B.2 (of Appendix B). This data may be compared with the Australian figures already presented. Figure B.3 illustrates the annual percentage growth for both domestic and global markets to further assist comparison. A comprehensive list of worldwide manufacturers of nitric acid (according to country of origin) is presented in Ref. MD4 (pp. 166-167). The breakdown of market distribution for the industry, on a worldwide basis, is shown below:

World Market Distribution (Ref. G5, p. 565)

Ammonium nitrate (fertilizers and explosives)	65%
Adipic acid	9.5%
Military (other than ammonium nitrate)	6%
Isocyanates	1.5%
Nitrobenzene	1%
Potassium nitrate	0.5%
Steel pickling	0.5%
Others	16%

A full analysis of all data presented and its implications is presented in the following section.

2.3.4 Market Analysis Discussion

Figure B.l highlights the experience of the Western Australian producer, KNC. The Australian market is not entirely stable. Although

exhibiting good growth over the last twenty years, the market has seen deviations in production levels of over $\pm$ 10% from year to year. This is certainly not an ideal environment for a possible new producer. The large fluctuations seen in the Australian market are not observed in the world scene. This may be attributed to the limited applications for nitric acid in Australia (this small base demand being a problem endemic within the entire domestic economy).

As highlighted in Section 2.3.2, the major proportion of all nitric acid produced in Australia is used for the production of ammonium nitrate. Ammonium nitrate is in turn used in the agricultural industry as a fertilizer and in the mining industry as an explosive. Both of these areas have shown large demand fluctuations over the last decade. The collapse of world commodity prices, the expansion and recent decline of the agricultural sector for overseas markets, and seasonal weather variations have all had a significant impact.

The problems experienced in Australia have been effectively buffered in the world market by a broad, diverse market base. A substantial proportion of world production is used in other areas of the chemical processing industry (as shown in Section 2.3.3), thereby establishing a more stable base.

Ignoring the cyclical trends of the domestic market, the overall growth is running at around 3% with an export sector that shows promise. Current export levels are at 490 tonnes per annum. Export figures (Table 8.3) indicate strong growth, with good prices recorded even for larger deliveries (A\$400-A\$500 per tonne). Exports have mainly come from the producers in New South Wales. The export market to date consists of only regional neighbours (Fiji, Indonesia, New Zealand, Papua New Guinea, Vanuatu). This may be attributed to the high shipping costs facing most Australian producers.

Nominal amounts of nitric acid are imported into Australia from Europe, North America and Japan, totalling approximately 15 tonnes per annum (see Table B.4).

In contrast to the volatile Australian demand, the world market for nitric acid over the last decade has shown a relatively steady growth (approximately 2% per annum). Again, the broader applications base is at least partially responsible for this superior performance.

An interesting trend exposed by an examination of world production figures, is the success of the larger producers at the

expense of the smaller ones (Ref. MD4). The North American and European giants have absorbed the falling production elsewhere in the world, and expanded still further to stimulate the overall growth in world capacity. Fortunately the Australian domestic market has enjoyed an inherent protection from overseas manufacturers in the form of high shipping costs. Whilst infiltration from outside remains difficult, these high transport costs have also prevented substantial expansion into lucrative export markets. Hence the domestic market has evolved into an isolated industry that is almost totally self-sustaining.

Having assessed the market performance to date, it is now possible to develop the feasibility analysis one step further. If a viable plant is to be constructed, a potential market for any new supplies of nitric acid must be determined.

Integral to the feasibility of this new project must be the establishment of another ammonium nitrate plant in Western Australia. Such a plant, with an annual capacity of 50 000 tonnes, is planned for construction. This would require approximately 39 500 tonnes of nitric acid per year (on a 100% basis), and provides the required preliminary base demand necessary for a feasible project.

However, the fundamental viability of this project also necessitates breaking away from the traditional industry structure in Australia, whereby acid is produced and consumed *in situ.* It is intended to seek a broader base for the new plant. This will provide an added market stability on which this new plant can prosper. With no significant local chemical process industry, widening the market base must take the form of targeting new export markets. The South-East Asian market is convenient, but to date it has remained relatively untapped by Australian producers. Several South-East Asian neighbours, notably China, Indonesia, South Korea, Singapore and Hong Kong, all import substantial quantities of nitric acid each year. A full summary of potential export targets within the South-East Asian region, compiled from Ref. MD4, is given in Table 8.7 (Appendix B). This market totals 32 600 tonnes each year (on a 100% basis) at 1980 levels.

The South-East Asian market has previously been supplied mainly from Japan (about 10 000 tonnes/year) and Europe (about 20 000 tonnes/year). These figures are relevant to 1980 data (Ref. MD4, pp.1 07-l 08). It may be possible to acquire up to 50% of this demand.

There are two main factors supporting an optimistic view of the possibility of entering this foreign market. The first is the present low value of the Australian dollar. The fall of the Australian dollar between 1984 and 1987 compared to the US dollar, many of the European currencies, and particularly the Japanese yen, has given all Australian export industries a new competitiveness in the world market. This competitiveness extends to the chemical industry, and in this case to nitric acid producers.

The second factor relates to a possible source of low-cost shipping, the very area that has prohibited export expansion in the past. It is proposed to export 60% nitric acid in one of the ships that currently import caustic soda solution into Western Australia for Alcoa of Australia Ltd. There is sufficient capacity available with this option. The ship's storage holds will have to be coated with polyvinyl chloride (PVC) to enable it to carry both caustic soda and nitric acid. This should be relatively inexpensive to implement. By using Alcoa's ships to export nitric acid, large shipping costs can be avoided. Full details of this proposal are included in Appendix B.

The economic advantages outlined above should enable the supply of a price-competitive product, assuming investment and production costs are not substantially higher than those of local and overseas competitors. A marketing campaign should be launched in the South-East Asian region to obtain a proportion of this valuable export market.

Current market prices for nitric acid are shown in Table 8.8. The market value for a 60% nitric acid product is currently As345 per tonne.

2.3.5   Market Assessment Conclusions

A detailed study of the nitric acid market both within Australia and overseas has suggested that there is a potential market of up to 72 000 tonnes per year, on a 100% acid basis.

This conclusion indicates the feasibility of a nitric acid plant with a total annual capacity of 55 000 tonnes (on a 100% basis). This figure represents 92 000 tonnes per year (280 tonnes per day) of 60% nitric acid product.

This latter figure is based upon an anticipated local demand from a new ammonium nitrate plant in Western Australia. This plant should account for approximately 65 000 tonnes of 60% nitric acid per year. The remaining annual production of 27 000 tonnes of 60% nitric acid, will be exported to various countries in South-East Asia (less than half that presently imported in this region).

This production philosophy sees a guaranteed local market absorbing 70% of production, with 30% capacity in reserve to exploit the suggested export market. This would seem to be a sound, stable base on which to launch a more detailed feasibility assessment.

The plant should operate on a standard 8000 hour/year basis, with approximately 330 days of production. It is assumed that a market price of As345 per tonne can be obtained for the product.

2.4   Case Study — Literature Survey

Summary

The references used have been catalogued according to the type of information they contain. Each category has been reviewed to highlight the most useful sources of information.

The **General Information** group contains references with a broad coverage of most aspects of nitric acid production. The best of the eight general references used were Kirk and Othmer (Ref. G1), and Lowenheim and Moran (Ref. G2). Both give a good introduction to the properties of nitric acid and background for the industry.

The **Process Technology** category contains the sources specific to new processes and technological innovations. The investigation of Harvin, Leray and Roudier (Ref. PTl) was undoubtedly the most useful to this project. It listed operating pressures and temperatures, reaction yield and equilibrium data, energy production data, and capital/operating cost figures relevent to the single-pressure process of C&l Girdler. Chilton (Ref. PT2) whilst an older source was also of great value.

The third grouping is **Cost Estimation.** These references were of direct use in the capital costing. They provided estimation techniques, typical cost data and useful correlations. The source

most often used was Peters and Timmerhaus (Ref. CE9). This book was important in costing the individual units as well as providing typical values for sundry operating charges.

Within the *Market Data* category, the best information was provided by the Australian Bureau of Statistics (Refs. MD5, MD6, MD7). These provided current Australian production figures and export/import details.

The section entitled *Thermodynamic Data* contains references which include thermodynamic properties and also chemical reaction equilibria data, physical property data, and specific formulae required in calculations. It is difficult to select the best source as each provided different information.

2.4.1   Introduction

The large number of references used in this project were divided into subject categories according to the content of the articles and the type of information provided, this made evaluation easier. Five major categories were selected. These are listed below:

1. General Information (G)
2. Process Technology (PT)
3. Cost Estimation (CE)
4. Market Data (MD)
5. Thermodynamic Data (TD)

Each category will be reviewed, with the better sources being highlighted.

2.4.2   General Information

This category provided general information for the background of the industry and current process technology. The references tend to be broad but shallow in their coverage, their style being typical of an encyclopedia.

References GI to G5 were all chemical and process technology-type encyclopedias, providing the starting point for this report. They provided a very diverse coverage including the properties of nitric

acid, applications in industry, and the techniques employed in production. This group was used extensively to form the introduction for this report (Sections 1.2.1 to 1.2.4).

The most comprehensive reference in this group was undoubtedly Kirk and Othmer (Ref. GI). It provided an excellent historical background into process development and market characteristics, and also provided a broad coverage of the physical and chemical properties of nitric acid, handling methods, and materials of construction. A warning on using this or other process encyclopedias is that much of the technology presented in the first five references has now been superseded.

Gregory (Ref. G6) is a very comprehensive source specific to the industrial and commercial applications of chemicals. It includes a three page list of the uses of nitric acid. Similar information is presented for many other chemicals in this book. The one fault of this source is that it is 50 years old! In that period many process technologies have changed and the applications list, whilst very comprehensive, tends to be dated.

Reference G7 provides an excellent coverage of the materials of construction used in nitric acid manufacture and storage. Reference G8 is a Firth-Vickers catalogue, used to obtain the physical properties and corrosion resistance data for stainless steel 304L.

The final reference in this category is the Australian Government Gazette (Ref. G9) which includes information regarding industrial effluent limits, as demanded by the Environmental Protection Authority (in Australia). These data are quoted in the Environmental Impact Analysis (Section 5.4.7).

2.4.3   Process   Technology

This category contains those references specifying process innovation, operating conditions, utility requirements, catalyst information, etc. All information relates directly to the process requirements.

Harvin, Leray and Roudier (Ref. PTI) and Chilton (Ref. PT2) were the two main references used to determine the process selection for this project. Reference PTI contains an 'objective' comparison between the dual and single-pressure processes. It compares them on nearly all aspects, including capital and operating costs, process

operating parameters, equipment requirements and energy consumption. This was particularly useful in the process selection stage of the project. Having opted for the single-pressure process, Ref. PT2 contains a comprehensive summary of the comparable Du Pont process which could then be applied in conjunction with the information available in Ref. PTI .

References PT3 and PT4 both relate to catalyst performance and recovery. They detail the expected catalyst loss and reaction yields under various process conditions. They also suggest preferred methods of catalyst recovery.

The next seven references listed (PT5 to PTI 1) all relate directly to the production of 'strong' nitric acid (a product quality of > 70% by weight). This information was relevent at the process selection stage. When the choice was ultimately made to produce a product of sub-azeotropic quality, these seven sources were of little further use.

References PTI 2 to PT17 were all useful for determining the expected tail-gas emissions from the plant, and ways in which these tail gases could be treated if emission levels became too high. References PT12 and PT13 are both investigations into the Humphreys and Glasgow/Bolme tail-gas recovery technique. This technique includes a catalytic combustion unit immediately following the tail-gas expander. Reductions in tail-gas emissions are reported from levels greater than 5000 ppm to less than 400 ppm by this method. Reference PT17 refers to the use of hydrogen peroxide for the scrubbing of tail gases. Whilst good results are achieved by many of the techniques mentioned in these references, they tend to relate more to the 'strong' acid production facilities which suffer most from tail-gas emission problems.

Handforth and Tilley (Ref. PTI 8) report on experimental work (in 1934) concerning platinum loss from reactor catalyst beds.

2.4.4 Cost Estimation

All eleven references in this category relate directly to Section 6.1, the **Economic Evaluation** of this project. They contain capital costing methods, cost correlations, typical plant and equipment costs, and economic indicators to account for cost changes with time.

Reference CE8 contains the 'economic indicators' including the M&S Equipment Costing Index and the CE Plant Cost Index. These indices and their component values are listed from 1977 to 1985.

Peters and Timmerhaus (Ref. CE9) was the most used book in this category. It explains the ratio method and factored cost method of capital cost estimation. It also contains nomographs and correlations so that many plant equipment items can be costed. There are tables of typical values for costs such as insurance, depreciation and engineering.

Hignett (Ref. CE1 1) is an excellent source for the determination of production and operating costs specific to nitric acid manufacture. It contains tabulated data for the typical production cost components and capital cost details of a 1700 tonnes/day plant. These figures could be compared with those presented in Ref. PTI for determination of an accuratevalue for the 280 tonne/day plant in this project.

The other references in this section all contain useful nomographs and correlations for the estimation of capital and equipment costs.

2.4.5 Market Data

This group of references all contain information regarding the production (and producers) of nitric acid both in Australia and worldwide. No single reference was outstanding as each contained different statistics.

Reference MD1 provided data for world nitric acid production for each year from 1966-76. References MD3 and MD4 are much more comprehensive in this respect, and they complete the world production figures up to 1984. They also list data for many countries including production figures, imports and exports, and the names of the producers in each country. Reference MD2 extends the list of Australian producers included in Refs. MD3 and MD4.

References M D5 to MD7 are publications of the Australian Bureau of Statistics. These papers provide data for Australian production, import and export figures, and prices paid/received. The real value of these sources is that information only a few months old is available. References M D8 and M D9 both provide data for nitric acid prices on

the world market. Reference MD9 lists prices in three currencies (Japanese yen, pounds sterling and US dollars). Reference MD1 0 was used as a source of current exchange rates.

2.4.6 Thermodynamic Data

This category was used to provide chemical and physical data for nitric acid. The references also contain extensive details of chemical equilibria appropriate to the process, and several formulae used in the mass and energy balance calculations.

Reference TDl contains an enthalpy table for ammonia at different pressures. Reference TD2 contains a series of tables in an appendix from which the specific heats of the reaction-gas mixture were calculated. Humidity charts were also useful. Reference TD3 is valuable for its steam tables, while Ref. TD4 contains both thermodynamic and chemical equilibria data for nitric acid. The final reference, Robertson and Crowe (Ref. TD5), contains formulae and tables for the sizing and choice of an air-feed compressor.

2.5 Case Study — Bibliography

It seems more appropriate to include the Bibliography after the Literature Survey rather than at the end of the report.

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CHAPTER 3
Process Selection

3.1   Process Selection — Considerations

FOLLOWING THE feasibility study, the next stage in the design project is the evaluation and comparison of the alternative process routes for manufacture of the chemical. The selection of an appropriate process is an important decision, all the subsequent work depends upon this choice. Although the selection can be changed or modified at a later stage, at least before the plant is built, such a decision results in a serious waste of time and money. However, probably not such a waste as building an uneconomic or unsafe plant!

Obtain a process flow diagram (flowsheet) or Process and Instrumentation Diagram (P & ID) for each process route under consideration. These diagrams are usually quite complex and include all the ancilliary equipment, instrumentation and control details. To obtain an initial appreciation of each process, and to make an initial comparison, draw a simple block diagram from each process flowsheet showing only the main chemical engineering plant items, e.g. reactor, absorption column, purification stages, etc. This simple diagram will not provide sufficient detail for final process selection, but it will provide an initial comparison of different processes and a quick familiarisation with the process stages comprising each route.

The reactor is the heart of most processes, developments in process technology often centre around improvements in the design and operation of the reactor. This is often the basis of 'new and improved' processes. Subsequent stages in the chemical process are usually concerned with the separation of various chemicals from the desired product, followed by the final purification stages.

Previously in the feasibility study (Chapter 2) the appropriate purity of the final product should have been established. The product purity *must* be dictated by the customer (market) requirements. In some

situations it may be possible to remove a low-purity product from the process for a particular application, and then purify the remaining chemical for another customer, The possibility of further purification should be determined in anticipation of new applications, environmental regulations, etc.

Processes often differ in terms of the process conditions, e.g. high-pressure and low-pressure processes, or the type of reactor that is used, e.g. gas-phase catalytic reactor or liquid-phase CSTR. These differences should be clearly marked on both the simple block diagrams for each process and the detailed process flowsheets, they often determine which process route is ultimately selected. Many older, established processes were operated under conditions of high pressure or high temperature (mainly in the reactor), whereas the newer processes have often been 'improved' by operation under less severe conditions.

For student design projects. the older and less efficient processes are often the only ones for which detailed process data (suitable for the detailed equipment design) are available. Information concerning the 'new' technology and relevant experimental design data may still be kept secret. Hence, the requirement sometimes for students to select a process route and perform a design study for a less efficient process than they would ideally choose. However, the training and experience in plant and process design is no less diminished, and the real thing is awaiting their talents in industry!

Many older processes (developed before the 1970s oil price crisis) were less energy efficient than those developed more recently. This should be apparent by consideration of the energy conservation features included in new plants, and the subsequent increase in complexity of the associated P & ID. Energy conservation is discussed in Section 8.2.1. The selection of a process route for production of a chemical will depend upon the following factors/considerations:

(a) Will the process produce what the customer requires?
(b) Is it possible to design, build and operate this plant economically (and safely)?
(c) The necessary design data, technology, fabrication methods and materials, raw materials, etc., must be available.
(d) The plant must operate in a safe manner, providing an acceptable hazard risk (see Section 8.2.3) to the plant employees and the public.

(e) The plant must conform to any environmental protection requirements (see Section 5.3), and any possible future restrictions.

(f) The plant must be as energy efficient (and energy self-sufficient) as possible.

(g) Maintenance requirements should be minimised.

(h) The plant should be designed to operate adequately under conditions of reduced throughput (say by 50%), and for increased production (say 25%). The latter case represents an overdesign of the plant and additional capital cost. In these situations, the energy consumption is greater than the optimum requirement. However, the need for operational versatility usually overrides these considerations.

(i) The production of any unusable by-products should be minimised.

(j) All necessary utilities should be available, e.g. electricity, cooling water, etc. The utilities specifications must be established, these are dependent upon whether a 'grass-roots' plant is to be built or the plant situated in an existing chemical complex. These specifications include the steam-pressure levels, cooling water and/or refrigerants available, etc. The unit costs must also be determined.

(k) Etc., etc.

As with many aspects of design work, the final choice will usually depend upon a compromise between various features of different processes. It is unlikely that one process will possess all the advantages and no disadvantages. Sometimes there will be one overriding factor that influences the selection of a particular process, e.g. availability of a particular raw material, minimum cooling water requirements, etc. However, whichever process route is finally selected, it must fulfill all the criteria established in the project brief (Section 1.1).

The aim of process design is to build the 'best' overall plant, all units within the designed plant may not operate at maximum efficiency or full potential. It is necessary to achieve a balance between the conflicting requirements of the individual units to produce the 'best' plant possible. Plant design is an exercise in compromise and optimization.

Although not mentioned so far in this section, the process control and instrumentation requirements (see Section 8.2.2) must be considered when selecting the process route. The ability to provide suitable operational control over the process and the availability and cost of necessary instrumentation are major considerations. If these aspects are ignored at this stage, it could well be that the detailed plant design is at

best more expensive than is necessary or at worst impossible to operate.
The attitude that 'whatever is built can be operated and controlled' is
both dangerous and unprofessional.

Action: **Select the 'best' process route for production** *of* **the chemical,
and justify   this decision.**
            **Prepare a detailed process jow   diagram, and as far as possible a
            P&ID.**

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3.2   Case Study — Process Selection

Summary

All commercially produced nitric acid is now prepared by the
oxidation of ammonia. The requirement for a nitric acid product of
60%(wt.) immediately restricts the choice of a recommended
production process. Only two processes are possible, both highly
efficient, each offering distinct advantages under different market
conditions.

   The first process is the so-called *dual-pressure process.* This
process features an atmospheric ammonia oxidation stage which
takes advantage of the higher reaction yields available at lower
pressures. The reaction mixture is subsequently compressed up to
approximately 1000 kPa for the absorption stage of the process. The
absorption of nitrogen oxide gases is favoured by the higher
operating pressures.

   The second process is the *single-pressure process.* This process is
essentially identical to the dual-pressure technique except that it

maintains a pressure of about 1000 kPa throughout the process,

The dual-pressure process has slightly lower operating costs due mainly to the higher efficiency in the ammonia oxidation stage. It does however suffer a significant capital cost disadvantage. It is shown that the operating cost advantage would take 21 years to outstrip the initial capital cost difference. This long payback period suggests that the single-pressure process is an economically preferable option, particularly for plant operation in the Western Australian cyclical market.

The single-pressure process is also a higher net exporter of medium-pressure steam and has a lower utility requirement, and is therefore selected as the preferred option.

The single-pressure ammonia oxidation/absorption process is selected based upon the operating criteria used in the single-pressure process developed by C&I Girdler.

It is also anticipated that this plant will be part of a larger chemical complex. Ammonia will be produced by steam reforming of natural gas. The nitric acid plant will take a portion of the ammonia product, and nitric acid and ammonia will then be used to produce ammonium nitrate.

References Used

ProcessTechnology (PT): 1,4,5,6,8,9,10,15,18.

3.2.1 Introduction

The designers of nitric acid processes have generally accepted the chemical steps outlined in Section 1.2.4. Hence, process developments have centred around optimizing the design within the bounds of available equipment, materials of construction, and economics.

Many new developments have occurred over the last 20 years, as discussed in Section 1.2.3, particularly for the production of strong nitric acid (> 90% by weight). Strong nitric acid production employs the technique of super-azeotropic distillation to concentrate weaker acid beyond the azeotropic point (68% by weight). However, it is not necessary to discuss this technology further because the market

described in Section 2.3 clearly requires a product of sub-azeotropic quality. For information regarding the strong acid technology consult Refs. (PT) 4, 5, 6, 8, 9, 10, 15.

The two important modern weak nitric acid (60-68% by weight) production processes which have been developed are the dual-pressure process and single-pressure process. The dual-pressure process was developed and used mainly in Europe, whereas the single-pressure process (developed by Du Pont) is common in the USA. Both processes are capable of highly efficient production of the same quality product, with the same limits on atmospheric pollution. The process flow diagrams for the dual-pressure process and the single-pressure process are presented in Figures 3.1 and 3.2 respectively (Ref. PTI; pp.l76,177).

Section 3.2.2 compares these leading process technologies in order to select the most applicable method of weak nitric acid production. The most appropriate process is discussed fully in Section 4.2. Summaries of both processes are included in Appendix C.

3.2.2 Process Comparison

Both processes follow the basic scheme outlined in Section 1.2.4 for the catalytic oxidation of ammonia. In summary, this involves an oxidation stage whereby ammonia is reacted with air in a catalytic converter at temperatures in the range of 850-950°C. Reaction gases pass through a series of energy recovery stages before entering an absorption column. The bottoms from the column are bleached of dissolved nitrogen peroxide using air, and the resulting solution is the weak nitric acid product. The major difference between the two processes lies in the initial conversion stage. The dual-pressure process employs a conversion stage operating in the range loo-350 kPa, and a reactor temperature of about 865°C. The single-pressure process however operates the converter at 800-l 100 kPa, with a reactor temperature closer to 940°C.

Both processes are capable of highly efficient production of the same quality product with the same limits on atmospheric pollution, therefore any comparison must be related to other aspects unique to each process. These distinctions will ultimately determine the most desirable process for this application. The process comparisons are
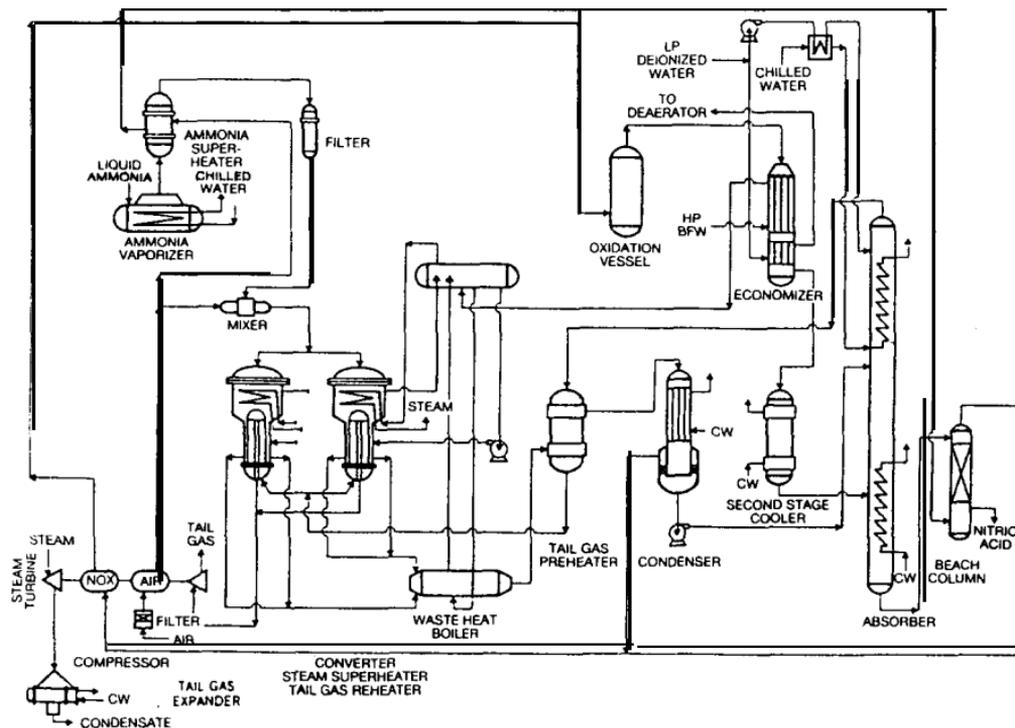
FIGURE 3.1 Flowsheet for dual-pressure nitric acid process. (From Ref. PT I, p. 176; reproduced with permission from American Institute of Chemical Engineers

FIGURE 3.2 Flowsheet for single-pressure nitric acid process. (From Ref. PTI, p. 177; reproduced with permission from American Institute of Chemical Engineers.)

Reproduced from:
Harvin, R.L.; Leray, D.G., and Roudier, L.R., "Single Pressure or Dual Pressure Nitric Acid: An Objective Comparison", *Ammonia Plant Safety,* Volume 2 I, pp. 173-I 83 (1979).

summarized in Tables C.l to C.6 and are presented in Appendix C. These tables are listed as follows:

Factors Favouring the Dual-Pressure Process

In the single-pressure and dual-pressure processes, the catalyst volatilizes at a rate determined by the converter exit-gas temperature. Experimental work indicates that the rate loss of catalyst (without a catalyst recovery system) is approximately three times more rapid at 973°C than at 866°C (Ref. PT18). From plant operation data, the loss from a dual-pressure converter (operating at 866°C) is estimated at about 0.10 g/tonne of 100% acid, and from a single-pressure converter (operating at 937°C) it is estimated at about 0.38 g/tonne of 100% acid.

In the single-pressure process, the ammonia content of the mixture should not exceed 10.5% because of the lower explosive limit. In the dual-pressure process with lower converter pressures, the explosive limit is higher providing additional operating flexibility.

Total production cost for a 1500 tonne/day plant built in the USA in 1979 (given in Table C.5) has a A$2.15/tonne acid advantage in favour of the dual-pressure process. This advantage is directly attributable to a higher ammonia yield and a smaller indirect cost structure (due mainly to a 20% lower catalyst inventory requirement and less frequent catalyst changes). This production cost advantage also prevails despite inferior steam production capacity. Production cost estimates for a similar sized plant in Europe confirm the dual-pressure process advantage of A\$S2.00/tonne of acid produced. Although the plant data relate to overseas operations where cost structures may be different, the figures do illustrate an important and valid trend.

Factors Favouring the Single-Pressure Process

The single-pressure process uses a higher ammonia conversion pressure. This higher pressure provides advantages in terms of equipment design, e.g. smaller converter dimensions and a single heat-exchanger-train layout.

The higher temperature and the favourable pressure both increase the energy recovery from the process. The single-pressure process provides an extra 10% high-level recoverable heat energy as shown in Table C.3.

Plant capital costs in the USA have been estimated at As8 million (USS5.1 million) for the single-pressure process and AS9.1 million (USS5.8 million) for the dual-pressure process. The AS1 .l million higher cost of the dual-pressure process is accounted for by the larger vessels required at lower operating pressures. Estimates made for the two plants in European locations show a differential of AS0.8 million, also in favour of the single-pressure process.

A discounted cash flow (DCF) analysis based on these US figures was performed by matching the capital cost advantage of the single-pressure process against the production cost advantage of the dual-pressure process. The results of these calculations are given in Table C.7, they indicate that it would take 21 years for the lower operating cost of the dual-pressure process to finally overcome its initial capital cost disadvantage. Although this plant should be aiming to establish itself as a long term supplier (>20 years), a pay-back period of 21 years is far too long particularly in the fluctuating local nitric acid market. For this reason, the single-pressure process is considered to be an economically superior proposition.

Other   Considerations

Harvin, Leray and Roudier (Ref. PTl; p.183) make several conclusions in their assessment of the advantages of each process. They conclude that the two processes may be economically competitive under particular conditions. Also selection of a process may depend on other factors such as:

(a) Better utility integration with other plants for a particular process.

(b) Limited space availability may favour the single-pressure process.

(c) Capacities of 1130-l 360 tonnes per day favour the dual-pressure process, because of the possibility of absorption up to 1550 kPa.

(d) Capacities of 270-540 tonnes per day favour the single-pressure process since lower conversion pressures and temperatures may lead to lower catalyst consumption.

(e) Requirements for 60-65% acid strength may favour the dual-pressure process because of higher nitrous gas concentrations to the absorber.

Their final comment was that: 'In the USA, the relatively low raw materials and utilities costs, together with the methods of plant finance, make the trade-off between these factors opt in favour of the single-pressure process for plants built up to 1979'.

3.2.3 Process Selection Conclusions

The single-pressure process appears to be preferred for a plant producing 280 tonne/day of 60% nitric acid. The capital cost advantage of this process surpasses the benefits of the superior operating cost structure of the dual-pressure process. Furthermore, the higher temperature and the favourable effect of pressure enable a greater recovery of energy from the process. This choice, made on both economic and operational grounds, is consistent with the conclusions presented in Ref. PTl .

CHAPTER 4
Process Description and Equipment List

4.1 Introductory Notes

AFTER THE process route has been selected, it is both possible and necessary to prepare an equipment list or equipment schedule. The process description, preliminary design calculations and equipment listing are usually performed at the same time as the preliminary mass and energy balances (for obvious reasons). However, for convenience of presentation and discussion the material and energy balances are presented separately here in Chapter 7. The equipment list should be as detailed as possible, but it will obviously not be complete until the detailed equipment designs are available (see Chapter 8). The reason for preparing a listing at this stage is to provide full documentation for the project and to establish a data base.

The equipment list is used in conjunction with the process flow diagram (or P & ID). Each item of equipment on the flowsheet should be assigned a unique reference number, and that number is used to cross-reference items in the equipment list. Particular letters may be used to identify similar items of equipment, e.g. reactors as Rxxx, pumps as P-001, etc.

The equipment list should include all available information that will be useful for the detailed design of the equipment. This should include the physical size of an item (if known), operating temperature and pressure, wall thickness, height of packing, etc. The original reference sources for all this information must be included. If one major publication was used, then it may be stated in the introduction to the equipment list that all data were obtained from this source unless stated otherwise. Although the majority of this information will be for similar plants, the final design details for the new plant may be very different. However, the equipment list provides a useful reference source at this early stage of the design.

The following information must be included in the equipment list so that the cost of purchasing and installing each piece of equipment can be determined.

(a) Specific type of equipment.
(b) Size and/or capacity.
(c) Material of construction.
(d) Operating pressure.
(e) Maximum operating temperature (if above ambient), or minimum temperature if refrigerated.
(f) Insulation required.
(g) Corrosion allowances (if large).
(h) Special features, e.g. jackets on heat exchangers.
(i) Duplication of plant items (for safety and/or reliability).

The heat exchangers and pumps cannot be sized until the energy balance for the plant is completed (Chapter 7). The final energy balance depends upon the energy conservation measures to be employed (Section 8.2.1), and the plant layout (Section 5.2). Other equipment should be sized reasonably accurately at this stage. Many approximate methods of equipment sizing are available, reference can be made to the books by Aerstin and Street (1978); Baasel (1976); or Ulrich (1984) for particular examples (see Appendix L here). References for the sizing and design of particular items of equipment are included in Baasel (1976; pp.133-139).

Action. *Prepare a complete equipment list for the chosen process route, include references for all information and data.*

References

Baasel (1976; Chapter 5) and Ulrich (1984; Chapter 4).

4.2 Case Study — Process Description

Summary

The process selected for the manufacture of 60%(wt.) nitric acid is a single-pressure type similar to that developed by C&I Girdler in the United States (see Figure 3.2). The operating pressure is

approximately 10 bar gauge (1000 kPa).

The major process units include an air compressor to provide feed air to the process, and an ammonia vaporizer and superheater for pretreatment of the feed ammonia. A reactor vessel with a fixed platinum/rhodium catalyst bed quickly oxidizes the ammonia at reaction temperatures approaching 950°C. The reaction yield is 95%. A heat exchanger train immediately following the reactor is used to recover reaction heat. Reaction heat is recovered for both gas expansion (to provide shaft power for the air compressors) and for production of medium-pressure steam (at 380°C and 4000 kPa). The high-level energy available in the process is shared approximately equally between gas expansion and steam production. About 40% of all steam production is delegated to 'in- house' process requirements, leaving about 3200 kg/hour available for export.

Cooled reaction gases are absorbed in water/weak acid using a sieve tray-type tower. The bottoms from this tower is a so-called 'red' product acid, the colour resulting from dissolved nitrogen oxide impurities. The red product acid is then bleached in a smaller sieve-tray stripping column. Air is bubbled through the red acid to strip out the dissolved nitrogen oxides. Bottoms from this column is the product nitric acid at 60%(wt.) concentration.

The design proposed in this section features two major differences from the flowscheme used by C&I Girdler. The first change is the elimination of the hot-water loop. This was considered to be contributing an unnecessarily high load on the refrigeration unit. The cooling service provided by this loop has been replaced by drawing from the standard 'process cooling water' circuit, The second change is the re-routing of steam condensate, produced at the ammonia superheater, into the vapour/liquid separator. This stream was being returned to the deionized-water circuit and constituted a waste of available heat energy. The condensate (at 250°C) is better utilized when it flows to the vapour/liquid separator and joins the waste-heat boiler feed.

References Used

Process Technology (PT): 1,2.
Thermodynamic Data (TD): 5,6,7.

4.2.1   Introduction

The single-pressure process was discussed in Section 3.2 and shown to possess both economic and operational advantages for this plant size. The process described in this design employs technology developed by C&I Girdler in their single-pressure process (see Ref. PTI, pp.173-183). This section of the report describes specific details of the process.

The single-pressure process shown in Figure 3.2 was developed to take full advantage of the high process pressure. Higher pressures allow the size of equipment to be reduced throughout the process. Elevated pressure is also shown to be of great advantage for absorption, while exhibiting only slightly decreased yields at the reaction stage (Ref. PT2; p.6). An initial two-stage compression step is used to raise the pressure throughout the entire process.

4.2.2 The Process

The process begins with the vaporization of ammonia at 1240 kPa and 35°C using process heat. Steam is then used to superheat the ammonia up to 180°C. Filtered air is compressed in a two-stage axial compressor to a discharge pressure of 1090 kPa and temperature of 232°C. Part of the air is diverted for acid bleaching, and the balance for ammonia oxidation is circulated through a jacket on the inner wall of the reactor. This preheated air and the ammonia vapour (10.3% ammonia by volume) are then mixed and passed through the platinum/rhodium catalyst gauze. The ammonia is rapidly oxidized and the heat of reaction raises the reaction-gas temperature to 927-937°C.

The reaction gas flows through a series of heat exchangers for recovery of energy as either high-pressure superheated steam, or as shaft horsepower from the expansion of hot tail gas. This heat exchanger train cools the reaction gases to 185°C. At this stage, 70% of the nitrogen monoxide initially formed will have already oxidized to nitrogen dioxide. Further cooling to 60°C (in the primary cooler/ condenser) is followed by separation of approximately one third acid product as 42% strength nitric acid. The gas reaches a state of oxidation to nitrogen dioxide of 43%, with the dimerization being approximately 20%. The **gas** is then combined with bleached air

containing additional nitrogen peroxide. This mixture further oxidizes using the additional oxygen supplied from the secondary air stream. It passes through an empty oxidation vessel and then through the secondary cooler. In cooling from 140°C to 65°C in the secondary cooler, the gas provides heat to a recirculating warm-water loop. This loop in turn provides the energy to evaporate the liquid ammonia feed. The reaction gas mixture entering the absorber is 95% oxidized to nitrogen peroxide and 27% dimerized.

In the absorber, deionized water is added to the top tray and weak acid from the low-pressure condenser is added to a tray corresponding to its acid strength. Down-flowing acid and up-flowing gases alternately mix as the chemical steps of action formation and nitric oxide oxidation take place with the accompanying release of heat. There are three operational zones in the absorber: the lower zone cooled with deionized cooling water; the middlezone cooled with chilled water; and the upper zone which is essentially adiabatic. Highly efficient heat removal in the middle and lower zones is particularly important due to its effect on the oxidation and dimerization reactions. The three-stage cooling system in the absorber provides good operational control over the unit. For this design, chilled cooling water and make-up water (both at 7°C) are used. The chilled water for the absorber is supplied by an absorption refrigeration unit, as a result the tail-gas exit temperature is approximately 10°C.

Acid from the bottom of the absorber is bleached at 1010 kPa using partially cooled, compressed air. The bleach air, containing nitrogen peroxide stripped from the acid, is then added to the main reaction gas stream at the oxidation vessel.

The cold tail gas leaving the absorber is warmed by heat exchange with the hot secondary air stream on route to the bleaching column. The tail gas is then heated to the expander inlet temperature of 620°C by passage through two further exchangers in the recovery train. The expander recovers 80% of the required compressor power. Expanded tail gas at 290°C then flows through an economizer providing heat to the high-pressure boiler-feed water and to the low-pressure deionized water. Subsequently, tail gas is exhausted to the atmosphere at 210°C and containing less than 1000 ppm of nitrogen oxides.

The cooling water circuit provides river water (at 20°C) to the steam condenser and compressor intercooler. The use of stainless

steels from the cooler/condenser onward has necessitated the use of deionized water for all other cooling applications. Any chloride contamination causes costly corrosion problems.

Mild steel is suitable for the construction of all the process units up to the cooler/condenser (see Figure 3.2), however, at this stage it is possible for nitric acid to be present in either liquid or vapour form in the reaction mixture. Stainless steel grade 304L is used for the cooler/condenser and all subsequent equipment. Grade 304L is resistant to corrosion by nitric acid, due to extensive nitriding during preparation and its extremely low carbon content. This stainless steel grade is also suitable for machining. A summary of the physical and mechanical properties of SS304L is given in Appendix D.

During the ammonia oxidation reaction, platinumand rhodium are lost in the form of volatile oxide vapours. Plant experience indicates that the loss from a single-pressure converter operating at 937°C is approximately 0.384 g/tonne (Ref. PTI; p.179). Due to this phenomenon, two platinum filter systems have been installed. The first is a gold/palladium alloy gauze placed immediately after the catalyst gauze in the reactor. The gold/palladium alloy forms a complex with the volatilized platinum and rhodium so that up to 70% of the volatilized catalyst may be recovered. The second platinum filter is a ceramic, alumino-silicate filter located between the waste-heat boiler and the tail-gas preheater. This second filter is able to trap 50% of the remaining precious metal catalyst for recovery at a later stage. However, the catalyst must still be reworked every five to seven weeks.

4.2.3 Requirements of Major Process Units

The major equipment items required for the single-pressure process are listed in Table 4.1 and the process flowsheet is presented in Figure 4.1.

4.2.4 Mechanical Design Features of Major Units

This section is included to outline some of the basic mechanical design features of the major process units as a preliminary to further design work presented in Part II.

FIGURE 4.1 Process flow diagram for a 'single-pressure' nitric acid plant.

TABLE 4.1

Schedule of major equipment

Unit	Function
Ammonia storage vessel	500m3 for one week supply of feed.
Ion-exchange unit	Provision of deionized water.
Deionized water cooler	Cool circulating deionized water.
Air filter	Remove solid particulates from feed air.
Two-stage air compressor	Provide feed air at 1090 kPa.
Ammonia vaporizer	Evaporate liquid ammonia feed.
Ammonia filter	Remove solid particulates from feed.
Ammonia superheater	Heat ammonia vapour to approx. 180℃.
Feed mixer	Mix the two gas-feed streams.
Reactor	Perform the oxidation of ammonia in air.
Waste-heat boiler	Cool reaction gases and produce steam.
Vapour/liquid separator	Remove entrained liquid as feed for the waste-heat boiler.
Steam superheater	Cool reaction gases and superheat steam.
Tail-gas preheater	Cool reaction gases and preheat tail gas.
Tail-gas warmer	Provide first stage of tail gas preheat.
Cooler/condenser	Condense out weak acid.
Oxidation unit	Final oxidation of reaction gases.
Secondary cooler	Cool reaction gases prior to absorption.
Refrigeration unit	Cool make-up water and provide some cooling.
Absorber	Absorb nitrogen oxides to form nitric acid.
Bleaching column	Remove dissolved NOx to give product acid.
Nitric acid product tank	2000m³ storage of one week supply of product.

Note   (a) The waste-heat boiler, tail-gas preheater, cooler/condenser, and oxidation unit form a sub-system of the process.

         (b) The operating parameters for each unit are given in Section 7.4: Mass and Energy Balances.

Note Heat-transfer coefficients for heat-exchanger design are calculated using the nomographs and typical values suggested in Refs. TD6 and TD7. This enables the determination of a preliminary heat-transfer area using the formula:

$$A = \text{Rate of Heat Transfer}/(U \times LMTD)$$

where the rate of heat transfer (W) is calculated in Appendix F;
U = overall heat-transfer coefficient $[W/(m^2\ K)]$;
A = heat-exchange area (m^2);
LMTD = logarithmic mean-temperature difference (K).

The formula has excluded the correction factors which account for the mode of flow. The exchanger areas calculated (in Appendix E) are estimates only, as required for Section 6.1: Capital Cost Estimation.

Process Units

1. Ion- Exchange Unit   This unit would probably service all three plants (ammonia, nitric acid and ammonium nitrate) in the chemicals complex, and would consist of a series of packed beds containing various organic polymer resins for the removal of unwanted divalent and monovalent ions.

2. Deionized- water Cooler   A finned fan-type cooler ,would be sufficient to provide the estimated 300 kW cooling duty.

3. Air Compressor   Air is compressed in twostages. The first-stage compression is a low-pressure compression from atmospheric pressure up to 310 kPa. An axial compressor is used which takes its shaft drive from a steam turbine. The second compression utilizes a centrifugal-type compressor. The centrifugal compressor is more efficient for the air flowrate (36 000 kg/h) and outlet pressure (1090 kPa) desired (Ref. TD5, p.581). The centrifugal compressor takes its shaft drive from the expansion of tail gas. Intermediate to the two compression stages is an intercooler which allows the air temperature to be lowered from 180°C to 45°C, with a pressure loss of 10 kPa. The temperature drop enables a more efficient second compression stage. Power calculations for the compressor are included in Section F.2.1 (Appendix F).

Compressor intercooler area = 1 30m².

4. Ammonia Vaporizer   A shell and tube-type heat exchanger with two passes per shell on the tube side. This unit should contain internal baffles. Operating pressure is 1240 kPa, with a design pressure of approximately 1400 kPa. This exchanger is made from mild steel.

Ammonia vaporizer heat-exchange area = 83m².

5. Ammonia Superheater   A shell and tube-type heat exchanger of

similar mechanical construction to the ammonia vaporizer. Also constructed from mild steel.

Ammonia superheater heat-transfer area = $30m^2$.

6. Reactor   The reactor is a pressure vessel operating in the range 1050 kPa to 1100 kPa. The design pressure should be about 1400 kPa. The vessel must be designed to ensure even passage of the feed gas mixture over the platinum/rhodium catalyst gauze. A 'Random Pack' flow distributor (Englehard Industries) would be suitable. The catalyst gauze and accompanying platinum filter gauze are fixed in position by lateral supports across the width of the reactor. The catalyst gauze and filter gauze are both 80 mesh, plain weave-type sheets (about 25 of each for a total mass of 24 kg). In the reaction section, the walls of the reactor should be lined with refractory material to prevent wear on the outer walls.

The bottom section of the reactor is jacketed. Air is preheated in this jacket prior to mixing with ammonia. The bottom section of the reactor also contains a shell and tube-type heat exchanger. This exchanger provides the final stage of tail-gas preheating. Tail gas enters at 235°C and the reaction gases leave the exchanger section of the reactor at 645°C. The reactor shell, jacket and associated heat exchanger are all constructed from mild steel.

Reactor exchanger heat-transfer area = $72m^2$.

7. Steam Superheater   This unit superheats saturated steam from 250°C (and 4000kPa) to 380°C. The product steam is of medium pressure and suitable quality for 'in-house' application and also for export. The superheater cools the reaction gases from the reactor exit temperature of 645°C to 595°C. Design pressure on the shell side is approximately 5000 kPa. The steam superheater is constructed from mild steel.

Steam superheater heat-transfer area = 1 $5m^2$.

8. Waste-heat Boiler   A shell and tube-type exchanger required to heat pressurized (4000 kPa) hot water from 117°C to a saturated vapour at 250°C. Design pressure on the tube side is approximately 5000 kPa. The waste-heat boiler cools reaction gases from 595°C to 280°C. It is made from mild steel.

Waste-heat boiler heat-transfer area = 11 $0m^2$.

9. Tail-gas Preheater   A shell and tube-type exchanger. It takes reaction gases leaving the platinum filter at about 315°C and 1020 kPa, and subsequently reduces their temperature to 185°C. The

cooling medium is tail gas. It enters at about 50°C and leaves the tail-gas pre-heater at 235°C. The design pressure for this unit is approximately 1200 kPa. It is constructed from mild steel.

Tail-gas preheater heat-transfer area = **89m²**.

10. Cooler/Condenser   This unit is the first to be constructed of type 304L stainless steel. It condenses weak nitric acid from the gaseous mixture and cools the remaining gases from an inlet temperature of 185°C to 60°C. This shell and tube-type heat exchanger uses deionized water as its cooling medium. It has a design pressure of about 1200 kPa.

Cooler/condenser heat-transfer area = **97m²**.

11. Oxidation Unit   The oxidation unit is an empty pressure vessel that takes input reaction gases and blends in additional air from the bleaching column. The extra oxygen provided enables further oxidation to occur and raises the gas mixture temperature to 140°C. At the top of the oxidation unit is a mist eliminator to prevent carry over of acid vapour by entrainment. At the bottom of the vessel is the weak-acid drain. The oxidation unit is constructed from SS304L and has a   design pressure **of 1200 kPa**

12. Secondary Cooler   The secondary cooler takes the exit gases from the oxidation unit at 140°C and cools them down to 65°C, a suitable temperature for entry into the absorption column. It is a shell and tube-type heat exchanger constructed of SS304L. The cooling medium is circulating warm water from the warm-water loop. The inlet temperature is 50°C and the exit temperature is about 80°C. The design pressure for this unit is about 1200 kPa.

Secondary cooler heat-transfer area = **140m²**.

13. Absorber   The absorber is usually a sieve tray-type column. It has a design pressure of 1200 kPa and an operating pressure around 990 kPa. A bursting disc is used for pressure relief. Each tray is provided with cooling coils to allow the cooling of the absorption liquor. There are two independent cooling circuits, each using deionized water. The top section has an inlet temperature of 7°C and an outlet temperature of 20°C. The bottom section cooling loop has an inlet temperature of 20°C and an exit of 40°C. The use of two cooling circuits provides greater flexibility in manipulating absorption conditions in the column. The tail gas leaves the column at about 10°C. Weak acid from the cooler/condenser is added to an appropriate tray midway up the column, and make-up water at 7°C is

added to the top tray. The acid drained from the bottom of the column contains some dissolved nitrogen oxides. The column and its cooling circuits are all constructed from SS304L.

14. Bleaching Column The bleaching column is a smaller sieve tray-type column. Impure acid runs down the column from the top tray and air is bubbled up through the liquor to remove dissolved nitrogen oxides. The acid from the base of the column is the final desired 60%(wt.) product.

4.2.5 Process Flow Diagram

The process flow diagram for the plant described in this case study is presented in Figure 4.1. It maintains many of the operating temperatures and pressures used in the C&I Girdler process, however, there are some significant changes in the flowscheme. First, the hot-water loop has been eliminated. This has effectively reduced the duty on the refrigeration unit, but instead requires more cooling water to be used at the compressor intercooler. This situation is preferred because the cooling water in this section is freely available with only the cost of pumping to be incurred. This pumping cost is offset against the cost of cooling-water processing in the refrigeration unit, together with the increased capital cost for a unit with a larger cooling duty.

The second modification to the flowscheme is that steam condensate from the ammonia vaporizer has been fed directly into the vapour/liquid separator, from where it may be drawn for waste-heat boiler feed. This route is in preference to feeding the condensate (at 250°C) back into the deionized-water circuit. By re-routing to the vapour/liquid separator, the heat duty of the waste-heat boiler can be reduced, and hence its size and cost are lower. The duty on the deionized-water fin-fan cooler and the economizer throughput are reduced. Both of these modifications result in a lower capital cost.

4.2.6 Process Performance Assessment

The process described in this section promises to be an efficient operation with the opportunity to become a net-energy producer.

Almost 6000 kg of steam can be produced every hour (see Appendix F for calculations). This leaves around 3200 kg/h for export after the process requirements are satisfied.

Increased consumption of ammonia due to the higher pressure operation can be minimized by a higher degree of ammonia oxidation. This is achieved by an added emphasis on complete ammonia/air mixing, and uniform flow distribution across the catalyst gauze. The higher oxidation temperature results in an increased consumption of platinum and rhodium and the necessity to rework the gauze every five to seven weeks (Ref. PTI; p.177). However, the higher temperature and the favourable effect of pressure make possible a greater recovery of energy from the process.

For an efficient absorption tower design, the plant should emit tail gases at less than 1000 ppm of nitrogen oxides. This level is about half the current emissions limit in Western Australia. Should emissions exceed this figure, then consideration must be given to the installation of a catalytic tail-gas combustor which enables emission levels to be lowered below 400 ppm. The plant does not normally produce any liquid waste streams (see Section 5.4.7: Environmental Impact Analysis).

The use of deionized water for cooling in the later process units is a costly, but necessary, requirement because of the use of stainless steels. By using a deionized-water loop the cost effect may be minimized.

CHAPTER 5
Site Considerations

5.1 Site Selection

MANY TEACHING institutions that undertake design projects omit consideration of the site selection for the chemical plant. The IChemE(UK) design project does not specify this aspect of the design to be considered, requiring only that the detailed equipment design for the plant is performed. However, in order to make the design project realistic and to provide the student with experience of as many aspects of design as possible, some consideration should be given to this topic.

In order to make the acquisition of relevant information easier, it is preferable to consider site selection on a local basis. If no local markets exist for the chemical or there are no areas suitable for chemical plant construction, it will be necessary to consider industrial regions elsewhere within the country or the state. Maps showing areas designated for heavy industrial development and the location of existing plants are usually available from government departments. However, even though an area has been designated for development does not mean that permission will be given to construct a new chemical plant, or that the location is actually suitable for the proposed plant. This comment can also apply to existing chemical plant sites, and in some cases further development of a site may be restricted by the government due to public opinion, environmental pollution problems, or concern regarding the risk of a major accident in a highly industrialised area. Workers usually prefer to live near their employment, and housing usually develops near industrial sites which were originally uninhabited, e.g. Bhopal! This can then lead to strong public opinion against further development.

Assuming that areas for industrial development can be identified, it is necessary to select a particular site. This should not be considered a 'one horse race'. Several sites need to be considered and the merits and

disadvantages of each site should be itemised. The final outcome will be a list of suitable sites in decreasing order of preference. To select one site and neglect all others leaves the project vulnerable to the dictates of government.

Several factors influence the selection of a site for the location of a chemical plant. The following list contains a few of the important considerations, but should not be considered exhaustive.

(a) Designation as a heavy industrial development area ('light' industry usually means assembly of electronic components, small metal fabricators, etc., and a major chemical plant would not be acceptable).
(b) Prior existence of similar chemical plants and location of other industrial centres.
(c) Existing roads and services, e.g. electricity, gas, water, etc.
(d) Appropriate terrain, sub-surface, drainage, etc.
(e) Suitable access for transportation of raw materials and chemicals, and for construction of a chemical plant.
(f) Proximity to major transportation networks, e.g. roads, railways, airports, waterways, ports. This is a major consideration in the location of a plant. In some cases direct pipelines for the transportation of chemicals or utilities (e.g. water, gas, oil) may be the most economical method. The cost of transportation by tanker (road, rail or sea) is reduced if a return load can also be carried.
(g) Availability of a local workforce and distance from local communities.
(h) Availability of domestic water and plant cooling water.
(i) Environmental discharge regulations.
(j) Proximity to both the raw materials supply and the market for the product chemical.
(k) Existence of services equipped to deal with a major industrial accident.
(1) Climatic conditions, e.g. humidity, maximum wind velocity and its prominent direction, rainfall, etc.
(m) Proposed or possible government restrictions regarding industrial development or discharge emissions.
(n) Room for expansion.
(0) Price of land.
(p) Public opinion.
(q) Possibility of earthquakes, subsidence, avalanches, etc.

(r) Availability of government regional development grants or tax incentives,   subsidies,   etc.

Several of these factors are discussed in more detail in Backhurst and Harker (1973; pp.374377); Baasel (1976; pp.2549); Coulson and Richardson Volume 6 (1983; pp.713-716);   references are given in Appendix L here.

Many other factors could be included, the most important factors are those specific to the requirements of a particular process or peculiar to a certain site. The three major site location factors are the location of raw materials, location of markets, and transportation. Once these aspects have been determined, other important (but secondary) considerations can be identified and evaluated. However, even if many of the basic requirements for a chemical process are satisfied by a particular site, the site may still not be acceptable if certain important criteria cannot be fulfilled, e.g. lack of government approval due to adverse public opinion, environmental regulations, availability of labour, etc. Therefore, it is necessary to consider the process of site selection as an overall package of *essential* requirements — all of which must be satisfied by the preferred site location, and achievement of as many secondary requirements as possible. Final site selection (i.e. obtaining a listing of site preferences) may be helped by listing all the features (required and existing) for each site, and assigning a subjective number value (say 1 to 5) for each feature. Each location can then be given a final points rating. Some discretion must be exercised because serious disadvantages must not be obscured by this approach. For example, a site that was otherwise ideal (or top of the list of preferences) but which had no local water available would be of little use as a chemical plant complex!

Action: ***Prepare a list of suitable sites for the construction of a chemical plant, arranged in descending order of preference.***
List all the merits and disadvantages *of* ***each site, and explain in detail why the first site is preferred.***
State any factors that need to be monitored throughout the project, e.g. government policy.

References

Baasel (1976; Chapter 2) and Peters and Timmerhaus (1980; Chapter 3). Both chapters include detailed reference lists.

5.2 Plant Layout

Having selected a suitable site for the chemical plant, it is possible and necessary to make a preliminary decision regarding the layout of the plant equipment. Although the equipment has not been designed in detail, preliminary estimates of the physical size of each item should be available in the equipment list (Chapter 4). Any sizing differences between the initial and final estimates should not be too excessive, and appropriate areas should be allowed around the plant items when determining the layout.

   A   preliminary   determination   of the plant layout enables consideration of pipe runs and pressure drops, access for maintenance and repair and in the event of accidents and spills, and location of the control room and administrative offices. The preliminary plant layout can also help to identify undesirable and unforeseen problems with the preferred site, and may necessitate a revision of the site selection (Section 5.1). The proposed plant layout must be considered early in the design work, and in sufficient detail, to ensure economical construction and efficient operation of the completed plant. The plant layout adopted also affects the safe operation of the plant, and acceptance of the plant (and possibly any subsequent modifications or extensions) by the community.

   There are two schemes that can be adopted for determination of the plant layout. First, the 'flow-through' layout (or 'flow-line' pattern) where plant items are arranged (sequentially) in the order in which they appear on the process flowsheet. This type of arrangement usually minimises pipe runs and pressure drops (and is often adopted for small plants). Second, the equipment is located on site in groupings of similar plant items, e.g. distillation columns, separation stages, reactors and heat exchanger pre-heaters, etc. The grouped pattern is often used for larger plants and has the advantages of easier operation and maintenance, lower labour costs, minimising transfer lines and hence reducing the energy required to transfer materials. These two schemes represent the extreme situations and in practice some compromise arrangement is usually employed. The plant layout adopted depends upon whether a new ('grass roots') plant is being designed or an extension/modification to an existing plant. Space restrictions are the most common constraints, however, space limitations are usually imposed even with new sites. Other factors to be considered are:

(a) Siting of the control room, offices, etc., away from areas of high accident risk, and upstream of the prevailing winds.

(b) Location of reactors, boilers, etc., away from chemical storage tanks.

(c) Storage tanks to be located for easy access, and a decision made as to whether all tanks (for raw materials and product) should be located together or dispersed around the site.

(d) Labour required for plant operation.

(e) Elevation of equipment.

(f) Requirements of specific plant items, e.g. pumps.

(g) Supply of utilities, e.g. electricity, water, steam, etc.

(h) Minimising plant piping systems.

(i) Suitable access to equipment requiring regular maintenance or repair.

(j) Plant layout to facilitate easy clean-up operations and dispersion of chemicals in the event of a spillage.

(k) Access to the plant in the event of an accident.

(l) Siting of equipment requiring cooling water close to rivers, estuaries, etc.

(m) Location of plant waste and water drainage systems (separate or combined?) and treatment tanks.

(n) Adopting a plant layout that will act to contain any fires or explosions.

(0) Spacing between items of equipment (insurance companies specialising in the insurance of chemical plants have specific recommendations for the distances required between particular items of equipment).

Particular factors that need to be considered in the plant layout stage of the design are discussed in more detail in Backhurst and Harker (1973; pp.377-391; Coulson and Richardson Volume 6 (1983; pp.718-724); and Baasel(l976; Chapter 6).

The layout of plant equipment should aim to minimise:

(i) damage to persons and property due to fire or explosion;
(ii) maintenance   costs;
(iii) number of plant personnel;
(iv) operating costs;
(v) construction   costs;
(vi) cost of plant expansion or modifications.

Some of these aims are conflicting, e.g. (i) and (iv), and compromises are

usually required when considering the plant layout to ensure that safety and economic operation are both preserved. The final plant layout will depend upon the measures for energy conservation within the plant (see Section 8.2.1) and any subsequent modifications, and the associated piping arrangements.

Action: **Prepare a preliminary plant layout, explaining the reasons for particular proposals.**

References

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Baasel (1976; Chapter 6); and Peters and Timmerhaus (1980; Chapter 3); detailed reference lists are also included in these chapters.

Mecklenburgh (1985) — see Appendix L here for full reference.

5.3 Environmental Impact Analysis

This topic needs to be considered at an early stage in the design, in relation to the site considerations, and reassessed during the detailed design stage (Chapter 8). The effect of the operation of the chemical plant upon the environment and the population **must** be considered during the feasibility study and during subsequent design stages. The formal environmental impact analysis *(EIA)* has two parts, these relate to:

(a) the treatment of unwanted chemicals (by-products) and the concentrations of liquid discharges and gaseous emissions during normal operation;
(b) the handling of a major chemical accident, including all chemicals within the plant and any subsequent reaction products, and containment and clean up.

The environmental impact analysis is specific to the plant under consideration, the comments/suggestions made here are of a general nature.

5.3.1 General Considerations

All waste chemicals from the plant must be disposed of in an acceptable manner. Dumping of the waste may not be allowed or, if it is, it may be prohibitively expensive. Some form of treatment, e.g. dilution, neutralisation, purification, separation, etc., may be necessary prior to disposal. It is necessary to determine whether it is more economical (and preferable for the efficient operation of the plant) to perform this treatment within the chemical plant itself. Consideration should also be given to the installation of separate drainage systems from certain sections of the plant and for particular wastes, e.g. rainwater, domestic waste, relatively pure process water, acid spills,. oils, etc. Separate drainage systems are best installed during the plant construction stage. This scheme will be more expensive than a central collection tank for all liquid waste, but it will allow separate treatment and possible recycling of   liquids.

The concentrations of all chemicals to be discharged, including gaseous emissions, must be determined and measures taken to ensure that these levels conform to allowable legislative standards. The cost of clean-up can be high, e.g. scrubbing systems, filters, etc., and may affect the economic analysis of the plant operation (Chapter 6). It is prudent to ensure that not only are present emission standards observed, but also that the plant could conform to any subsequent legislative reductions in these emission levels (while still operating economically).

The effects of all emissions on the environment, and upon company employees and the population must be assessed. The possibility of future litigation, say in 10 years time, should not be ignored. Noise levels from the plant need to be considered, and finally its aesthetic acceptability.

The second part of the environmental impact analysis (EIA) relates to the effects of a major accident or spill within the plant. The safety aspects of an explosive gas discharge (for example) should be considered in conjunction with the loss prevention studies for the plant (Section 8.2.3). However, proposals for containment, clean-up, and discharge of major chemical spills should be part of the EIA report. Any proposals should ensure the safety of personnel, minimise the discharge and its effect on the environment, and preserve the integrity of the plant. The worst situation should be evaluated, not just the most likely scenario. Factors to be considered include the quantity and location of chemicals

stored within the plant, prevailing wind direction, location of plant personnel and the general public, access to the plant and to particular high-risk   areas.

There is a large amount of information available concerning chemical discharges, environmental protection, clean-up and the applicable legislation. The Environmental Protection Agency in the USA has produced a prodigious number of reports concerning a wide range of chemical plants. Much of the available information is specific to particular chemicals and process routes, however so much information has been published that locating a relevant source is usually the main problem. Most countries now require a company to file an environmental impact assessment with any proposal for a new chemical plant. This document is usually open to public scrutiny and can be a source of much useful information. Copies of the relevant legislation are readily available and advice can be obtained from appropriate government departments.

Remember that however attractive a particular chemical plant appears (technically and economically), public and political opinion especially in relation to the environmental issues can prevent the project proceeding, or seriously reduce its economic feasibility.

5.3.2 EIA Policy and Scope

Environmental Impact Assessment (EIA) has evolved as a comprehensive approach to project evaluation, in which environmental factors, as well as economic and technical considerations (e.g. Cost Benefit Analysis), are given appropriate consideration in the decision-making process. The purpose of an EIA study is to determine the potential environmental, social and health effects of a proposed development. It attempts to define and assess the physical, biological and socio-economic effects, so that logical and rational decisions are made. The identification of possible alternative sites and/or processes may assist in the reduction of potential adverse impacts.

There is no general and widely accepted definition of EIA and although many organisations may share common objectives, the concepts and scope of an EIA may vary considerably. The results of an EIA are assembled into a document known as an *Environmental Impact Statement (EIS)* which contains a discussion of beneficial and adverse

impacts considered relevant to the project, plan or policy. In some countries and states, the draft EIS must be made available for public inspection, debate and comments, and a final EIS may be required to include relevant public opinion and discussion.

An EIA should improve the efficiency of decision-making by ensuring that subjectivity and duplication of effort are minimised, and by avoiding the short-term and long-term consequences of inappropriate decisions. However, in order to be effective an EIA should be implemented in the early stages of project planning and design, and it must be an integral component in project design rather than a technique to be utilised after the design stage is completed. An EIA procedure should be applied to all actions likely to have a significant environmental effect.

The responsibility for undertaking an EIA can be allocated in several ways, the choice depends upon the particular EIA system planned or operating, and the nature of the project. The four major alternatives (although others exist) that are commonly used to assign responsibility are: the authorising agency; the developer; shared responsibility between the authorising agency and the developer; or an independent specialist body. The cost of an EIA study can also be allocated (or recovered) in a variety of ways. The impartiality of an EIA may be achieved in several ways, for example:

(a)   Establish guidelines or minimum standards for the form and content of an EIA.
(b)   Supervision by a reviewing or controlling body with no vested interest in the project.
(c)   Mandatory consultation with relevant and competent organisations.
(d)   Publication and provision for public discussion of the impact statements.

5.3.3 EIA Reports

Many different EIA manuals are available (the majority produced in the USA) which provide advice about identifying, predicting and displaying potential impacts. Manuals often contain advice beyond these basic aspects including the structure and scope of the assessment, responsibility for the study, review procedures, implementation,

monitoring, etc. Manuals are available dealing with specific types of developments; the EIA and resulting EIS requirements will depend upon the particular type of proposed project. The following discussion relates to general guidelines for performing an EIA study and preparation of the EIS document, it should not be considered to apply in every (or any) particular situation.

(i)   Acquisition of information on a proposal

The developer should be invited to prepare a *short* brief concerning the proposal(s) including a list of key siting criteria, brief details of the proposed development, and the processes to be used when operational. Information available may be limited at this stage and the discussions are usually of a feasibility/evaluation nature.

As the project design and planning proceeds, the developer is asked to complete a detailed **Project Specification Report** containing the following details:

Physical characteristics of application site
Employment requirements
Financial data
Infrastructure requirements
Factors of environmental significance
Emergency services
Hazards

Information can also be obtained by a site inspection, consideration of existing planning policies, and consideration of similar proposals.

(ii)   Identification and assessment of possible impacts

It is necessary to establish the nature of the existing environmental, social and economic conditions in the area surrounding a proposed development. The possible (or probable) impacts can then be identified and their implications assessed. Information may be required concerning the physical characteristics of the site and its surroundings, local human activity patterns, infrastructure services, social and community services, existing levels of environmental pollution, etc.

For each possible impact it is necessary to undertake an analysis of the scale and significance of the potential change. The environmental

consequences of a project will not all be similar in degree or kind — they may tend to cancel each other out or to reinforce each other. The assessment should therefore distinguish between impacts which are:

Permanent/irreversible	or	temporary/reversible
Short term	or	long term
Local impact	or	strategic/widespread
	or	distant impact
Primary	or	secondary
Direct	or	indirect

It may also be necessary to assess not only the consequences of the project under consideration, but also any subsequent actions which become inevitable (or more likely) as a result of the proposal proceeding. Quantitative assessment of the environmental impact should be presented whenever this is possible, even though the available numerical techniques often vary in their precision.

(iii) Presentation off findings

The conclusions of the EIA study regarding the potential impacts are presented in the EIS report. The EIS should include:

A brief description of the proposed development.
A brief description of the local area.
Potential   impacts.
Mitigating actions.
Examination of effects on the area if existing trends were to continue.
Consultations and objections.

Some consideration of the probable situation without the proposed development is useful, and should include:

Employment   prospects,
Dynamic   environmental   factors.
Changes due to planning policies,

The findings of the draft EIS report can be interpreted and summarised in a *Major Issues Report,* thus providing a shorter and simplified summary of the options. Both reports should be available for

public scrutiny and comment, feedback from all interested parties can provide valuable information and should be considered (and included, if appropriate) in the preparation of the final EIS report.

5.3.4 Australia

In Australia, Federal (Commonwealth), State and local government all have responsibilities and powers in relation to environmental assessment and protection. If proposals are jointly financed by a State Government and the Commonwealth, then each government shares the responsibility for the EIA. The Commonwealth Environment Protection (Impact of Proposals) Act (December 1974) provides for the assessment of the environmental effects of proposals so that these factors are included with the consideration of economic, technical, financial, and other factors, before decisions are made. The Act requires environmental examination of all proposals which involve the Commonwealth Government, and which have a significant effect on the environment. The Act contains administrative procedures for the early provision of appropriate information to the Minister responsible for the Act, in order to decide whether an EIA statement is required. The State Governments have major responsibilities for the environment. In several States, legislation is being developed to provide support for impact statement requirements.

5.3.5 United Kingdom

Development projects in the UK are subject to a wide range of controls and consultations under which permission is needed from public authorities prior to commencement. Particular permissions required depend upon the scale of the project, the activity or use for which it is intended, the detailed design, the location, and other factors. Planning control and pollution control, including hazard control, are of primary importance. The assessment of environmental impacts is achieved through existing legislation for planning and pollution control on development projects, including appropriate changes to the existing systems where necessary. New (and specific) legislation has not been implemented, partly due to the flexibility of the existing system.

No significant development of any kind can be undertaken in the UK without the prior permission of a public authority. The local authority normally considers such proposals, although sometimes the decision is

made by central government. Public hearings are often held for major development projects and a final decision is made by the appropriate Secretary of State. Pollution and hazards of various kinds are subject to separate controls and all major sources of pollution are subject to authorisation procedures, e.g. air, water and solid wastes, and noise nuisances.

5.3.6 United States

The National Environmental Policy Act *(NEPA)* of 1969 has shaped the US environmental protection efforts. It has asserted a collective social responsibility for the quality of the nation's environment, and has provided:

(a) a general policy statement of Federal environmental responsibility;
(b) formation of the Council on Environmental Quality within the Executive Office of the President;
(c) requirements and guidelines for preparation of environmental impact statements.

All Federal agencies are required to produce an EIS for any Federal action which significantly affects the quality of the human environment.
Under the NEPA, all Federal departments and agencies are required to improve, co-ordinate and orientate their planning and development programmes towards:

(i) minimising the long-term environmental effects of all Federal actions;
(ii) the right of society to a safe, healthy and aesthetic environment;
(iii) multiple use of environmental resources;
(iv) preservation of historic and natural landmarks;
(v) balanced population growth and resource utilisation;
(vi) recycling of scarce natural resources.

The Environmental Protection Agency *(EPA)* is an operating line agency responsible for administering and conducting all Federal pollution control programmes, focussing attention on pollution control as a strategy for securing environmental quality as well as preservation of wildlife and natural resources.

Under the guidelines for preparation of an EIS, the following must be included:

(a) a comprehensive technical description of the proposed action;
(b) an analysis of the probable impact on the overall environment;
(c) a description of any probable adverse environmental effects;
(d) analysis, studies and descriptions of possible alternatives to the recommended course of action, and their environmental effects;
(e) detailed consideration of any irreversible or irretrievable commitments of scarce environmental resources.

Many states have now passed legislation and adopted administrative actions to establish procedures similar to the NEPA.

Federal agencies issue permits for a wide range of activities, e.g., air and water emissions, dock construction, etc., covering most industrial projects. Therefore, virtually all major plant projects require an EIA. The company prepares the EIS (or has it prepared by a consultant) for the appropriate federal agency which then issues the EIS document. The preparation of an EIS report requires detailed knowledge of the plant site and its surrounding territory. It must also include a great many aspects, e.g., air and water quality, visual aspects, impacts on local flora and fauna, etc. The EIS may be several hundred pages long (especially for a new plant), therefore this section provides only an introduction and covers a few of the more important points to be considered.

Action: Prepare an environmental impact analysis for the proposed chemical plant, including aspects *of* normal operation and major spills.
Detail the relevant legislation affecting   this plant.

References

Detailed reference lists are included in Peters and Timmerhaus (1980; Chapter 3, pp.89-95, 110-l 12).

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5.4 Case Study — Site Considerations

Summary

Nitric acid and the nitrogen oxide gases present in the process are highly toxic. It is imperative that the choice of site selection and plant layout reflects the potential hazard from both of these sources.

The nitric acid plant will be part of a chemical plant complex that will include an ammonia plant, a nitric acid plant and an ammonium nitrate plant. Consideration is given to the needs of all three of these plants in the site selection procedure.

The site chosen for the nitric acid plant, as a part of the larger chemical plant, is a 10 hectare plot in the Bunbury district of Western Australia. The site is located between the port handling facilities (for alumina) belonging to Alcoa of Australia Ltd, and Leschenault Road. The site runs parallel to the Preston River and is less than 1 km from the main port area. It features easy port access, adjacent rail facilities, and an abundance of river water (for process cooling purposes).

The layout suggested for the chemical plant complex includes a central control room to operate all three plants. Administration, laboratory, and workshop areas are also common. The nitric acid plant is small, occupying less than 1 hectare. There is space on the 1 hectare plot for inclusion of a second parallel process train for possible future expansion.

Normal operation should be well within the environmental regulation limits set by the EPA. Liquid waste is virtually non-existent and can be sent to the normal sewerage drains. Any acid spills should be diluted. Tail-gas emissions are thought to be less than 1000 ppm of nitrogen oxides (about half the current EPA limit). Should tail-gas emissions exceed this figure, then a catalytic combustor would be necessary to reduce nitrogen oxide levels to below 400 ppm.

References Used

General Introduction: G8.

5.4.1   Site Considerations — Introduction

There are several aspects to be considered regarding the siting and operation of the nitric acid plant. First, a suitable site must be chosen and second the plant layout must be planned after the site characteristics are assessed. Finally, an environmental impact analysis needs to be performed to ascertain the expected effect of the plant and the chemicals on the surrounding areas.

5.4.2   Site Selection

There are a number of considerations concerning the choice of site locations for a new nitric acid plant within Western Australia. Some of these are general considerations whilst others relate directly to the process and its requirements. Those considerations relevant to this study include:

(a)   Close proximity to the anticipated market.
(b)   An availability of fundamental infrastructure (including port facilities and both rail and road access).
(c)   Utility costs and availability.
(d)   A suitable local labour force.

   In the determination of a site location for this nitric acid plant, a broader spectrum of factors must be considered. This plant will almost certainly have to be part of a larger chemicals complex involving the production of ammonia (for nitric acid plant feed) and ammonium nitrate (to use most of the nitric acid product). A site of about 10 hectares is required.

   After assessing all these factors, there are several possible options for the site location. These include a series of industrial belts within the Perth metropolitan region and three regional centres (Bunbury, Geraldton and Karratha). Figure 5.1 shows the location of these areas within Western Australia. Both metropolitan and country areas offer promising initial opportunities. These are now investigated in more detail.

5.4.3   Perth Metropolitan Region

The primary constraint when choosing a suitable site location within
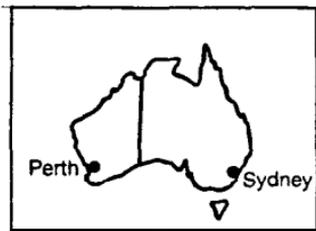
Perth to:
Bunbury 175km
Kalgoorlie 600km
Karratha 1450km
Sydney 4200km
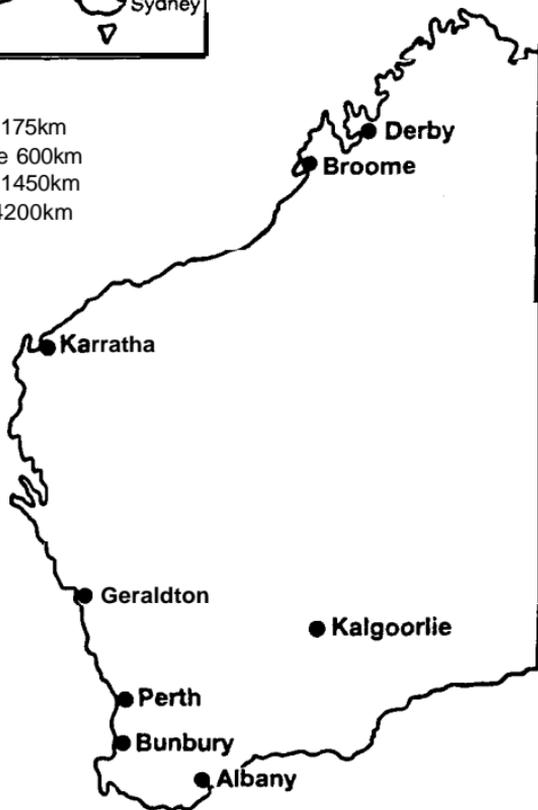
FIGLXE5.1   Possible nitric acid plant locations and major towns in Western Australia.

the Perth metropolitan area is the restriction governing land usage as laid down by the Town Planning Authority. Their zoning schemes designate separate zones for use as residential, commercial and industrial applications. A chemicals plant must be sited in an industrial zone. Within the 'industrial' category there are several sub-groupings. Plants utilizing hazardous chemicals (such as nitric acid), or potentially hazardous processes, are categorized under the title of 'Special Industry'. There are four Special Industry areas at present

within the Perth environs. These are South Fremantle, Kwinana, Viveash and Jandakot (see Figure 5.2). The South Fremantle region has no room for expansion and Jandakot is not due to be released until the 1990s. Therefore, only the other two regions will be considered.

The industrial belt of Canning Vale is also considered (see Figure 5.2). Normal industrial belts may be used for chemicals manufacture through direct negotiation with the relevant city council, and with permission from the Environmental Protection Authority (EPA). Canning Vale is a relatively new industrial area that has already accommodated several large processing facilities, including the new Swan Brewery and the glassworks of Australian Consolidated Industries (ACI). Also present are several fibreglass resin and paint manufacturers. The Canning City Council was contacted and the Chief Town Planner saw no particular difficulties in locating a nitric acid plant and its associated works in this region. For these reasons Canning Vale is also assessed.

1. Kwinana

The proposed site location is adjacent to Paterson Road, and approximately 1 00m south of Claymore Street, running adjacent to the Kwinana Nickel Refinery (KNR) of Western Mining Corporation.

Advantages
- accessible by road, rail and sea
- seawater available for cooling in the other plants in the complex (not used in the nitric acid plant due to use of chloride-susceptible stainless steels)
- availability of scheme water for drinking and for the process uses (needs pre-treatment)
- natural gas available
- close to KNR which is a possible potential market for excess steam produced in the process
- labour market availability
- removed from residential areas.

Disadvantages
- Kwinana industrial strip is already crowded, with any new
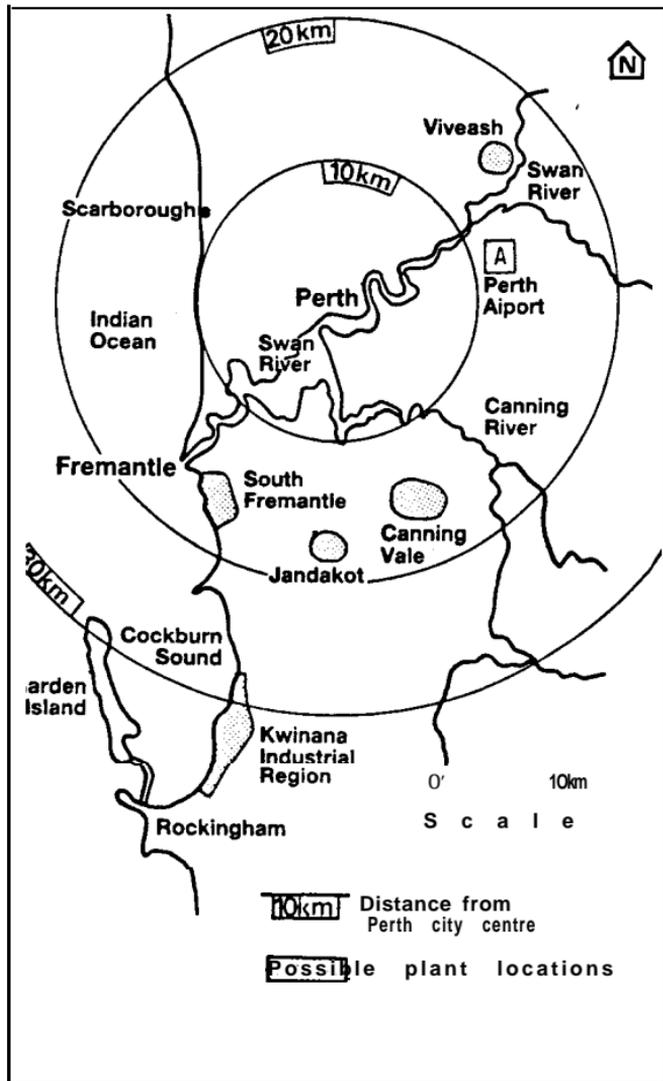
FIGURE 5.2 Possible nitric acid plant locations in the Perth metropolitan area.

proposals receiving strong opposition from a growing' local lobby

— the explosion risk to the ammonia complex from fires/explosions at adjacent KN R

the site is particularly marshy and would require an extensive and rigorous site preparation so that good plant foundations can be laid.

2. Viveash

The proposed site location in Viveash is along the east side of the Swan River, and to the south of Middle Swan Road.

Advantages
- readily accessible by road and rail
- cooling water from the Swan River (not for use in nitric acid plant)
- scheme water
- natural gas
- labour force availability.

Disadvantages
- no port nearby
- quite close to residential areas (not ideal as far as the accompanying ammonia plant is concerned)
- inconsistent with present planning in the area, and may well meet with considerable local government opposition.

3. Canning Vale

The proposed site location may be selected from a number of vacant lots in this industrial estate along Baile Road, and adjacent to the Swan Brewery.

Advantages
- road and rail access
- scheme water and natural gas
- potential market for excess steam (the brewery and the glassworks)
- optimistic response from Chief Town Planner at the Canning City Council
- close enough to labour force but far enough removed from residential areas.

Disadvantages
- the lack of a convenient supply of cooling water, however, an agreement may be possible with the Swan Brewery
- no port facilities.

5.4.4 Country Districts

Several country locations are also considered in an effort to find a superior location to those suggested in the Perth metropolitan region.

1. Bunbury

The site considered for the new chemical complex containing the nitric acid plant is immediately between the Alcoa alumina handling facilities and Leschenault Road. There is an area of approximately 20 hectares of which 10 hectares are required. The site is less than 1 km from the Bunbury port facilities and also has one border along the Preston River. Figure 5.3 provides further details.

The Preston River represents a suitable supply of process cooling water. The railway line servicing the Bunbury dock area runs past the site, providing rail despatch facilities to the plant.

Advantages
- port facilities
- road and rail access
- natural gas available
- cooling water available (sea or river)
- a good local labour market
- no anticipated objection from local council
- removed from residential areas.

Disadvantages
- added transport costs in shipping/railing any of the products to Perth (more applicable to ammonium nitrate plant).

2. Karratha

Karratha has several key advantages relating to its port facilities, proximity to the South-East Asian market and direct access to North-West Shelf natural gas (both for heating and for producing ammonia synthesis gas). However, the disadvantages of increased labour costs (above the 26th parallel) and poor labour availability, together with the large engineering costs in site preparation, tend to make this option unattractive.
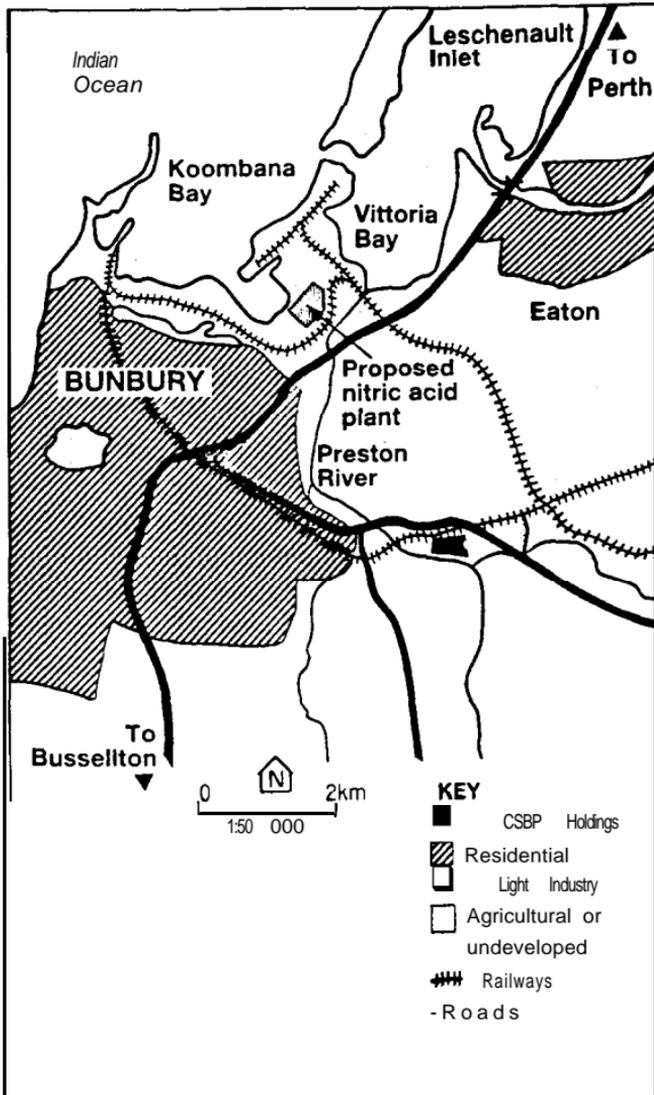
FIGURE   5.3   Proposed Bunbury   site for nitric acid plant.

3 .   *Geraldton*

Geraldton offers port facilities and cheap land, but the problems with labour availability also tend to make this option unattractive.

5.4.5 Site Location Conclusions

The philosophy of *in situ* consumption of much of the product nitric

acid, together with the remaining product aimed directly at the export market, both relieves the need for a metropolitan site location and also makes the need for port facilities of paramount importance. A Kwinana location is not imperitive since the ammonia plant will almost certainly not be taking its feed from an oil refinery.

These factors essentially determine the choice between the Kwinana and Bunbury sites. Both sites are considered suitable, however, the former has been disregarded because of the present controversy and difficulty being experienced in gaining new-site approval in an area already crowded. The Bunbury site is therefore selected as a totally satisfactory solution.

5.4.6 Plant Layout

There are several plant layout considerations relevant to the Bunbury site. First, the prevailing wind direction, shown with a dark arrow in Figure 5.4. This would indicate that the tank farm is best placed on the river side of the site to avoid vapours drifting back across the plant area. The second major consideration is the need for rail access to the ammonium nitrate despatch/storage area.

A suggested layout for the overall complex is shown in Figure 5.4. It features the three chemical processes flowing approximately one to the other. The joint administration and workshop facilities are located in the south-western corner of the plant. There is a central control room for all operations, and the rail link is extended into the despatch area.

The nitric acid plant is relatively small and should not occupy more than 1 hectare. Even this small area should leave plenty of room for expansion. Figure 5.5 shows the equipment layout for the nitric acid plant. The plant layout includes space for a parallel production train should the market expand sufficiently in thefuturesuch that a second plant is required.

The layout shown in Figure 5.5 also offers plenty of room around each unit for maintenance work, with units remaining close enough for process streams to be transferred easily.

5.4.7 Environmental Impact Analysis

The final consideration of this section, and by no means the least
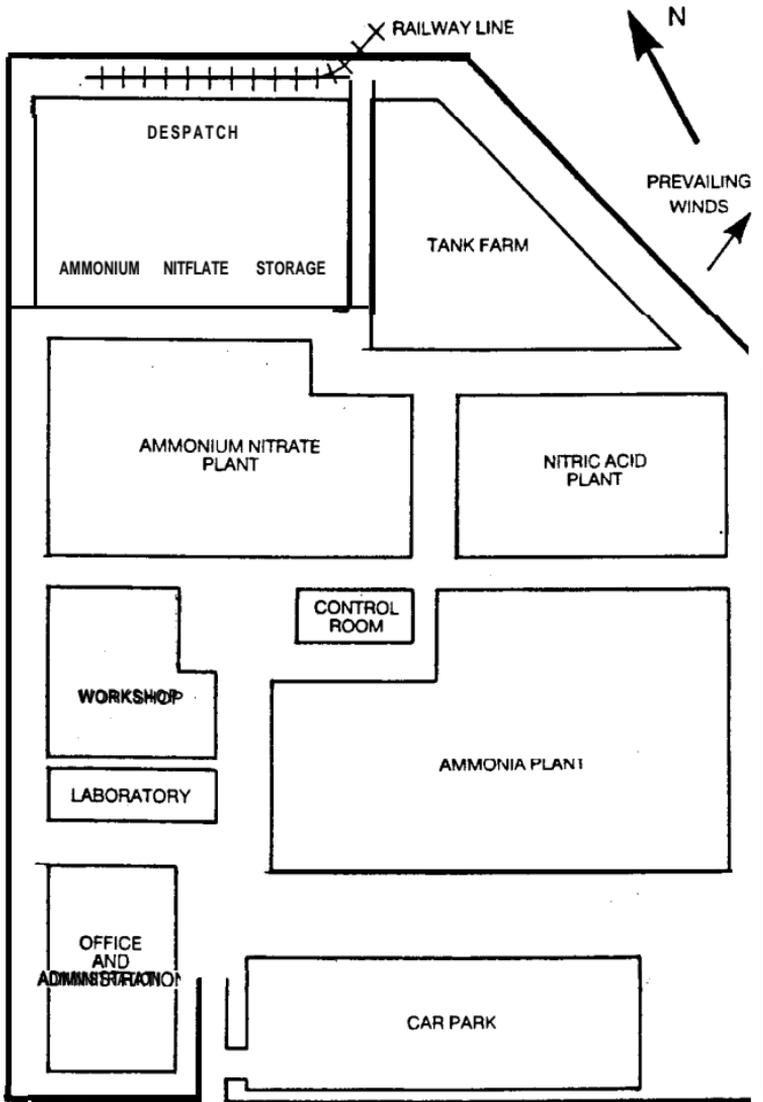
FIGURE 5.4   Overall   chemicals-complex   layout.

important, is the environmental impact study. Associated with nearly every chemicals plant are unwanted by-products which must be disposed of both safely and economically. This necessitates a compromise between the safety demand limits and tolerances, and the costs of treating wastes to a level beyond regulation stipulations.

Fortunately the single-pressure nitric acid plant is a remarkably clean plant. It has no liquid waste streams in normal operation and
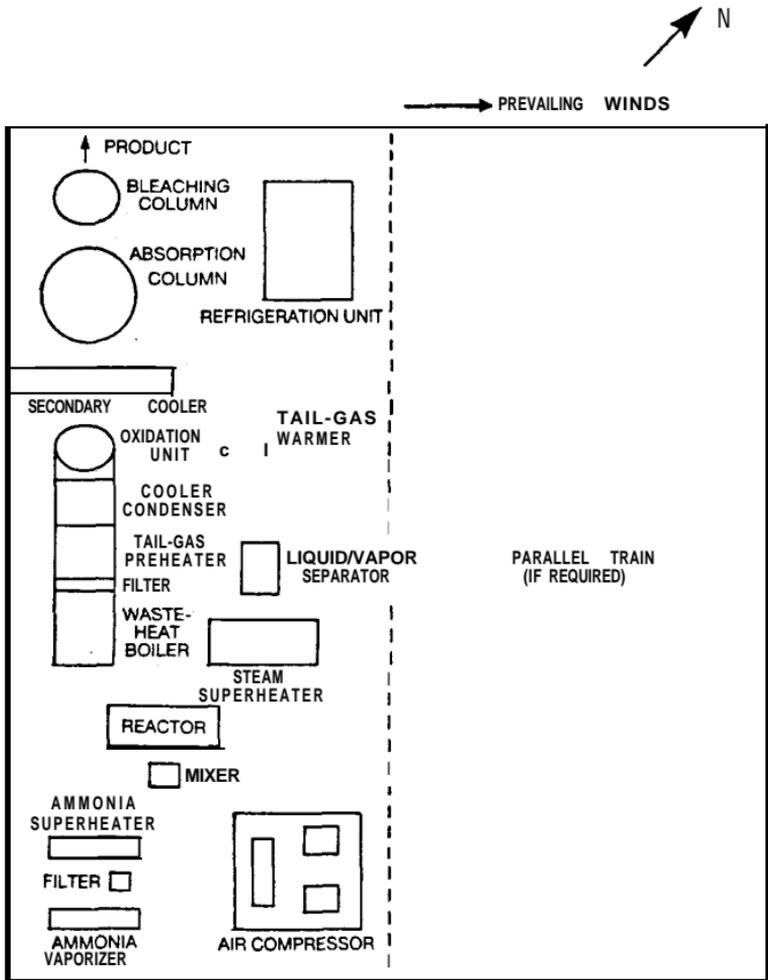
FIGURE 5.5   Nitric acid plant layout.

the gaseous emissions are limited to process tail gas. This cleanliness is the result of working with pure feed streams and highly efficient absorption stages.

The other environmental impact consideration is noise. The air compressor with its tail-gas expander and steam-turbine drive could present a problem. Therefore, this unit should be enclosed in a brick building to contain the noise, with those working inside the building required to wear approved ear protection.

In the industrial emissions standards, an item of concern is the levels for nitrogen oxides. Reference G9 itemizes each of the limits for

waste-water streams and waste-air streams, as set by the Environmental Protection Authority (EPA). The limit for dissolved nitrogen oxides in water is 0.20 g/m^3 of liquid waste. The only foreseeable liquid waste of this nature would be drainage from the ground following maintenance on equipment. The concentrations attained from this source would not approach the limit set by the EPA.

The limit for the nitrogen oxide component in gaseous streams is 20 g/m^3 (STP). The design specification for tail-gas emissions is set at less than 1000 ppm nitrogen oxides. This figure of 1000 ppm is less than half the EPA limit. Should the normal tail-gas emissions be seen to greatly exceed the design specifications, then a catalytic combustor would be required. This unit enables the reduction of nitrogen oxide levels in tail gas to less than 400 ppm. The economic viability of this project would not be affected by the installation of such a unit.

In the event of a major plant accident, large quantities of nitrogen oxides could be released into the atmosphere and product nitric acid may well be released to the drains. There is little that can be done to provide for such a situation. The emphasis would be on dilution of any liquid wastes expelled. It would be left to the prevailing winds to disperse the nitrogen oxide cloud. Fortunately this gas cloud is not flammable, but it is highly toxic. Nitrogen oxide fumes either from nitric acid or from process gases may have a deceptive delayed action. The victim may feel no pain or discomfort at the time of the low-level inhalation, or for up to 48 hours thereafter, but then serious respiratory and cardiac problems can occur.

For a large gas release, the areas in the path of the cloud may well have to be evacuated. Workers sent in to make repairs or assess damage should be equipped with air-tight, acid-proof clothing and a self-contained breathing apparatus.

The threshold limit for nitric acid is only 2 ppm. A few breaths of 200 to 700 ppm may be fatal within 5 to 8 hours. Anyone exposed to nitrogen oxide fumes should be observed for 48 hours. It is obviously essential that a tall, well designed stack be employed to aid in the effective dispersion of tail gas.

CHAPTER 6
Economic Evaluation

6.1 Introductory Notes

THE ECONOMIC evaluation of a chemical plant covers a wide range of topics. The scope is much too broad to allow inclusion of a detailed description in this short section, indeed entire books have been written on this topic alone. Sections 6.2 to 6.4 contain brief notes covering the main topics to be considered in the introductory stages (feasibility study) of a design project.

Several excellent reference sources are available, those written by and for chemical engineers are especially useful. The most prominent text in this field is by Peters and Timmerhaus, 1980 (see reference in Appendix L here). The following chapters, containing over 200 pages, are included in that text:

4. *Cost and Asset Accounting;* presents a survey of accounting procedures for the analyses of costs and profits as used for industrial operations.
5. *Cost Estimation;* provides information regarding the estimation of fixed capital costs, and also recurrent operating expenditure.
6. *Interest and Investment Costs;* discusses the concept and calculation of interest, i.e. payment as compensation for the use of borrowed capital.
7. *Taxes and Insurance;* taxes represent a significant payment from a company's earnings and although insurance rates are only a small fraction of annual expenditure, adequate insurance cover for a plant is essential.
8. *Depreciation;* this is a measure of the decrease in value of an item with respect to time and can be considered as a cost incurred for the use of the equipment.
9. *Profitability, Alternative Investments, and Replacements:* the profitability of an investment is a measure of the amount of profit

generated. It is important to assess the profitability accurately, and also the profits that could be obtained from alternative investments.

Other useful chemical engineering design books containing chapters for the economic evaluation of projects are those by Baasel (1976); Backhurst and Harker (1973); and Ulrich (1984). The advantages of the books by Baasel (1976); Peters and Timmerhaus (1980); and Ulrich (1984) are the inclusion of extensive references to the literature and substantial compilations of economic data. The IChemE(UK) has published the following guides:

Economic Evaluation of Projects by D.H. Allen
Project Procedure by J.C. Rose, G.L. Wells and B.H. Yeats
A New Guide to Capital Cost Estimating, report of a working party.

The book by Ulrich (1984) contains the following chapters in the section covering economic analysis: Chapter 5 describes capital cost estimation; Chapter 6 discusses procedures for evaluating plant operating costs (and interpretation/preparation of balance sheets); Chapter 7 covers techniques for determining the optimum design from several alternatives; Chapter 8 discusses profitability analysis, discounting to present day values, and determination of the viability of a project. It should be remembered that government taxation policy varies widely between countries, and it can have a significant effect on project profitability.

The first task is usually to obtain a capital cost estimate for the plant, remember that different methods have different errors. A pre-design estimate (mentioned in Section 2.1: Initial Feasibility Study) based upon capacity corrections and the use of cost indices, when applied to data for an existing plant, will have an accuracy of $\pm$ 20-30%. A more accurate original cost estimate can be obtained either from a supplier or by the use of tabulated data in the literature (see Ulrich, 1984; pp.281-323). This data is usually presented graphically as capacity (or size) plotted against purchase price for particular types of equipment. Even at this stage of the design project, it may still be necessary to wait until the detailed design of certain equipment is available before final and accurate (within 5%) cost estimates can be completed. This aspect of the design work cannot be completed at this stage in the project.

The reader can refer to Ulrich (1984) for more detailed descriptions of these topics, especially pages 279-280 containing a review of the capital cost estimation procedure. A more detailed (although still necessarily brief) discussion is included here in Sections 6.2 to 6.4, the references quoted provide an excellent in-depth coverage of this subject.

6.2 Capital Cost Estimation

During the feasibility stage of a design project, approximate cost estimates ($\pm$ 20-30%) are required — and usually in the minimum time possible. Cost correlations and factored estimates are usually sufficiently accurate for the initial cost estimation study, and they significantly reduce the calculation time required for more detailed estimates. Detailed (and more accurate) cost estimates are usually required after the detailed design work has been completed, including the design and sizing of all equipment, determination of pipework layouts, and specification of the control and instrumentation schemes. The additional time and effort (and cost) required to produce a more accurate ($\pm$ 5-10%) cost estimate is rarely justified in the feasibility stage of the design.

6.2.1   Cost of Equipment (Major Items)

Major items of equipment include reactors, heat exchangers, columns, vessels, etc. Ancillary equipment such as process piping and insulation can be estimated after the total cost of the major items is known.

I Cost   correlations

Cost correlations provide a convenient method of estimating the capital cost of major items of equipment. Correlations are usually provided graphically as plots (log-log coordinates) of capital cost of a particular item versus capacity (e.g. volume, surface area, throughput, or power rating). Even at zero capacity, there is some cost (e.g. overheads) associated with the equipment. The cost (C) increases to infinity (i.e. slope of the curve = I), at which point it is more economic to use multiple units of the same size. For an intermediate capacity (Q) range,

the slope *(m)* of the curve is approximately constant and the following convenient equation can be used: $C = C_B (Q/Q_B)^m$, where C_B and Q_B are for a unit of basic size. Useful data can be obtained from: A *Guide to Capital Cost Estimating,* (IChemE, UK).

The cost of equipment or of a complete plant must be up-graded to account for the reduced purchasing power of the dollar (or pound) from a given time datum to the present. Varicus cost indices are published annually (and monthly, *see Chemical Engineering* magazine). Those of particular relevance to chemical plant costs are the Nelson Refinery Construction Index, Chemical Engineering (CE) Plant Cost Index, and the Marshall and Swift (M & S, previously Marshall and Stevens) Equipment Cost Index. These indices apply to complete plants rather than individual items of equipment. The appropriate equation is: $C_2 = C_1 (I_2/I_1)$, where Zis the relevant index, and the suffixes represent different time periods.

Values of the *CE* and M & S indices are tabulated in Baasel (1976; p.238, Table 9.1) for the period 1953-1973. The *CE* Plant Cost Index is composed of four components having the following weightings:

Equipment, machinery and supports	0.61
Erection and installation labour	0.22
Building materials and labour	0.07
Engineering and supervision manpower	0.10

It is possible to modify the index to represent a particular situation by changing the component weightings, and hence recalculate the values of the modified index for particular years. An alternative index is the Process Engineering (PE) Plant Cost Index, values are published monthly in *Process Engineering* journal.

The cost of a complete plant can be estimated using the 'six-tenths rule' which states that: 'the ratio of the costs of two plants producing the same product is proportional to the ratio of their capacities raised to the power of 0.6'. This statement can be written as: $(C_1/C_2) = (Q_1/Q_2)^{0.6}$, the accuracy of such an estimate is $\pm$ 30%. Using this rule, if the capacity is doubled then the cost increases by 52%. This emphasises the economic advantages of building very large plants, provided duplicate plant items are not required. However, the effects of a major accident (or failure of an item of equipment) are more serious in a single large plant than for two smaller plants!

II   *Factored   estimate   method*

The complete plant cost can be estimated by extrapolation from the cost of the major items of equipment. An accuracy of $\pm$ 15-25% is possible if appropriate factors are chosen. The technique is described by the equation:

$$C_F = C_E (1 + \Sigma f_i) (1 + \Sigma f_i')$$

or

$$C_F = C_E F_L$$

where C_F is the fixed investment cost of a complete plant; C_E is the cost of the major items of processing equipment; f_i are factors for the direct costs e.g. piping, instruments, etc.; f,' are factors for the indirect costs, e.g. contractor's fees, contingency allowance   etc.

The value of the *Lang factor* (FL) depends upon the type of process. For most plants it has a value of approximately 3. The following values can be used as a guide:

F_L = 3.10 for solids processing plants
F_L = 4.74 for fluids processing plants
F_L = 3.63 for a mixed fluids-solids processing plant

If possible, the Lang factor should be obtained from the cost files of the particular company. It is important (although sometimes difficult) to define the type of plant under consideration because of the large difference between the values of F_L (3.10 to 4.74).

The complete cost of a plant can be estimated by using a particular value of F_L, this can be done in the early stages of the design when the preliminary flowsheet and equipment list are available. Greater accuracy can be achieved by considering the individual cost factors (f_i and f_i') that are compounded into the Lang factor (F_L). Lists of direct and indirect costs and typical values of their factors are given in Coulson and Richardson, Volume 6 (1983; pp. 187-l 89). In the early stages of a design study some items may be overlooked, this results in a low estimate of the capital cost. A contingency allowance of between 10% and 50% (depending upon the completeness of the information) is included to account   for   this   possibility.

6.2.2 Module Costs

An estimate of the plant capital cost can also be obtained by considering the plant as a set of modules. Each module consists of similar items of equipment, e.g. heat exchangers, steam generators, etc. The standard cost of each process module is calculated (for a particular year), corrections are applied for the use of materials other than carbon steel and for high pressures. Variations in the process piping cost are corrected in the contingency allowance. A modified Lang factor is calculated for each module, including freight costs and sales taxes. Site development and structures are considered as separate modules. This method is described in more detail in the paper by Guthrie (see References).

6.2.3 Auxiliary Services

It is necessary to include the full cost of auxiliary services, e.g. steam, water and electricity supply and distribution, roads, buildings, communications, etc., in the capital cost estimate for a new plant. If a chemical plant is to be constructed on an existing complex, then a proportion of the cost of existing services (based upon the estimated consumption) is usually charged. A very approximate estimate of the cost of auxiliary services can be obtained by taking 20-25% of the total installed (new) plant cost. However, in some situations a figure of 40-70% may be more appropriate. It must be emphasised that a more accurate estimate should be obtained during the preparation of the feasibility study.

6.3 Operating Costs

The plant capital cost represents a one-off expenditure, although the capital will usually have to be repaid over several years. In order to assess the economic viability of a project it is also necessary to estimate the operating costs which are incurred annually in the production of the chemical. The operating costs should be considered when the alternative process routes are being evaluated, and they can significantly influence the final choice that is made. Operating costs can be divided into two groups (although in some cases the division is arbitrary), these are:

(a) *Fixed* costs such as laboratory costs, operating labour, capital repayment charges, insurance, etc. These costs do not depend upon the production rate, and they must be paid even if no chemical is produced.
(b) *Variable* costs such as raw materials, utilities (services), shipping, etc. These costs are dependent upon the amount of chemical produced.

The plant supervisor/manager has no control over fixed operating costs, whereas he/she is held accountable for the variable costs. In addition to those costs incurred due to the construction of the plant and/or its operation, each plant, site or product is usually required to contribute towards the general operating expenses of the company. These expenses include general overheads, research and development costs, sales expenses, etc. Each company decides how these costs are apportioned, however as a general indication they may add 20-30% to the direct production costs at site.

   The following items represent the more common operating costs, although the list should not be considered complete for any plant. Operating costs are usually calculated on an annual basis, and subsequently calculated per tonne of product (for example) when determining the profitability and selling price of the chemical.

(a) *Raw materials* are determined from the process flowsheet and from material balances.
(b) *Miscellaneous materials* include items such as safety clothing, chart recorder paper, etc., that are not included as raw materials or maintenance materials. These are usually calculated as 10% of the total maintenance   cost.
(c) Utilities (services) include electricity, water, steam, compressed air, etc. Quantities are determined from the flowsheet and from energy balances, current costs (and anticipated price rises) should be obtained for these items.
(d) *Shipping and packaging costs* depend on factors specific to the process, the location and the type of product. In some cases these costs are negligible, but sometimes they can be significant.
(e) Maintenance includes materials and labour costs. This cost is typically between 5-15% of installed capital cost and should be estimated from data for a similar plant.
(f) *Labour   costs* should be estimated from reasonably detailed manning estimates. The operating labour costs may not decrease if production is

reduced, however overtime payments will be required for significant increases in production. Operating labour costs do not normally exceed 15% of the total operating cost (most plants employ few personnel).

(g) *Supervision* includes the management team directly responsible for the overall plant operation and for directing the work of the plant operators (item (f) above). The personnel requirements of the management team should be determined, although an approximate figure of 20% of item (f) can be used to provide an initial estimate.

(h) *Laboratory costs* for analysis associated with quality control and process monitoring. An approximate estimate can be obtained as 20-30% of the operating labour cost (item (f) above), or 2-4% of the total production cost.

(i) *Plant overheads* include general operating costs such as security, canteen, medical, administration, etc. This item is often estimated as 50-100% of item (f).

(j) Capital charges are recovered from the project to repay the initial capital investment. The procedure adopted depends upon the accounting practice of the company. Capital is often recovered as a depreciation charge of 10% per annum (for example) based on a plant operating life of l0 years, although the plant is not necessarily replaced after that time! Interest must also be paid on the capital borrowed to finance the plant. The capital may be obtained from company reserves, but it should be repaid with interest-based upon a consideration of the alternative investments and their return on capital, i.e. applicable market rates.

(k) *Rates* are payable to the local authority or shire based upon the assessed rateable value of the site. A typical figure is l-2% of the capital cost.

(l) *Insurance* for the site, the plant and employees is usually about l-2% of the fixed capital.

(m) *Royalties and licencefees* are payable to the company or individual responsible for developing the process. This payment may be a lump sum or an annual fee, and is typically either l-5% of sales price of 1% of the fixed capital.

More detailed discussion of these items and a worked example can be found in Coulson and Richardson, Volume 6 (1983; pp. 194202).

6.4 Profitability Analysis

The first stage in evaluating the profitability of a proposed new product is to compare the total cost of the product (for example, per tonne produced) with the current market price. It is also necessary to estimate future demand for the product, and to determine the trend in the selling price over several previous years. It is also necessary to estimate the number of years that the product can be sold at a satisfactory profit, this is more useful than estimating the possible operating life of a plant! Backhurst and Harker (1973; p.47) make the following observation:

'...it is suggested that a large chemical organisation will not invest in a new process unless it is possible to sell the product for less than half the current market price'.

Two other measures that can be used to evaluate the profitability of a product are the return on investment and the payback period. The return *on investment (ROZ)* is the expected profit divided by the total capital invested, expressed as a percentage return. It must be clearly stated whether the profit is based on pre-tax or after-tax earnings. The after-tax ROI is compared with the earnings that could be achieved by an alternative investment, e.g. capital bonds. An after-tax ROI of at least 15-20% is usually expected (or 30-40% pre-tax ROI), assuming that the project is not particularly risky!

The payback period(or payout time) is the number of years from plant start-up required to recover all expenses involved in a project, if all the pre-tax profits were used for this purpose. Depreciation charges are not included in the operating costs. Expenses not incurred directly in the design and construction of the plant are excluded, the analysis is intended to demonstrate the best means of allocating the present and future resources of a company. A payback period of less than five years is usually required for a project to proceed. However, the payback period does not consider the timing of the payments or the profits earned by the plant after the payback period.

For more detailed discussion of plant economics and profitability analysis, the reader should refer to a specific textbook such as Peters and Timmerhaus. The following quotation from Backhurst and Harker (1973; p.47) provides an interesting summary:

'The viability of investment lies, sad to say, mainly in the hands of the economist and the financial expert, and it is important to realise that an increase of) per cent in the bank rate has probably far more effect on the profitability of a project than an increase of, say, 10 per cent in the efficiency of a distillation column'

Action: *Perform a detailed economic analysis of the project including:*
             Capital cost estimates (by different   methods)
             Operating   cost   estimates
             Profitability   analysis
             Re-assess the economic feasibility of the project based upon the complete   economic   analysis.

References

Detailed reference lists are included in the relevant chapters in Baasel(l976); Peters and Timmerhaus (1980); and Ulrich (1984).

Holland, F.A.; Watson, F.A., and Wilkinson, J.K., Engineering Economics for Chemical Engineers, a 20-part series in *Chemical Engineering,* from 25 June 1973 (p. 103) to 20 October 1974. The series is available in total as *Chemical Engineering* Reprint, no. 215 (1975).

Guthrie, K.M., Data and Techniques for Preliminary Capital Cost Estimating, *Chemical Engineering,* 24 March, pp.1 14-142 (1969).

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Humphreys, K.K., and Wellman, P., *Basic Cost Engineering,* 2nd Edn, Marcel Dekker, Inc., New York (1987).

Modern Cost Engineering: Methods and Data, Volume I (1979) and Volume II (1984), reprints of papers from *Chemical Engineering,* McGraw-Hill Publications, New York.

6.5   Case Study — Economic Evaluation

Summary

The scope of the economic evaluation includes estimation of capital cost figures for a nitric acid plant producing 280 tonnes/day of a 60% product. Following this capital cost estimation the total annual operating costs are estimated, both variable and fixed components being considered. Also considered is the cost of providing finance for the initial capital outlay at 25% interest per annum.

The capital cost is estimated by the ratio method at AS1 2.9 million.

The factored cost method produces a slightly larger figure of AS1 3.5 million. The two estimates were considered sufficiently close, and appeared consistent with actual plant costs detailed overseas.

The investigation into return on investment centres around several key assumptions. The initial capital is secured at 25% interest per annum. Interest payments and operating costs are deducted from product sales to obtain gross profit figures. This figure is subject to 49% company tax. The viability of the project is assessed with the requirement that the project show a return on investment within five years. Two scenarios were considered:

1. Supplying only a domestic market of 65 000 tonnes/year.
2. Adding an overseas market to bring the total production figure to 92 000 tonnes/year.

With the market price for nitric acid at A\$345/tonne, the results obtained are encouraging. Scenario 1 suggests a breakeven cost of A\$220/tonne for the first five years (while repaying capital) and AS1 17/tonne thereafter. Scenario 2 suggests a breakeven cost of A\$205/tonne for the first five years and AS1 31 /tonne thereafter. These figures indicate clear post-tax profits of over 30% on investment. The project appears to be highly feasible and well suited to cope with the cyclical Australian market.

References Used

Process Technology (PT): 1,2.
Cost Estimation (CE): 1,8,9,11.

6.5.1 Introduction

The economic evaluation is an important and integral part of the overall feasibility study of the project. First, a capital cost estimate is obtained using two estimation techniques. The *ratio* method and *factorial cost estimation techniques* are used to determine separate capital cost estimates for the proposed 280 tonne/day plant. Finally an investigation into the expected return on investment from this project is performed.

6.5.2 Capital Cost Estimation

Two methods are used to determine a figure for the estimated capital cost. First, the ratio method is applied to known capital cost figures for an existing plant, providing an update to a current cost. Second, the factorial method is used to estimate the total capital cost by estimating the anticipated fixed costs and then adding to that figure the value of the working capital.

(a) The Ratio Method

The ratio method is a simple technique whereby known capital cost data for an existing chemical plant are adjusted to provide a cost estimate for the desired plant capacity. This method is also able to update figures to account for inflationary effects of past years. Finally the capital cost figure is adjusted for exchange rate differences between countries. The method is centred around the use of key cost estimation indices such as the CE Plant Cost Index and the Marshall and Stevens (M&S) Index.

Several values for the capital costs of nitric acid plants have been located. They range from a 1967 figure of USS5 million for a plant producing 1000 tonne/day (Ref. CE1) to a 1985 figure of USS23 million for a 1800 tonne/day plant (Ref. CEI 1). The most relevant figure obtained, and the one used for this ratio analysis, is a 1979 value of USS12.2 million for a nitric acid plant producing 907 tonnes/day of 60% acid (Ref. PTI; p-1 81).

The method used is outlined in Ref. CE9 (Chapter 4). The capital cost estimate is adjusted to the correct plant capacity by applying a scaling index. The appropriate scaling index for nitric acid plants is 0.6 (Ref. CE9; Table 19, p.184). Therefore:

Cost of-280 t/d plant = (Cost of 907 t/d plant) $\times$ $(280/907)^{0.6}$
                   = US$6.0   million

The plant cost must be updated to account for inflation and technological advances over the 7 years from 1979 to 1986. This is achieved by the use of the CE Plant Cost Index (Ref. CE8; p.7). The cost is calculated from:

Present Cost = Previous Cost x (Present Index/Previous Index)

The present index is evaluated using a formula that applies a weighting for the various capital cost components. The weighting scheme is shown below:

	Weighting
Equipment, machinery, supports	0.61
Construction, labour	0.22
Buildings	0.07
Engineering and supervision	**0.10**
	1.00

These weightings are combined with numerical values given in Ref. CEB (Table 2, p.7):

Present Index = 0.61 (347.6) + 0.22(265.3) + 0.07(303.8) + 0.1 O(333.8)
              = 325.0

The past index is obtained from Table 1 of the same reference; it is 238.7 for the year 1979. Therefore, the present cost can be calculated as:

Present Plant Cost = US$6.0 million x (325/234.8)
                   = USS8.2 million
                   = AS1 2.9 million

The accuracy of the ratio method improves if more recent plant cost data are available. This figure is probably accurate to within + 30%. The exchange rate used was taken at 16th October 1986 (US$l = As0.6332).

(b)   The Factorial Method

The factorial method of cost estimation requires the purchase cost of the major equipment items in the process. Other fixed costs are

estimated by applying indices to this estimate. These indices are expressed as factors of the equipment costs. The accuracy of this estimate depends upon the reliability of the data available for the equipment costs.

The basis for estimating the purchase cost of the heat exchangers was the area available for heat transfer. Factors were then applied to allow for different construction materials, operating pressures and tube dimensions.

The air compressor purchase cost was estimated on the basis of power consumption. The oxidation vessel and reactor costs were estimated using correlations appropriate for pressure vessels. Capacity formed the basis of the storage tank purchasing cost. The absorption and stripping column were costed according to diameter, operating pressure and number of trays.

The fixed capital cost is estimated by first assessing the purchase cost of the major equipment items listed in Table 4.1. These costs are calculated in Appendix E, and are shown in Table 6.1.

Installation costs for the various units are estimated using the data of Ref. CE9 (Table 6; p.169). This data expresses typical installation costs for each unit as a percentage of purchase price. Compressors are rated at 30% of purchase cost; heat exchangers at 35%; towers/. columns at 60%; metal tanks at 30%. Those items not included are estimated as having an installation cost at 30% of the purchase cost.

Total Installation Cost is estimated to be As870 000.

Other fixed costs are estimated using similar factors of the total purchase cost. Table E.l in Appendix E shows several other fixed costs to be considered, as well as the appropriate factor. These factors are compiled from values suggested in Ref. CE9 (Chapter 4).

Total Physical Plant Cost = [(I + ΣFactors) x Purchase Cost] +
                            Installation
                          = [(I + 1.56) x 2 623 300] + 870 000
                          = As7 586 000

Other factors are required to account for the engineering and contractors costs involved with construction. These factors relate to the total physical plant cost, A breakdown of these costs is shown in Table E.2 of Appendix E.

TABLE 6.1
Purchase cost of major equipment

Unit	Purchase cost (A$)
1st stage compressor	490 000
Compressor intercooler	20 500
2nd stage compressor	915 000
Ammonia vaporiser	23 100
Ammonia superheater	11400
Reactor	66 800
Steam superheater	7 500
Waste-heat boiler	29 000
Tail-gas preheater	24 300
Cooler/condenser	62 500
Oxidation unit	34 000
Secondary cooler	84 200
Absorption column	55 000
Refrigeration unit	55 000
Bleaching column	25 000
Ammonia storage tank	310 000
Nitric acid tank	410 000
Sub-total	AS2 623 300

Total Fixed Capital Cost = [(1 + ΣFactors) x Physical Plant Cost]
$$= (1 + 0.40) \times 7\,586\,000$$
$$= 10620000$$
$$\approx \text{AS1 1 million}$$

The total working capital must be added to this estimate of fixed capital to obtain the full composite figure for the capital cost. The working capital represents an investment in storage of feed ammonia for one week, product acid for one week, stores (e.g. catalyst) and others. Table E.3 in Appendix E summarizes these incurred costs.

Total working capital is estimated at As2.45 million.

The factorial method therefore yields a capital cost estimate of:
Capital Cost = Fixed Costs + Working Capital
$$= \text{AS1 1 million} + \text{As2.45 million}$$
$$= \text{AS1 3.5 million}$$

(c) Capital Cost Conclusions

The ratio method provides an estimate of AS1 2.9 million. This can be regarded as reasonably accurate ($\pm$ 30%) considering the original plant cost data is 7 years old. The factorial method has produced a surprisingly similar result. This is probably due to the fact that the plant is not particularly large, and the possibility of estimation cost inaccuracies is reduced. The estimate of AS1 3.5 million determined by the factorial cost technique should therefore also be regarded as an   acceptably   accurate   value.

6.5.3   Investment   Return

It is necessary to determine the minimum selling price for nitric acid that must be obtained in order to substantiate this project as being economically feasible. If this figure is below the market value for nitric acid, then due to the relatively stable consumption pattern envisaged, the project might be considered viable. If the project is found to be viable then the rate of return can be calculated based upon sales of nitric acid at market value.

   Viability is assessed according to a required maximum 5 year return on investment. All annual costs must be considered. These costs include the required return on the capital outlay, as well as daily production costs, annual sundry charges such as insurance, tax, maintenance charges, and interest on borrowed capital.

   Table E.4 in Appendix E provides a breakdown of the total production costs encountered in the manufacture of nitric acid. The costing is based upon paying the full market price for ammonia feed (at A\$300/tonne). All tangible input and output valuesarecalculated using the results of the mass and energy balances detailed in Section 7.3. Labour requirements are evaluated assuming only two operators per shift and the usual labour maintenance requirements for nitric acid plants (see Ref. CE1 1).

   Production costs are calculated to be A\$97.07 per tonne of 100% nitric acid produced. This corresponds to As59.30 per tonne of 60% nitric acid. This figure compares with a cost of As36 per tonne of 60% nitric acid (Ref. CE1 1; US\$22.70,   1985). This lower value might be attributed to several factors:

(i) Lower unit cost of ammonia-the market price for ammonia is around A$300/tonne,   however, in the context of a nitric acid plant forming part of a larger chemicals complex, the cost would be expected to be lower. Ammonia represents about 80% of the total production costs. The plant in Ref. CEI 1 received ammona at A$1 90/tonne.

(ii) Lower unit cost of labour — United States labour costs are fractionally lower than those in Australia.

(iii) Economies of scale — the plant in Ref. CE11 produces approximately 1700 tonnes per day of nitric acid compared to 280 tonnes per day in this design.

Other annual operating costs to be calculated are shown in Table E.5 of Appendix E. These figures sum to A$3.7 million.

Assume that the capital for the project is borrowed at 25% interest with a required return on investment of 5 years. Five annual payments are made to the creditors. The calculations made to determine the size of these payments and their repayment schedule are given in Appendix E (Table E.6). The five payments needed are each of A$5.02 million. These will cover both the 25% interest rate and the A$13.5 million principal. Operating and production costs are tax deductible as is the interest on borrowed capital, only gross profit is taxed. In the calculations, the A$5.02 million capital cost term is therefore broken into two components. The first is the tax-free interest component and the second must come from operating profit. This profit is assumed to attract 49% company tax.

Total Cost of Production
= Capital Finance Cost + Variable Production Cost + Fixed Operating Cost
$$= [3\ 375\ 000 + (1\ 645\ 000/0.49)] + [t\ (59.3)] + 3\ 700\ 000$$
$$= 6\ 732\ 150 + [t\ (59.3)] + 3\ 700\ 000$$
where t = annual production of 60% nitric acid.

Two scenarios have been considered to establish the stability of the project.

Scenario 1

The first scenario assumes that only the domestic market of 65 000
tonnes is available.

Total cost of production   =   6 732 150 + (65 000 x 59.3)
                                     + 3 700 000
                          = AS1 43 million

In order to cover these production costs the product must be sold at a
minimum price of:

Minimum sales price = AS1 43 million/65 000
                          = A$220/tonne

Scenario 2

The second scenario assumes that the full annual plant production
capacity of 92 000 tonnes is sold. The additional 27 000 tonnes must
include the shipping costs estimated at $120/tonne.

Total cost of production
= 6 732 150 + (92 000 x 59.3) + 3 700 000 + (27 000 x 120)
= AS191 million

In order to cover these production costs the product must be sold at a
minimum price of:

Minimum sale price = AS1 91 million/92 000
                          = A$208ltonne

Under both scenarios the plant offers an excellent opportunity for a
good return on investment given that the current market price is
As345 (US$200). Table 6.2 summarizes the economic evaluation
results.

TABLE 6.2
Economic evaluation conclusions

Fixed capital cost: Ratio method	A\$12.9 million
Fixed capital cost: Factored cost method	A\$13.5 million
Annual variable production cost	A\$59.3/tonne (60% acid)
Annual fixed operating costs	A\$3.7 million
Capital finance costs (for first 5 years only)	A%5.02 million

Scenario 1: Domestic market

In first five years:

Breakeven cost	A\$220/tonne of 60% acid
Clear profit after tax	A\$40 million annually
(A\$345/tonne)	(30% of investment)

In period thereafter:

Breakeven cost	A% 117/tonne of 60% acid
Clear profit after tax	A\$7.3 million annually
(A\$345/tonne)	(54% of investment)

Scenario 2: Both domestic and export markets

In first five years:

Breakeven cost	A\$208/tonne of 60% acid
Clear profit after tax	A\$6.3 million annually
(A\$345/tonne)	47% of investment

In period thereafter:

Breakeven cost	A\$1 3 1 /tonne of 60% acid
Clear profit after tax	A\$9.7 million annually
(A\$345/tonne)	(72% of investment)

CHAPTER 7
Mass and Energy Balances

7.1 Preparation of Mass and Energy Balances

MASS AND energy balances must be completed for each item of equipment before the detailed design of these items can begin. A mass balance is performed for each individual unit shown on the process flowsheet and included in the equipment list (Chapter 4). Sometimes an energy balance is performed over a group of units (or a section of the plant) where energy is being transferred between various process fluids, e.g. the feed preheating heat exchangers or the reactor and a subsequent cooler/condenser. After the energy balances have been completed it is possible to consider the efficient utilization of energy within the plant, and hence prepare an energy conservation scheme (Section 8.2.1).

Even at this stage in the design, it may not be possible to finalise all details of the mass and energy balances. A specification of performance should be available for each unit, but only estimates of certain flowrates and temperatures may be available. These mass and energy balances will need amendment and checking following the detailed and final design of each item of equipment. For example, it may be specified that a liquid-washing absorption tower is to remove at least a certain percentage of a gaseous product stream. It is usually possible to estimate the flowrate of liquid required, but this quantity will not be known accurately until the detailed design of the unit is completed. The design will consider the type and height of packing material required, and the liquid flowrate necessary for efficient operation of the tower. The final design may be capable of removing more gas than the minimum specified for the original estimated liquid flowrate (however, it is important that the unit designed does not remove less of the gas stream!).

A summary of the mass and energy balances should be presented in tabular form using a separate sheet of paper for each unit. This

information *must* be presented clearly and concisely. The calculations are detailed separately, often included as an appendix to the design report. The mass and energy balance for each unit should include the following details:

(a)Mass   flow of all streams into and out of the equipment per unit time. Molar or volume flowrates must not be used. A convenient and appropriate unit of time is chosen, e.g. per minute, hour, day, etc., so that the numbers being handled can be easily appreciated, e.g. between 0 and 1000, and as far as possible excluding the use of 10^3 or 10^{-6}, etc. All balances *must* be performed for the same time period.
(b) Composition (mass %) of all streams.
(c) Sometimes the molar flowrates (and mole %) of gas streams are also included.
(d) Temperature of each stream (use °C or K consistently; for temperatures below freezing point (0°C) absolute temperatures may be preferred).
(e) Pressure of each stream, or state the pressure once if no significant pressure   drop   occurs.
(fl Enthalpy content of each stream (J, MJ, GJ, etc.).

The tabular format for mass and energy balances should be familiar to students when they undertake a design project, otherwise refer to Himmelblau, 1982, p. 100 (reference at the end of this section) or Austin and Jeffreys, 1979 (see Appendix L here). The convention is to include information for streams entering the unit on the left side of the page, and for output streams on the right (flow from left to right, as on the P & ID). Each sheet of the mass and energy balances should be numbered consecutively, e.g. Sheet 1 of 32 (or l/32). The reference number for each unit (see Chapter 4) should also be included. If an energy balance is performed over several units, then the combined and summarised balance sheet may be presented. The total mass and total energy into and out of each unit should be given at the bottom of the left hand and right hand sides of the page. Hence, the use of the term 'ledger balance', this provides an immediate indication of whether a balance has been obtained.

Data such as flowrates, compositions, temperatures, etc., that are summarised in the mass and energy balances should be transferred to the process flowsheet and the equipment listing (noting any amendments from the original estimates). The process flowsheet should be

continuously updated as the project proceeds, and the date of the most recent amendments clearly printed.

The mass and energy balances should not be considered merely as a set of data for use in the detailed design of equipment, they should also be used and analysed. The mass balances can be used to show those areas of the plant having a high chemical requirement, or production, or process water requirement. Similarly, the energy balances can identify equipment with a high energy requirement or large surplus of energy to be removed. It is often useful to draw a simple block diagram showing particular sections of the plant, including the major mass flows and the heat output and energy requirements. Such a diagram enables an immediate appreciation of the operational requirements of the plant.

Note to students

Mass balances often do not balance exactly, some error is allowable (1%, 5%, l0%?). When an exact balance is obtained, this is usually because the flow of one stream has been obtained by difference. Therefore, an exact balance does not mean that the calculations have been performed correctly!

When presenting mass and energy balances, some discretion should be exercised regarding the number of significant figures quoted. An eight-figure calculator display does not imply eight-figure accuracy in the calculations — and such accuracy is not required. Quoting data to three significant figures (or even two figures) is usually sufficient. The inclusion of more figures merely makes the balance less clear and it becomes difficult to obtain an immediate impression of the situation. Students often seem unwilling to adopt this suggestion! However, an exception to this general rule is the calculation of quantities of impurities within the process streams. Low-levels of impurities may be present at different stages and if ignored they may build-up to unacceptable levels in recycle streams or in the final product stream. It is important to identify and calculate the amounts of impurities at each stage of the process.

Action: ***Preparation of individual mass and energy balances for each item of chemical engineering equipment (or heat balances over several items of equipment).***
Transfer of information to the process flowsheet and equipment list.

References

Peters and Timmerhaus (1980; p.50) for a reference list.

Himmelblau, D.M., ***Basic Principles and Calculations in Chemical Engineering,*** 4th Edn, Prentice-Hall, Inc., New Jersey (1982).

Houghen, O.A.; Watson, K.M., and Ragatz, R.A., ***Chemical Process Principles Part II,*** 2nd Edn, John Wiley and Sons, Inc., New York (1959).

7.2 Preliminary Equipment Design

Following completion of the mass and energy balances, preliminary equipment designs can be undertaken. The final equipment designs and specifications are completed after performing a detailed economic evaluation of the project and an energy conservation study (Section 8.2.1).

   The preliminary designs reassess and modify the data and assumptions made in earlier stages of the project. Any major modifications that become apparent and any changes to the plant or process design aspects are noted, and their effects upon the project feasibility are assessed. The specific type of equipment to be used (as detailed in the equipment list) is confirmed, the appropriate design method is determined, and all necessary data is obtained (hopefully). Any major differences from the original equipment specifications are investigated. At this stage the design engineer stops short of performing the detailed design calculations.

Action: ***Prepare detailed specifications   for all major plant items to be designed.***
Obtain all necessary design data,

References

Baasel(1976; Chapter 5); Peters and Timmerhaus (1980; Chapters 13-l 5); Ulrich (1984; Chapter 4); also refer to Aerstin and Street (1978); Backhurst and Harker (1973); and Coulson and Richardson Volume 6 (1983), details in Appendix L here.

7.3 Computer-Aided Design

(Program names in capitals are registered trademarks.)
The term ***computer-aided design (CAD)*** is widely used within engineering but is often interpretted in different ways. It should not be

confused with the use of computers to prepare engineering drawings via draughting packages, this is sometimes referred to as computer-aided design and draughting (CADD). For the chemical engineer engaged in a design project assignment, computers and CAD now have several common applications. These include the preparation of flowsheets and the P & ID, calculation of the mass and energy balances for the overall process and individual units, costing and economic analysis, the optimization of process schemes, and the detailed design of individual chemical engineering equipment, e.g. heat exchangers, distillation columns, etc. Although there is no generally accepted standard terminology, the detailed design of major units is best described as CAD, whereas the other computer-based operations listed above can be referred to as computer-aided process engineering (CAPE) or simply . computer-aided engineering (CAE).

It is now expected that all graduating chemical engineers will have had at least some basic experience of CAD in their course, and in many cases the students will have had extensive hands-on training with a variety of CAD programs and packages. This experience will vary depending upon the university, the funding and resources available, and the department philosophy to the teaching of design. A book of this type cannot devote more than a few pages to this subject, and even then presenting only an overview of the use of computers for design. This discussion must be limited to a general description of the types of CAD packages available, their usefulness and what the student can realistically expect to achieve. Specific details of particular packages are omitted because of the wide range of CAD materials now available and the variation between departments in terms of what is available for student use. However, some of the more-widely used packages are mentioned.

At this stage it is worth considering briefly what experience in CAD a university course should provide for a student, and what experience the student should expect to have when leaving university and beginning a career in industry. Hopefully these two aspects will not be too far different. Most employers of chemical engineering graduates now expect that the students will have had some hands-on experience of CAD in their course. How much experience they consider necessary probably depends upon the use of CAD packages within the company. However, just as students are expected to be adept at computer programming and to be able to apply their skills to whatever system is available, it would also be expected that CAD experience would be such that the student becomes capable and confident of being able to use new packages as they

become available (with some instruction and help). What is really important is that the student appreciates the need for and the use of CAD packages for solving chemical engineering design problems and that the operation and structure of such packages is understood, rather than the need to become fully conversant with the use of a large number of packages. Wherever the graduate engineer is employed, if CAD packages are used then instruction in their use and subsequent advice and assistance should be available. It is necessary for the student to become familiar with the use of such packages, and to be able to adapt to the use of new CAD facilities.

When using a calculator to perform numerical calculations it is necessary to know the sequence of operations and (just as important) to be able to assess the 'correctness' of the answer obtained. The same requirements apply to the use of CAD packages. The programming operations depend upon the particular package, but an assessment of the validity of the results is essential and depends upon the student understanding the fundamental principles of chemical engineering. Students need to be taught (or rather they need to develop the approach) to assess the results produced by the use of a CAD package and not blindly to assume that the results are correct or necessarily apply to the particular situation. The aim of using CAD packages is to reduce the time spent performing manual calculations, and also to enable the engineer to consider a wide range of alternative situations/values/ variables in order to optimize the problem and arrive at the 'best' solution. It is neither useful nor practical to suggest that the student (or engineer) should perform many detailed calculations manually to check that the computer results are valid. However, some checks should be performed for appropriate calculations and it should be possible for the students to perform the calculations if necessary, and to appreciate the design method or calculation sequence being performed. CAD packages will not replace the need for engineers who understand and can apply the basic principles of chemical engineering, however they can make the engineer's work more productive by reducing the time spent performing laborious or repetitive calculations.

The use of computer programs or packages to carry out material and energy balances is now commonplace, and probably one of the first encounters the student has with CAD. Unless the equipment or process is fairly simple, hand calculations become difficult and time consuming due to the iterative nature of the method. The design project may be the

student's first encounter with the calculation of material and energy balances for an entire process. Calculation by computer not only saves time but allows many changes to be made to the process variables in order to specify the optimum process conditions.

The computer program for the material balance contains several parts. First, a description of each item of equipment in terms of the input and output flows and the stream conditions. Quite complicated mathematical models may be required in order to relate the input and output conditions (i.e. performance) of complex units. It is necessary to specify the order in which the equipment models will be solved, simple equipment such as mixers are dealt with initially. This is followed by the actual solution of the equations. The ordering may result in each equation having only one unknown and iteration becomes unnecessary. It may be necessary to solve sets of linear equations, or if the equations are non-linear a suitable algorithm applying some form of numerical iteration is required.

The work involved in preparing a material balance computer program can be considerable, even for a simple process involving few units. The usefulness of these programs has lead to the development and availability of commercial 'Process Flowsheet' programs or packages such as LOTUS l-2-3. These programs contain sets of programmed, general-purpose unit-operation subroutines, and an executive program which can order the subroutines based upon the data input. The flowsheet-type program can also be connected to a physical property data bank so that appropriate data can be obtained automatically for a wide range of components without the need for user input. Further details of the use of flowsheeting programs for material and energy balances are given in Wells and Rose (1986; pp.285-294).

Flowsheeting programs can provide much more information than just the material and energy balances for the process. In performing these calculations, each item of equipment has effectively undergone a preliminary design evaluation. The results (or output) can usually include details of equipment heat loads, reflux ratios, utility requirements, and preliminary sizing. If the program is connected to a physical property data bank then extensive component property data can also be obtained for each process stream. Although a large number of flowsheet packages are commercially available, many have similar features and program structure. When evaluating packages for use (or purchase) it is desirable to compare the scope of operations included, the

range of equipment models, the ease of use (user-friendliness), the size of the data bank, and the quality of the support documentation.

All design studies require physical property data and the efficient use of CAD methods means that this data must be available in a convenient form for computer use. Data in the form of graphs or tables is of limited use, although some data banks do contain only measured data. A more useful form of data storage is to correlate the available data and then store the calculated parameters required for use in appropriate empirical equations (stating the limits over which the correlations apply). Some data banks contain only the calculated parameters whereas others also store the available raw data. The following are examples of some common physical property data banks:

ASPEN-PLUS	AspenTech, USA
CHEMCO	Eurecha, Switzerland
DATABANK	Imperial Chemical Industries, UK
FLOWTRAN	Monsanto, USA
PPDS	IChemE, UK

The most useful data banks for chemical engineering design problems are those where missing data can be obtained from appropriate predictive methods, e.g. vapour-liquid equilibria (VLE) data obtained from a group contribution method. Unavailable data may also be obtained by substituting data for a corresponding and similar compound. If the required data has no significant effect on the calculation then a rough estimate can be used. The user of CAD packages needs to know which predictive methods are being applied, and their accuracy and effect upon the calculated results. This is where an understanding of the structure and application of CAD packages is required, and a knowledge of fundamental engineering principles. More details concerning the use of physical property data banks are given in Wells and Rose (1986; pp.343-364).

To gain the maximum benefit from the use of a flowsheet program, the operator/designer must be adequately trained. A suitable program will have 20-30 standard units available, numerous equation-solving procedures, control facilities and probably optimization facilities. The unit-equipment subroutine must adequately represent the process equipment, recycle streams need to be specified, and suitable solution convergence is required. For the effective use of CAD packages, it

should be obvious that engineering software will not replace the engineer but a better trained engineer with wider abilities and judgement will be required.

The use of a good flowsheet package includes the following four stages: simulation, design, case-study, and optimization. The initial simulation stage requires definition of the input data and selection of appropriate equipment models. Simple units can replace more complex equipment at this stage, e.g. 'component splitter' for a distillation column. At this stage repeated substitution usually achieves convergence, and this can then be followed by use of a more suitable accelerated convergence algorithm. The simulation stage provides a simplified model which will converge, although the output and product specifications are probably incorrect. This initial stage is followed by the upgrading of the simulation model into a suitable design mode, usually by adding further recycles (as information loops). The design process is iterative and suitable convergence procedures are required. The required result is a model that produces a meaningful solution in a reasonable number of iterations. The case-study stage involves the use of actual plant data and rules-of-thumb for calculating initial design variables. The design model is used to provide a sensitivity approach to indicate which variables are significant and whether it is worthwhile implementing full-scale optimization procedures. The optimization stage requires a significant allocation of the designer's time and subsequent computer costs. The appropriate software must be available and the user should possess the necessary experience with these procedures. It will be necessary to optimize sections of the plant separately, and subsequently to make appropriate changes and repeat the calculations. A full economic analysis will also be required at this stage.

The application of computers to chemical engineering design, and the potential savings, have lead to the commercial development of a large number of CAD packages. Some of these packages deal with particular aspects of design work, such as detailed heat exchanger design (e.g. HEATEX by Humphreys and Glasgow, UK), flowsheet drawing and draughting packages (e.g. PRODESIGN II by American Small Business Computers, USA; a general drawing package which is easy to use and can be programmed to include a personal library of flowsheet drawing symbols and detailed engineering drawings), and data banks (e.g. Physical Property Data Service (PPDS) by the IChemE, UK). Many

other programs of the flowsheet-type are available for simulation and design of entire processes, including calculation of the material and energy balances. These include HYSIM (by Hyprotech Ltd, Canada) which can be used on a personal computer, and more extensive packages for use on mainframe computers, e.g. PROCESS (by Simulation Sciences, Inc., USA), ASPEN PLUS (by Aspen Technology, Inc., USA), and SPEED-UP (developed at Imperial College, London). The UNICORN flowsheet program was developed by Eurecha, Switzerland (European Committee for the Use of Computers in Chemical Engineering Education) specifically for teaching purposes, and it can be used with the data bank CHEMCO. Details of these two packages are given in Wells and Rose (1986; pp.649-692). It is important that undergraduate chemical engineers obtain experience in the use of CAD packages, but the actual packages used will depend upon the individual department. It is more important to appreciate the applications of CAD and the skills required by the computer-based designer, than to have superficial hands-on experience of many different systems.

A word of caution. The software suppliers accept no responsibility for any damage or loss caused by the use of CAD packages, even if the results obtained from the package are in error. The design engineer chooses to use the program and is responsible for checking and approving the results obtained.

7.4 Case Study — Mass and Energy Balances

Summary

There are several important observations from the series of mass and energy balances presented in this section. At 280 tonne per day capacity, the process is able to produce about 5800 kg of medium-pressure superheated steam (380°C and 4000 kPa) every hour. About 60% of this steam is available for export. This contributes a valuable source of operating revenue to the plant.

The plant is found to be highly efficient with a reaction yield of approximately 95%. Effluent levels from the process are small. There are no liquid waste streams and the tail gas should contain less than 1000 ppm of nitrogen oxides.

7.4.1   Overall Process Mass Balance

	AMMONIA		REACTOR FEED		REACTION GAS		REACTION GAS	
	WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%
AMMONIA	2006	100	2006	6.33	-	-	-	-
NITROGEN	-	-	22249	70.18	22331	70.43	22331	70.43
OXYGEN	-	-	6471	21.26	2116	6.67	2023	6.39
WATER	-	-	353	1.11	3539	11.16	3539	11.16
INERTS	-	-	356	1.12	356	1.12	356	1.12
NITROGEN MONOXIDE	-	-	-	-	3363	10.61	3195	10.09
NITROGEN DIOXIDE	-	-	-	-	-	-	258	0.81
NITROGEN TETROXIDE	-	-	-	-	-	-	-	-
NITRIC ACID	-	-	-	-	-	-	-	-
FLOWRATE (kg)	2006		31706		31706		31706	

	AIR FEED		REACTOR AIR		BLEACH AIR		REACTION GAS		REACTION GAS	
	WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%
AMMONIA	-	-	-	-	-	-	-	-	-	-
NITROGEN	22968	74.91	22249	74.91	4719	74.91	22331	70.43	22331	70.43
OXYGEN	8172	22.70	6742	22.70	1430	22.70	1771	5.59	1728	5.59
WATER	428	1.19	353	1.19	75	1.19	3539	11.16	3539	11.16
INERTS	432	1.20	356	1.20	76	1.20	356	1.12	356	1.12
NITROGEN MONOXIDE	-	-	-	-	-	-	2716	8.57	2634	8.30
NITROGEN DIOXIDE	-	-	-	-	-	-	963	3.04	1083	3.42
NITROGEN TETRAOXIDE	-	-	-	-	-	-	30	0.09	34	0.11
NITRIC ACID	-	-	-	-	-	-	-	-	-	-
FLOWRATE (kg)	36000		29700		6300		31706		31706	

REACTION GAS		BLEACH GAS		REACTION GAS		ADDED WATER		TAIL GAS	
WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%
-	-	-	-	-	-	-	-	-	-
22331	85.31	4719	69.81	27050	82.13	-	-	27050	94.33
399	1.52	1430	21.16	1381	4.19	-	-	850	2.96
-	-	75	1.11	75	0.23	2338	100	286	1.00
356	1.36	76	1.12	432	1.31	-	-	432	1.51
696	2.66	-	-	122	0.37	-	-	0	0.00
1240	4.74	-	-	2203	6.69	-	-	32	0.11
1154	4.41	460	6.80	1673	5.08	-	-	27	0.09
-	-	-	-	-	-	-	-	-	-
26176		6760		32936		2338		28678	

WATER

TAIL GAS

COOLER CONDENSER — OXIDATION UNIT — SECONDARY COOLER — ABSORBER — BLEACHING COLUMN
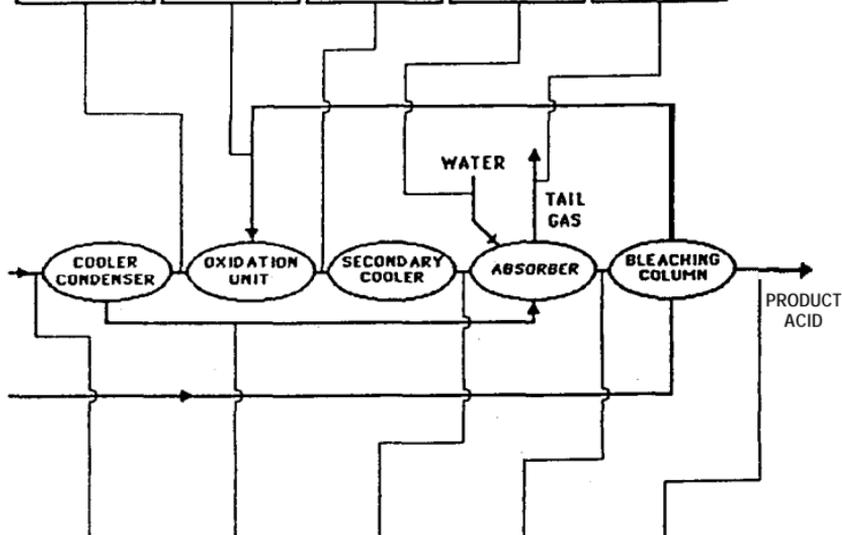
PRODUCT ACID

REACTION GAS		WEAK ACID		REACTION GAS		RED PRODUCT		PRODUCT ACID	
WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%	WEIGHT	%
-	-	-	-	-	-	-	-	-	-
-	70.43	-	-	27050	82.13	-	-	-	-
22331	5.49	3208	58.00	1428	4.33	-	-	-	-
1376	11.16	-	-	75	0.23	4667	38.48	4667	40.00
3539	1.12	-	-	432	1.31	-	-	-	-
356	8.30	-	-	102	0.31	-	-	-	-
1975	3.42	-	-	2164	6.57	-	-	-	-
1947	0.11	-	-	1685	5.12	460	3.79	-	-
181	-	2323	42.00	-	-	7000	57.73	7000	60.00
31706		5531		32936		12127		11667	

A refrigeration unit is required for the cooling of plant make-up water and also for cooling the absorption tower cooling water. The cooling duty required from the refrigeration unit is estimated to be 325 kW.

The air-feed compressor is a dual-stage unit with a calculated duty of 3 MW. Compressor shaft power is provided by tail-gas expansion (80% of required power) and a steam turbine.

Nitrogen oxides dissolved in 'red' product acid are stripped from the acid using a secondary air stream. This secondary air is recycled back into the oxidation unit, where the additional oxygen produces an extra 40% oxidation in the main reaction stream.

Detailed calculations concerning the mass and energy balances are presented in Appendix F.

7.4.2 Unit Mass and Energy Balances

7.4.2.7 Ammonia Vaporizer

SATURATED AMMONIA		
COMPONENT	WEIGHT	%
Ammonia	2006	100
-	-	-

Flowrate (kg) 2006
Temperature (°C) 35
Pressure (kPa) 1240

WARM WATER LOOP		
COMPONENT	WEIGHT	%
Water	21900	100
-	-	-

Flowrate (kg) 21900
Temperature (°C) 50
Pressure (kPa) 101

AMMONIA VAPOURIZER

LIQUID AMMONIA		
COMPENENT	WEIGHT	%
Ammonia	2006	100
-	-	-

Flowrate (kg) 2006
Temperature (°C) -15
Pressure (kPa) 1240

WARM WATER LOOP		
COMPONENT	WEIGHT	%
Water	21900	100
-	-	-

Flowrate (kg) 21900
temperature (°C) 80
Pressure (kPa) 101

7.4.2.2 Ammonia Superheater

SUPERHEATED STEAM		
COMPONENT	WEIGHT	%
Water	305	100
-	-	-
Flowrate (kg)	305	
Temperature (°C)	380	
Pressure (kPa)	4000	

SUPERHEATED AMMONIA		
COMPONENT	WEIGHT	%
AMMONIA	2006	100
Flowrate (kg)	2006	
Temperature (°C)	177	
Pressure (kPa)	1220	

AMMONIA SUPERHEATER

STEAM CONDENSATE		
COMPONENT	WEIGHT	%
Water	305	100
-	-	-
Flowrate (kg)	305	
Temperature (°C)	250	
Pressure (kPa)	4000	

SATURATED AMMONIA		
COMPONENT	WEIGHT	%
Ammonia	2006	100
-	-	-
Flowrate (kg)	2006	
Temperature (°C)	35	
Pressure (kPa)	1240	

7.4.2.3 Two-stage Air Compressor

COOLING WATER		
COMPONENT	WEIGHT	%
Water -	36000	100 -

Flowrate (kg)   36000
Temperature (°c)   20
Pressure (kPa)   101

COOLING WATER		
COMPONENT	WEIGHT	%
Water -	36000	100 -

Flowrate (kg)   36000
Temperature (°C)   40
Pressure (kPa)   101

180 °C   310 kPa
45 °C   300 kPa

Intercoolr

Mechanical Drive - steam Turbine

STAGE I

Mechanical Drive - Gas Turbine

STAGE 2

Compressor

SECONDARY BLEACH AIR		
COMPONENT	WEIGHT	%
Air	6300	100

Flowate (kg)   6300
Temperature (°C)   232
Pressure (kPa)   1090

PROCESS AIR		
COMPONENT	WEIGHT	%
Air -	36000	100 -

Flowrate (kg)   36000
Temperature (°C)   35
Pressure (kPa)   99

REACTOR FEED		
COMPONENT	WEIGHT	%
Air -	29700	100 -

Flowrate (kg)   29700
Temperature (°C)   232
Pressure (kPa)   1090

7.4.2.4 *Reactor Feed Mixer*

SUPERHEATED AMMONIA		
COMPONENT	WEIGHT	%
Ammonla	2006	100

Flowrate (kg) 2006
Temperature (°C) 177

Pressure (kg'a) **I240**

REACTOR AIR FEED		
COMPONENT	WEIGHT	%
Air	**29700**	**100**

Flowrate (kg) 29700
Temperature (°C) 262
Pressure (kPa) 1090

MIXER

REACTOR FEED		
COMPONENT	WEIGHT	%
Air	29700	89.7
Ammonia	2006	10.3
–	–	–

Flowrate (kg) 31706
Temperature (°C) 250
Pressure (kPa) 1060

7.4.2.5 Reactor

REACTOR FEED

COMPONENT	WEIGHT	%
Ammonia	.2006	10.3
Air	29700	89.7
Flowrate (kg)	31706	
Temperature (°C)	250	
Pressure (kPa)	1060	

PREHEATED AIR - MIXER

COMPONENT	WEIGHT	%
Air	29700	100
-	-	-
Flowrate (kg)	29700	
Temperature (°C)	262	
Pressure (kPa)	1090	

REACTOR

TAIL GAS EX-PREHEATER

COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
NO2	32	0.1
N2O4	27	0.1
-	-	-
Flowrate (kg)	28678	
Temperature (°C)	235	
Pressure (kPa)	960	

COLD AIR TO MIXER

COMPONENT	WEIGHT	%
Air	29700	100
-	-	-
Flowrate (kg)	29700	
Temperature (°C)	262	
Pressure (kPa)	1090	

TAIL GAS TO EXPANDER

COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
NO2	32	0.1
N2O4	27	0.1
-	-	-
Flowrate (kg)	28678	
Temperature (°C)	620	
Pressure (kPa)	950	

REACTION GASES

COMPONENT	WEIGHT	%
Nitrogen	22331	70.4
Oxygen	2116	6.7
Water	3539	11.2
Inerts	356	1.1
NO	3363	10.6
-	-	-
Flowrate (kg)	31706	
Temperature (°C)	645	
Pressure (kPa)	1035	

7.4.2.6 Steam Superheater

REACTION---GASES		
COMPONENT	**WEIGHT**	**%**
Nitrogen	2233	70.4
Oxygen	2116	6.7
Water	3539	11.2
Inerts	356	1.1
NO	3363	10.6
Flowrate (kg)	**31706**	
Temperature (°C)	**645**	
Pressure (kPa)	**1035**	

SUPERHEATED STEAM		
COMPONENT	**WEIGHT**	**X**
Water	5775	100
Flowrate (kg)	**5775**	
Temperature (°C)	**380**	
Pressure (kPa)	**4000**	

STEAM SUPERHEATER

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	22331	70.4
Oxygen	2023	6.4
Water	3539	11.2
Inerts	356	1.1
NO	3195	10.1
NO2	258	0.8
—	—	—
Flowrate (kg)	31706	
Temperature (°C)	595	
Pressure (kPa)	1025	

SATURATED STEAM		
COMPONENT	WEIGHT	%
Water	5775	100
		—
Flowrate (kg)	5775	
Temperature (°C)	250	
Pressure (kPa)	4000	

7.4.2.7 Waste- Heat Boiler

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	22331	70.4
Oxygen	2023	6.4
Water	3539	11.2
Inerts	356	1.1
NO	3195	10.1
NO2	258	0.8
		–
Flowrate (kg)	31706	
Temperature (°C)	595	
Pressure (kPa)	1025	

SATURATED STEAM		
COMPONENT	WEIGHT	%
Water	6000	100
–	–	–
Flowrate (kg)	6000	
Temperature (°C)	250	
4 Pressure0 (kPa)	0	0

TO PLATINUM FILTER

WASTE HEAT BOILER

HOT WATER		
COMPONENT	WEIGHT	%
Water	6000	100
–	–	–
Flowrate (kg)	6000	
Temperature (°C)	117	
Pressure (kPa)	4000	

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	2233 1	70.4
Oxygen	1741	5.5
Water	3539	11.2
Inerts	356	1.1
NO	2716	8.6
NO2	963	3.0
N204	30	0.1
Flowrate (kg)	31706	
Temperature (°C)	280	
Pressure (kPa)	1020	

7.4.2.8 Platinum Filter

REACTION GASES		
COfIPONENT	WEIGHT	%
Nitrogen	2233	70.4
oxygen	1771	5.6
Water	3539	11.2
Inerts	356	1.1
NO	2716	6.6
NO2	963	3.0
N204	30	0.1
Flowrate (kg) 31706		
Temperature (°C) *280*		
Pressure (kPa) 1020		

TO
TAIL GAS PREHEATER

FROM
WASTE HEAT BOILER

PLATINUM FILTER

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	2233 I	70.4
Oxygen	1728	5.5
Water	3539	11.2
Inerts	356	1.1
NO	2634	8.3
NO2	1083	3.4
N204	34	0.1
Flowrate (kg) 31706		
Temperature (°C) 315		
Pressure (kPa) I020		

7.4.2.9   Tail- Gas Preheater

REACTION GASES		
COMPONENT	WEIGHT	%
Nf trogen	2233	70.4
Oxygen	1726	5.5
Water	3539	11.2
fnerts	356	1.1
N O	2634	8.3
N O 2	1083	3.4
N204	34	0.1
Flowrate (kg) 31706		
Temperature (°C) 315		
Pressure (kPa) 1020		

TAIL GAS EX- 1° HEATER		
COMPONENT	WEIGHT	%
Nftrogen	27050	94.3
Oxygen	850	3.0
Water	266	1.0
fnerts	4 3 2	1.5
N O 2	32	0.1
N204	27	0.
Flowrate (kg) 3 1706		
temperature (°C) 4 7		
Pressure (kPa) 970		

TO COOLER-CONDENSER

from
PLATINUM   FILTER
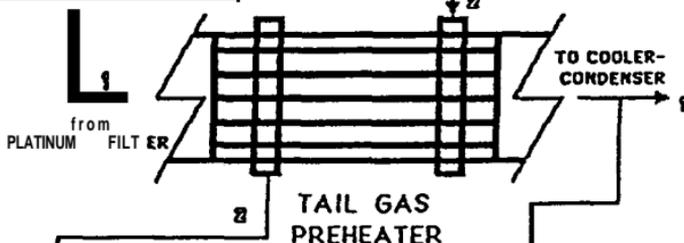
TAIL GAS
PREHEATER

TAIL GAS TO 3' HEATING		
COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
N O 2	32	0.1
N204	27	0.1
Flowrate (kg) 31706		
Temperature (°c) 235		
Pressure (kPa) 960		

REACTION GASES		
COtlPONENT	WEIGHT	%
Nitrogen	22331	70.4
Oxygen	1376	4.3
Water	3539	11.2
Inerts	356	1.1
N O	1975	6.2
N O 2	1947	6.1
N204	181	0.7
Flowrate (kg) 31706		
Temperature (°C) 185		
Pressure (kPa) 1018		

7.4.2.10 Oxidation Unit

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	27050	82.1
Oxygen	1381	4.2
water	7s	0.2
Inerts	432	1.3
NO	122	O.4
NO2	2203	6.7
N2O4	1673	5.1

Flowrate (kg) 32936
Temperature (°C) 140
Pressure (kPa) 1000

OXIDATION UNIT

From
Cooler/Condenser

REACTION GASES		
COMPONENT	WEIGHT	%
Nitrogen	22331	85.3
Oxygen	399	1.5
Water	0	0.0
Inerts	356	1.4
NO	696	2.7
NO2	1240	4.7
N2O4	1154	4.4

Flowrate (kg) 26176
Temperature (°C) 60
Pressure (kPa) 1015

BLEACHING AIR		
COMPONENT	WEIGHT	%
Air	6300	93.2
N2O4	460	6.8
-	-	-

Flowrate (kg) 6760
Temperature (°C) SO
Pressure (kPa) 1020

7.4.2.7 7 Cooler/Condenser

COOLING WATER

COMPONENT	W E I G H T	%
Water	77000	100
-	-	-

Flowrate (kg) 77000
Temperalure (°c) 40
Pressure (kPa) 101

REACTION GASES

COMPONENT	WEIGHT	%
Nitrogen	22331	85.3
Oxygen	399	1.5
Water	0	0.0
Inerts	356	1.4
NO	696	2.7
NO2	1240	4.7
N2O4	1154	4.4

Flowrate (kg) 26176
Temperature (°C) 60
Pressure (kPa) 1015

FROM
TAIL GAS PREHEATER

TO
OXIDIZER

TO
ABSORBER

COOLER

REACTION GASES

COMPONENT	WEIGHT	%
Nitrogen	22331	70.4
Oxygen	1376	4.3
Water	3539	1.2
Inerts	356	1.1
NO	1975	6.2
NO2	1947	6.1
N2O4	181	0.6

Flowrate (kg) 31706
Temperature (°C) 185
Pressure (kPa) 1018

COOLING WATER

COMPONENT	WEIGHT	%
Water	77000	100
-	-	-

Flowrate (kg) 77000
Temperature (°C) 20
Pressure (kPa) 101

WEAK ACID

COMPONENT	WEIGHT	%
Water	3208	58
Nitric Acid	2323	42
-	-	-

Flowrate (kg) 5531
Temperature (°C) 66
Pressure (kPa) 1015

7.4.2.72 Secondary Cooler

REACTION GASES

COMPONENT	WEIGHT	%
Nitrogen	27050	82.1
Oxygen	1381	4.2
Water	75	0.2
Inerts	432	1.3
NO	122	0.4
NO2	2203	6.7
N2O4	1673	5.1

Flowrate (kg)   32936
Temperature (°C)   140
Pressure (kPa)   1000

WARN WATER LOOP

COMPONENT	WEIGHT	%
Water	21900	100
-	-	-

Flowrate (kg)   21900
Temperature (°C)   50
Pressure (kPa)   101

SECONDARY COOLER

WARM WATER LOOP

COMPONENT	WEIGHT	%
Water	21900	100
-	-	-

Flowrate (kg)   21900
Temperature (°C)   80
Pressure (kPa)   101

TO ABSORBER

REACTION GASES

COMPONENT	WEIGHT	%
Nitrogen	27050	82.1
Oxygen	1428	4.3
Water	75	0.2
Inerts	432	1.3
NO	102	0.3
NO2	2164	6.6
N2O4	1685	5.1

Flowrate (kg)   32936
Temperature (°C)   65
Pressure (kPa)   995

7.4.2.73 Absorber

TAIL GAS

COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
NO2	32	0.1
N2O4	27	0.1
-	-	-

Flowrate (kg) 28678
Temperature (°C) 10
Pressure (kPa) 980

MAKE-UP WATER

COMPONENT	WEIGHT	%
Water	2338	100
-	-	-

Flowrate (kg) 2338
Temperature (°C) 7
Pressure (kPa) 995

CHILLED WATER LOOP

COMPONENT	WEIGHT	%
Water	19100	100
-	-	-

Flowrate (kg) 19100
Temperature (°C) 7
Pressure (kPa) 101

WEAK ACID

COMPONENT	WEIGHT	%
Water	3208	58
Nitric Acid	2323	42
-	-	-

Flowrate (kg) 5531
Temperature (°C) 65
Pressure (kPa) 1015

COOLING WATER

COMPONENT	WEIGHT	%
Water	19100	100
-	-	-

Flowrate (kg) 19100
Temperature (°C) 20
Pressure (kPa) 101

CHILLED WATER LOOP

COMPONENT	WEIGHT	%
Water	19100	100
-	-	-

Flowrate (kg) 19100
Temperature (°C) 20
Pressure (kPa) 101

COOLING WATER

COMPONENT	WEIGHT	%
Water	19100	100
-	-	-

Flowrate (kg) 19100
Temperature (°C) 40
Pressure (kPa) 101

REACTION GAS

COMPONENT	WEIGHT	%
Nitrogen	27050	82.1
Oxygen	1428	4.3
Water	75	0.2
Inerts	432	1.3
NO	102	0.3
NO2	2164	6.6
N2O4	1685	5.1

Flowrate (kg) 32936
Temperature (°C) 65
Pressure (kPa) 995

RED PRODUCT ACID

COMPONENT	WEIGHT	%
Water	4667	38.5
Nitric Acid	7000	57.7
N2O4	460	3.8
-	-	-

Flowrate (kg) 12127
Temperature (°C) 46
Pressure (kPa) 990

7.4.2.14 Bleaching Column

BLEACHING AIR

COMPONENT	WEIGHT	%
Air	6300	93.2
N2O4	460	6.8
-	-	-

Flowrate (kg)   6760
Temperature (°C) 50
Pressure (kPa)   1020

**PRODUCT
BLEACHING COLUMN**

RED PRODUCT ACID

COMPONENT	WEIGHT	%
Water	4667	38.5
Nitric Acid	7000	57.7
N2O4	460	3.8
-	-	-

Flowrate (kg)   12127
Temperature (°C) 45
Pressure (kPa)   990

SECONDARY BLEACH AIR

COMPONENT	WEIGHT	X
Air	6300	100

Flowrate (kg)   6300
Temperature (°C) 80
Pressure (kPa)   1070

PRODUCT ACID

COMPONENT	WEIGHT	X
Water	4667	40
Nitric Acid	7000	60

Flowrate (kg)   11667
Temperature (°C) 55
Pressure (kPa)   990

7.4.2.15 *Vapour/Liquid Separator*

SATURATED STEAM

COMPONENT	WEIGHT	%
Water	577s	100

Flowrate (kg)   5775
Temperature (°C)   250
Pressure (kPa)   4000

DEIONISED WATER

COMPONENT	WEIGHT	%
Deionised Water	–	–
	5470	100

Flowrate (kg)   5470
Temperature (°C)   96
Pressure (kPa)   4000

VAPOUR/LIQUID SEPARATOR

SATURATED STEAM

COMPONENT	WEIGHT	%
Water	6000	100
–	–	–

Flowrate (kg)   6000
Temperature (°C)   250
Pressure (kPa)   4000

STEAM CONDENSATE

COMPONENT	WEIGHT	%
Water	305	100

Flowrate (kg)   305
Temperature (°C)   250
Pressure (kPa)   4000

HOT WATER

COMPONENT	WEIGHT	%
Water	6000	100
–	–	–

Flowrate (kg)   6000
Temperature (°C)   117
Pressure (kPa)   4000

7.4.2.16 Tail- Gas Warmer

DEIONISED WATER

COMPONENT	WEIGHT	%
Deionised Water	5470	100

Flowrate (kg)   5470
Temperature (°C) 96
Pressure (kPa) 4000

SECONDARY BLEACH AIR

COMPONENT	WEIGHT	%
Air	6300	100

Flowrate (kg)   6300
Temperature (°C)   80
Pressure (kPa)   1070

TAIL GAS

COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
102	32	0.1
N204	27	0.1

Flowrate (kg)   28678
Temperature (°C) 10
Pressure (kPa) 980

TAIL GAS WARMER

TAIL GAS

COMPONENT	WEIGHT	%
Nitrogen	27050	94.3
Oxygen	850	3.0
Water	286	1.0
Inerts	432	1.5
NO2	32	0.1
N204	27	0.1

Flowrate (kg)   28678
Temperature (°C)
Pressure (kPa) 98

SECONDARY BLEACH AIR

COMPONENT	WEIGHT	%
Air	6300	100

Flowrate (kg)   6300
Temperature (°C)   232
Pressure (kPa)   1090

DEIONISED WATER

COMPONENT	WEIGHT	%
Deionised Water	5470	100

Flowrate (kg)   5470
Temperature (°C) 100
Pressure (kPa) 4000

7.4.2.17 Refrigeration Unit

COOLING WATER

COMPONENT	WEIGHT	%
Water	42080	100

Flowrate (kg) 42080
Temperature (°C) 20
Pressure (kPa) 101

WARM CONDENSATE

COMPONENT	WEIGHT	%
Water	2338	100

Flowrate (kg) 2338
Temperature (°C) 40
Pressure (kPa) 101

REFRIGERATION
UNIT

CHILLED WATER LOOP

COMPONENT	WEIGHT	%
Water	19100	100

Flowrate (kg) 19100
Temperature (°C) 20
Pressure (kPa) 101

Coolant

REFRIGERATION UNIT
DUTY-1 190 150 kJ

COOLING WATER

COMPONENT	WEIGHT	%
Water	42080	100

Flowrate (kg) 42080
Temperature (°C) 22
Pressure (kPa) 101

CHILLED WATER LOOP

COMPONENT	WEIGHT	%
Water	19100	100

Flowrate (kg) 19100
Temperature (°C) 7
Pressure (kPa) 101

CHILLED MAKEUP WATER

COMPONENT	WEIGHT	%
Water	2338	100

Flowrate (kg) 2338
Temperature (°C) 7
Pressure (kPa) 990

Comments

The reader should by now have made two important observations about plant and process design studies. First, decisions are usually based upon a compromise between conflicting factors, the final design is not necessarily the 'best' in all aspects but it must be technically and economically acceptable for the task. Second, a design study does not follow the usual format for solution of a traditional undergraduate tutorial problem. There is a preferred sequence of activities to be followed, but this does not lead to a set of completed individual stages and a final single correct solution. The design project requires a continual re-evaluation of previous decisions, changes are made as new developments occur or as' additional information becomes available. Several aspects of the design need to be considered at different stages of the work. A preliminary economic feasibility study is performed (Chapter 6), and a final economic evaluation after the detailed design is completed. The equipment design is considered in the feasibility study, the preparation of the equipment list, the 'mass and heat balance calculations, and during the preliminary and final design stages. The same work is not repeated each time, but new information is assessed, previous data and decisions are re-assessed, and the design is continually refined and developed. This is the nature of the design project-it is the ultimate iterative, trial-and-error type problem.

The student undertaking a design for the first time often finds this unnerving. There is often a reluctance to carry out the early stages of the project because the detail required for final decision-making is incomplete. However, it should be apparent that to wait until all information is available, and then make the decisions and carry out the calculations in 'one-hit' is not the best way to proceed.

Action: Adopt a new way of **thinking and approaching engineering problems — be prepared to re-assess decisions at every stage** of **the design, and incorporate necessary changes. Do not assume every problem has a particular method** of **solution and a single correct   answer.**

References

Ray, M.S., *Elements of Engineering Design : An Integrated Approach,* Prentice-Hall International, UK (1985).
A general introductory textbook on engineering design, including discussion of creativity, innovation, problem solving and decision making, and other aspects of engineering design such as economics, ergonomics, legal factors, etc.
Sears, J.T.; Woods, D.R.; and Noble, R.D. (Eds), *Problem Solving, AIChemE Symposium Series,* Volume 79, Number 228, New York (1983).

PART II

Detailed Equipment Design

Note   For Part II, references are included at the end of each chapter. General references are given in Sections 2.4 and 2.5, and Appendix L.

CHAPTER 8
The Detailed Design Stage

8.1 Detailed Equipment Design

AFTER ALL the preliminary work has been completed, the detailed design work can begin. The equipment can be designed in its final (?) form and full specification sheets prepared for each item. The process flowsheet and equipment list can be checked and amended. The cost estimates can also be revised to account for any significant changes from the preliminary design specifications. Space precludes the inclusion of design details for particular items of equipment in this book. A general discussion of equipment design is included and some appropriate references.

Many different process operations are performed in a chemical plant, e.g. distillation, heat transfer, absorption, etc., and for each of these operations a wide range of equipment has been developed to suit particular applications and situations. These include different types of heat exchangers, e.g. plate, shell and tube, carbon block, etc.; packed and plate distillation columns; CSTR or tubular reactors, etc., etc. Many books have been published describing particular process operations, e.g. distillation, heat transfer, etc., a few of the more prominent volumes are listed in the references at the end of this chapter. By the time the student undertakes a design project, these books should be very familiar from use in other parts of the course. Some books provide an introduction to the subject, aimed mainly at undergraduate courses and emphasising the basic theory. Other books concentrate upon the application of basic principles to the design of equipment, and include details of relevant design methods. This latter type of book is needed for design work, some examples are also listed at the end of this chapter.

The reader requiring an overview or a review of the different types of equipment available and their relevant design features, should refer to

the chapters on equipment design in Baasel (1976); Peters and Timmerhaus (1980); and Ulrich (1984). These books, provide extensive references to the literature for different types of equipment and detailing new developments in the design of equipment. More detailed design information can be found in Perry's Handbook (1984). It is not sufficient to base a design on a method presented in a book published 10 years ago (or even 5 years ago), the design engineer must be familiar with new methods and improvements as they occur. Papers published in *Chemical Engineering* (including Feature Articles describing the current state-of-the-art in particular fields), *Hydrocarbon Processing* and *Chemical Engineering Progress* are good sources of recent information. Other journals and reference sources are listed in Appendix L.

The design of an item of equipment must conform to the following criteria:

(a) It must perform its intended function, i.e. fulfill the design brief.

(b) It must operate safely, minimising the risks of explosion, fire and danger to operating personnel.

(c) It must be an economic design.

(d) It should operate with minimum maintenance and repair, and with minimum operating costs and utilities requirements.

(e) It is usually required to operate 'satisfactorily' under conditions of both increased and reduced capacity.

(f) It must operate for the expected life of the plant.

(g) It must be capable of construction and operation.

(h) It must be 'acceptable' (?).

(i) And probably several other requirements.

When the equipment is designed, a full design report must be prepared giving details of all relevant design calculations and the data used, and assumptions/restrictions inherent in the design method. A full design specification sheet and a detailed (scaled) engineering drawing are prepared. Sufficient details must be available so that the item can be costed accurately — and constructed (if necessary).

8.1.1   Equipment Design — HELP!!!

Not all the students in a chemical engineering course have mastered all the basic principles that have been taught, many students feel uncertain about their ability to transform basic theory into an engineering design

adapting to open-ended (and often ill-defined) design problems after a 'diet' of predominantly precise textbook-type exercises. It is necessary to realise that the first stage of a design problem is to provide some order to the available (and sometimes contradictory) information that constitutes the basis of the equipment design. These notes are intended to provide some immediate help and guidance to those students who feel apprehensive about tackling an equipment design problem. Reference to application. Sometimes the better academic students find difficulty these notes should also be useful as the design proceeds, to ascertain that important aspects have not been overlooked. The following brief notes do not represent a detailed analysis of the design process but rather a few general pointers to indicate the general direction.

How to begin to design an item of equipment?

1. Whatever you have to design there is always some information available. In the absence of 'hard' data, agree with the supervisor the necessary assumptions/estimates that will enable the design to proceed. Write down **all** the available information which might be applicable to the design of this item of equipment. Itemise the information that is required, but not available, for the design.

   DO NOT flounder at this stage, decide to begin (and begin to decide!) and make the list of available (and required) information.

   DO NOT expect the data to be absolute, this is not a textbook problem, some data will be contradictory and some decisions will be based upon estimates. Some students become paralysed, unwilling to proceed, because the problem is not clearly defined and all necessary information is not available. Make a start, check the validity of any estimates as the design proceeds.

2. **Formulate a statement of the design criteria for the unit.** What is the general design criteria? What is the design (or the unit) intended to achieve? For example, a reactor is usually required to achieve a particular conversion of reactants to products, a distillation column must provide a particular separation, an absorption column provides removal of a specific level of impurity, etc.

3. **Identify the design parameters** that will be required to complete the design, e.g. temperatures, pressures, flowrates, compositions, etc. Complete a **preliminary design specification** by listing the known information, and providing estimates (clearly stated as such) if possible for any unknown requirements.

i.e. different types of equipment that could be used, alternative methods of operation, varying the particular requirements, etc.

5. *Decision time* — select the particular type of equipment to be used. Justify this choice — list the advantages *and* disadvantages compared with alternative equipment or process schemes.

   Never consider a design decision to be the only possible correct choice, or to be irreversible. As the detailed design proceeds, more appropriate choices may become apparent. It may be necessary to make major changes, this is unfortunate in relation to the design time already spent but not as serious as the consequences of producing an unacceptable design!

6. *Revise the design criteria* and preliminary design specification, if necessary.

7. *How is this type of equipment designed? You* need a *design method* to be implemented. Refer to design books, handbooks, journal articles, etc. Remember to do some background reading, determine the basic principles of operation and the theory related to the design of the equipment. Never proceed with a design method unless you are completely satisfied with its theoretical basis, and with your knowledge of the associated chemical engineering principles. If you are not satisfied with your understanding of the justification of the proposed design method — obtain reliable advice.

8. Ensure that the proposed design method is actually applicable to the type of equipment to be designed. This appears obvious, but many times students have applied design methods unsuitable for a particular application.

9. Study the design method in detail. Are all the necessary data, physical and empirical constants, etc., available?

10. List all the assumptions/limitations inherent in the design method. State all the estimates that must be made to apply the method — both in the method itself and in the design specification parameters.

11. Apply the design method to your situation.

12. *Assess the results,* i.e. the design specification, obtained using this method. Does the design appear to be sensible? Is the design acceptable? Do not assume that the results obtained by applying an established design method are necessarily acceptable or sensible.

    The design is not finished at this stage. Establish the acceptability of the results by comparison with other sources, e.g. published information in the literature, comparison with the performance or

4. Consider general ways in which the design criteria can be achieved, specification of other types of equipment, manufacturer's data, economic considerations, materials and fabrication requirements for the design, etc.

   If the design is not acceptable in all respects, then either corrections or modifications must be made to the design method employed, or a completely new approach must be adopted (return to point 2 above).

13. A complete design includes consideration of all aspects of the equipment specification such that the item could be built and operated. These aspects include the mechanical design, materials selection, fabrication requirements, operational details, safety features, versatility (e.g. turn down), economic considerations, etc. The scope (i.e. depth) of the design should be clearly specified by the supervisor.

Summary

The most relevant advice when faced with the task of designing an item of equipment is to get started! Survey the appropriate literature for useful background information and details of available design methods. Make necessary assumptions/estimates in order to proceed with the design. It is extremely unlikely that a design will be completely original with no precedent. Do not expect the design method to be completely rigorous, many designs are based upon empirical correlations and experience (i.e. 'rules of thumb').

8.2   Additional   Design   Considerations

The detailed design stage described so far, describes fairly accurately the way in which many student projects are carried out. However, several important aspects have been ignored which should be considered as the detailed equipment designs are being performed. In many student projects, these other factors are considered after the equipment design is completed and are usually adapted to fit the final (?) design specification. After reading these notes, it should be obvious that this procedure is unacceptable and the following additional factors *must* be considered *concomitantly*   with   the   detailed   equipment   design.

8.2.1 Energy Conservation

Chemical plants have always been designed to operate efficiently and economically due mainly to product competition. However before 1970, the objective of building a low first-cost plant was generally considered more important than low operating cost. This concept changed due to the oil crisis and the effects of the U.S. Environmental Protection Agency in emphasising non- or low-polluting fuels to replace heavy fuel oils and coal. Since the oil (energy?) crisis of the 1970s particular attention has been paid to such topics as energy conservation schemes, process (heat) integration, heat exchanger networks, cogeneration, etc. This attention is evidenced by the large number of books and journal articles published on these topics in recent years, a few useful references are included at the end of this chapter.

The design engineer must consider appropriate energy conservation/integration schemes that are designed to:

(a) utilise as much of the energy available within the plant;

(b) minimise the energy requirements for the plant.

The energy balances performed for the plant items (or plant sections) provide the initial key to identifying areas of high energy availability or demand. An attempt can then be made to utilise excess energy in areas where energy must be provided. However, this is not always possible because:

(a) a high energy load (from the energy balance) may constitute a large volume of liquid at a relatively low temperature, exchanging this energy may require large and expensive equipment;

(b) the energy source may be distant from the sink, piping and insulation costs may make this uneconomic; perhaps a re-arrangement of the plant layout (Section 5.2) is required;

(c) the energy source may be corrosive.

Any energy conservation scheme must also consider the cost of removing or transferring the excess energy, i.e. heat exchanger capital cost, piping, valves, pumps, insulation, operating costs of pumps and maintenance. Energy conservation is only worthwhile if the reduction in energy costs exceeds the cost of implementation. A scheme may be devised for a plant and then held over until energy prices make the proposal attractive. This type of forward planning requires that the plant layout adopted can be easily modified.

Energy conservation can be achieved at three 'levels', these are:

(a) correct plant operation and maintenance;

(b) major changes to existing plants and/or processes;

(c) new plants and/or new processes.

The time required to implement energy conservation measures, the capital cost required, and the potential savings, all increase from level (a) to (c) above. The cost of downtime for level (b) can be significant, and level (c) offers the greatest long-term potential for energy conservation. This latter objective can be achieved either by designing new, energy-efficient plants for established process routes, or adopting new and less energy-intensive process routes. The areas immediately obvious for consideration of energy utilisation are the feed preheating section of the plant (probably utilising high-temperature reaction products), and the distillation section by attempting to utilise condenser and reboiler duties on various columns. The following examples illustrate some applications of basic engineering principles to the design of equipment for improved energy efficiency.

(i) *Plant operation*

Energy savings can be achieved by good engineering practice and the application of established principles. These measures may be termed 'good housekeeping' and include correct plant operation and regular maintenance. The overall energy savings are usually small ($< 10\%$) and may not be easy to achieve, significant time may be required for regular maintenance and checking. However such measures help to establish the commitment of a company to a policy of energy conservation.

(ii) *Heat recovery*

Heat recovery (or reuse) is an important and fundamental method of energy conservation. The main limitations of this method are:

(a) inadequate scope for using recovered waste heat because it is too low grade for existing heat requirements, and/or because the quantity of waste heat available exceeds existing requirements for low-grade heat;

(b) inadequate heat-transfer equipment.

Developments and improvements are continuing in the design and operation of different types of heat exchangers including the use of

extended (finned) heat-transfer surfaces, optimizing heat-exchanger networks, heat recovery from waste fuels, heat exchanger fouling, and the use of heat pumps.

(iii) *Combined heat and power systems*

Significant energy conservation is achieved by the well-established method of combined heat and power generation (cogeneration). The heat is usually in the form of intermediate or low-pressure steam, and the power as direct mechanical drives or as electricity generated with turbo-alternators. The choice of systems is usually between back-pressure steam turbines, or gas turbines with waste-heat boilers for the process steam. The amount of power generated is usually determined by the demand for heat.

It is not usually possible to balance exactly the heat and power loads in a system. The best method of achieving this aim is to generate excess electricity for subsequent sale, other balancing methods tend to be less efficient. It is therefore important to forecast the heat:power ratio accurately at the design stage to avoid large imbalances and reduced system efficiency.

(iv) *Power-recovery systems*

A power-recovery (expansion) turbine can recover heat from an exhaust gas, and then use this heat to provide part of the energy required to drive the shaft of a motor-driven process-air compressor. Other examples are the use of a steam-turbine drive and a two-stage expansion turbine with reheating between the stages.

A hydraulic turbine can be incorporated on the same shaft as a steam turbine (or motor drive). This arrangement can be used to provide about 50% of the energy needed to recompress the spent liquor in a high-pressure absorption/low-pressure stripping system.

Power generation using steam or gas turbines is now well established, however power recovery by the pressure reduction of process fluids is more difficult and less common. In general, the equipment is not considered to be particularly reliable. Rankine cycle heat engines have been developed/adapted to use relatively low-grade waste-heat sources (particularly from organic fluids) to generate power in the form of electricity or direct drives. They tend to be used when the heat source

would otherwise be completely wasted, the low efficiencies (10–25%) do not then represent a significant disadvantage.

(v) Furnace efficiency

Incorporating an air heater (to utilise low-level heat) on a steam boiler can be more economic than using a hot-oil system which is designed for high-level heat only.

(vi) *Air coolers versus water coolers*

Air coolers have a higher installed cost but lower operating costs than water coolers.

(vii) *Low-pressure steam*

Energy savings can be achieved by the efficient use of low-pressure steam, e.g. use of 60–80 kPa gauge (10–12 psig) steam in absorption–refrigeration units (steam condensate is returned to the boiler).

(viii)   *Distillation*

All separation processes use energy but distillation is the most significant energy-consuming process, and has therefore received considerable attention. Other less versatile processes are also being developed as alternative separation methods. The other main energy-consuming processes are drying and evaporation. The three main ways of reducing the energy requirements for distillation are discussed below.

Improved control and operation of existing columns includes using a lower reflux ratio (or better control of reflux ratio), reducing (if possible) the overhead and bottom product purity specifications, lower operating pressures, and correct feed plate location. Energy can be saved by designing tray-distillation columns to include more trays and hence operate at a lower reflux ratio. A reflux ratio of between 1.10 and 1.20 times the minimum value is used for many installations, except for low-temperature (cryogenic) columns because of their high utilities costs (use a reflux ratio of 1.05-l. 10 times the minimum value) and in systems using low-level heat or recovered heat (use 1.20-l SO times minimum).

The second method involves recovery of the heat content of overhead

vapours to be used for generating low-pressure steam (if there is a suitable use for it). The overheads can be used in the reboiler of another column. The overhead vapours can also be used in the same column in an open-cycle heat-pump system known as vapour-recompression distillation. The overhead vapours are compressed and then condensed in the reboiler. Heat of compression raises the temperature of the vapour and provides the necessary AT for the reboiler. Condensed vapour is collected in a pressure-controlled reflux drum (no reflux pump), and is returned to the top of the column. This technique is useful if the relative volatilities of the components are 'enhanced' at reduced pressures, and therefore fewer trays and/or a lower reflux ratio are required.

Energy savings can often be achieved in multicomponent distillation systems by appropriate sequencing of the columns.

(ix) *Drying operations*

Dryers may be classified as direct (convection) dryers, radiant-heat dryers, and indirect (conduction or contact) dryers. Direct-contact dryers are inherently less efficient than the other types due to the difficulty of economically recovering heat from the exit gases. Energy conservation could be increased by improving the thermal efficiency (typically < 30%). Thermal drying is an energy-intensive process and consideration should be given to the use of alternative drying methods, e.g. mechanical separations such as filters and centrifuges for water removal. General methods of reducing the energy requirements of drying operations include:

(a) improved operation and insulation of existing dryers;
(b) reduced moisture content in the feed;
(c) optimization of the exhaust air rate, humidity and temperature;
(d) heat recovery from the exhaust gas.

Energy conservation in the design of complete processes may be achieved in four ways:

(A) major modifications to existing plant;
(B) new plant using an existing process route;
(C) new process routes and/or alternative raw materials;
(D) new processes for new (substitute) products that are less energy intensive.

Items (A) to (C) represent short-term and medium-term energy conservation measures. Item (D) requiring use of new products or processes is more appropriate for long-term energy efficiency planning, due to the long lead times for the application of new technology in the chemical industry. Although energy conservation is an obvious objective of all equipment manufacturers and plant designers, more attention is necessary in relation to education, training and the application of new and existing technology to ensure significant medium-term and long-term savings.

Energy conservation must be considered at various stages of the project, e.g. feasibility study, process selection, plant layout, energy balances (identifying major areas), and in conjunction with the detailed equipment design. If energy utilisation is only an afterthought, either unnecessary and costly modifications may be required to the design work, or the plant may not be as economically feasible as it originally appeared.

8.2.2 Process Control and Instrumentation

Process control and instrumentation have largely been ignored in the notes presented for the preliminary design (Part I), certainly not because these topics are unimportant or because they can be safely left until late in the design work. The opposite applies in both cases. However, this subject is broad and is changing so rapidly that like the economic analysis, to provide a complete coverage would require a separate volume to be written. The design of process control systems and the specification of instrumentation are subjects that apply to specific plants and processes, any general description would be inadequate in most situations, however some typical guidelines and ideas can be presented.

Process control and the associated instrumentation were mentioned in relation to the process selection (Section 3.1). To reiterate those comments, these aspects of plant design must be considered during the process selection stage and at other subsequent stages in the design of the plant. Do not assume that any item of equipment or an entire plant that can be designed can also be operated and controlled efficiently. It should be established early in the design that the control and instrumentation aspects are not only feasible but that the cost is within acceptable limits.

Ignoring the control and instrumentation requirements may mean that (at best) the capital cost is higher than necessary, and that operation and control remain difficult.

Process control is such an important topic in the design of increasingly complex chemical plants that the student should have been exposed thoroughly to the basic principles, and to applications in chemical engineering situations. The design project provides an opportunity to apply this basic knowledge to design problems. A detailed control strategy for an entire plant is beyond the scope of a student project (and the time available), however at least one unit or a section of the plant should be considered in detail.

Instruments which can monitor the important process variables during plant operation must be specified. These instruments must be capable of measuring the variables and should have an acceptable accuracy and repeatability of measurement, usually the latter attribute is more important than the former on chemical plant measurements. The instruments may be used for manual measurements or included in automatic control loops. Automatic alarms may also be required to indicate deviations outside acceptable limits. If possible, direct measurement of the process variable should be made, however it is often easier to measure a dependent variable, e.g. temperature measured as an indication of composition for distillation column top product.

The specification of a control scheme and the associated instrumentation for a chemical plant should satisfy several main objectives. First, the plant should operate at all times in a safe manner. Dangerous situations should be detected as early as possible and appropriate action initiated, also the process variables should be maintained within safe operating limits. Second, the plant should operate at the lowest cost of production. Finally, the production rate and the product quality must be maintained within specified operating limits. These objectives may be conflicting, and the final control scheme to be adopted is based upon a realistic and acceptable compromise between the various factors. The main conflict is between the need to design and operate as safe a plant as possible and the desire to produce the chemical at the lowest cost. Safe plant operation can be expensive, both in terms of the capital cost of instrumentation and the annual operating costs, e.g. maintenance.

Experienced process control engineers are usually responsible for the design and specification of automatic control schemes on large chemical plants. The book by Shinskey (1979) provides details of the practical

aspects of process control-system design. The P & ID is used to prepare the preliminary instrumentation and control scheme. Control loops are identified for level, flow, pressure and temperature controls to ensure steady plant operation. The first steps in the design of a good control system are to define which variables need to be measured and which need to be controlled. To specify pressure, temperature, flow and level controllers on all units at different locations, and to assume that the measurements will provide the basis for adequate control is unrealistic. It is necessary to specify:

(a) the variables that *need* to be measured;
(b) the location of these measurements;
(c) the variables that *need* to be controlled;
(d) how the desired control can be achieved.

A control strategy hopefully can then be evolved. It is also necessary to determine the ancillary instruments required for process monitoring, and for plant commissioning and start-up. It is worthwhile at the design stage including connections for instruments which may be required in the future. The location (local or a distant control room) and types of recorders must be specified, and the alarms and interlocks that are required. Control does not occur independently in different units, and an overall and integrated approach to the control of the plant is required.

Action: *Prepare detailed designs for the chemical engineering units in the plant. Consider energy conservation measures and the process control and instrumentation required as the designs are performed. Prepare a design specification sheet for each unit.*
Detail the specific energy conservation schemes considered and adopted,
Discuss the process control strategy for the overall process and the instrumentation required.
Prepare a complete P & ID for the process.

8.2.3 Safety, Loss Prevention and HAZOP

Factors relating to the safety of a chemical plant have so far been ignored. This is mainly because it is impossible to discuss all the aspects of plant design initially in a set of notes. The designer will obviously design items of equipment, and perform other tasks in process design, in

order to ensure safety during operation — no one is going to deliberately design a plant to operate in an unsafe manner. However, the designer cannot envisage all possible malfunctions that can occur, whether due to human error or mechanical failure. Also, the occurrence of minor operating changes in one unit can have serious consequences on equipment performance in another section of the plant.

All process design activities, e.g. process selection, plant layout, etc., and the equipment designs must be carried out with the design objective of ensuring (within acceptable risk levels) the safe operation of the plant. To achieve this aim, safety and plant operation must be considered at several stages in the preliminary design work. After the detailed plant design is completed, it is necessary to perform a detailed and complete safety analysis. This study should identify aspects of the design that may cause operational or safety problems, and also any modifications necessary to the equipment design (or the control scheme) which minimise the effects of changes in the plant operating conditions. The following activities are carried out in order to achieve these aims.

(a) A safety review of all equipment designs, e.g. design to appropriate standards and observation of design code recommendations. This is really checking that a 'good' design has been performed. However, mistakes do occur and a second opinion and review by a group of experts can help to uncover some errors.

(b) *Loss Prevention Studies.* Loss prevention is the general name given to the activities that help provide anticipatory safety measures for the *prevention* of accidents. The techniques that are used include:

(i) *Hazard and Operability (HAZOP) Studies,* sometimes simply referred to as Operability Studies, provide a systematic and critical examination of the operability of a process. They indicate potential hazards due to deviations from the intended design conditions. The techniques can be applied to existing plants, and should be standard procedure at the process design stage of new plants.

(ii) *Failure Analysis* is described by Leach (1972) and provides a crude criterion of acceptance. The overall reliability of a chemical plant can be calculated from the reliability of individual components, consideration is given to parts of low reliability e.g. duplication of equipment. Powers and Tompkins (1973) describe this procedure as applied to a chemical plant.

(iii) *Hazard Analysis (HAZAN)* is described by Lawley (1973), and is a

technique for the quantitative assessment of a hazard, **after** it has been identified by an operability study, or similar technique. HAZAN is used to compare the risks to life per hour of exposure. Attention is concentrated on those risks which exceed a given acceptable value, e.g. 3 $\times 10^5$ deaths per year.

A *HAZOP* study is a detailed and formal evaluation of a process and is performed section-by-section (or line-by-line of the flowsheet), in order to consider operational deviations and their possible effects. The study is based upon a set of 'guide words' (identifying possible deviations), examples are: 'NONE', 'MORE OF', 'LESS OF', 'PART OF', etc. A list of guide words and their precise interpretation is included in the Chemical Industries Association booklet (1979), and in Coulson and Richardson, Volume 6 (1983; pp. 292-294). Each guide word is applied systematically to the 'property words' (e.g. flow, pressure, temperature, etc.) which are chosen as being relevant to the design conditions and the design intention. A typical sequence of events comprising a detailed operability study is presented in Coulson and Richardson, Volume 6 (1983; p. 295, Fig. 9.3). The results of the *HAZOP* study are usually presented in tabular form with the following headings:

Property word;   Guide word;   Cause;   Consequence;   Action.

It is important to identify not only what may happen (consequences) and why (causes), but also what should be done (actions) to prevent serious accidents. Examples of operability studies in particular situations are included in Coulson and Richardson, Volume 6 (1983; pp. 292-294, 296298). Modifications initiated in one section of the plant can have effects in other sections, it is therefore necessary to consider the effects of any changes that are made on the overall operation of the plant. The existence of recycle lines requires careful consideration in the study. A second operability study may need to be performed after all the actions have been incorporated into the design and included on the flowsheet.

Note In student design projects, the aim is to produce a competent design and also to communicate to the supervisor what has been done. It is therefore essential to include a summary of the major findings obtained by performing the *HAZOP* study (this is usually only performed on one

item of equipment or a section of the plant for student projects, a complete study would require too much time and is not necessary for teaching purposes). The student should not assume that the market will read, analyse and digest several pages of an operability study in detail. Therefore, identify in a summary the most important consequences and actions arising from the study.

Safety check lists can provide a useful guide for the main items to be considered during the process design stage. It must be remembered that no list is ever complete and other considerations and techniques should also be applied. Safety check lists can be found in Coulson and Richardson, Volume 6 (1983; pp. 301-303); IChemE Flowsheeting for Safety (1976); Wells (1980); and Balemans (1974).

The preceeding discussion may give the impression that loss prevention, operability and safety are only considered during the design phases of a project. This is certainly not the case and the safe operation of the plant must also be evaluated during the construction phase, commissioning, and during normal operation. Plant start-up, shutdown, and performing modifications to the plant, are particularly high-risk activities in respect of safe operation. These departures from normal operation must be carefully assessed. The following summary describes the relevant studies during the various stages of a project.

Project Stage	*Details of Study*
1. Feasibility evaluation	Identify major hazards and inherent process risks (toxic, explosive).
2. Process design	Systematic quantitative analysis of hazards and definition of measures to reduce them.
3. Engineering design	Formal examination of engineering diagrams with quantitative hazard analysis to ensure the above measures have been implemented.
4. Construction	Systematic check that hardware (as constructed) and operating instructions are as intended in the design.
5. Operation	Post start-up review of all departures from design affecting safety, operating difficulties,

equipment failures, etc., with hazard implications. Plant modifications and changes in operating procedures or conditions, checked for effect on plant safety.

Much has been written about Loss Prevention, and **HAZOP** studies in particular, the volume of publications has grown considerably since the Flixborough disaster in 1974. Standards, Codes of Practice and Acts of Parliament provide essential information, and the references included here provide a good overview of the subject, 'and a basis for more detailed study.

Action: **Prepare a HAZOP study for a particular item (or section)** *of* **the chemical plant.**
Identify **major hazards and necessary actions.**
Summarise **the conclusions in relation to the design** *of* **the plant.**

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Case Study — Summary for Part II: Detailed Equipment Design

Part II of the Design Project (Case Study) for the production of nitric acid concerns the design of two main plant units (the NOx absorption column and the steam superheater), a pump to deliver 'red' product acid from the absorption column to the product bleaching column, and finally a product storage tank.

Absorption of the nitrogen oxide components (NOx) of the process gas stream is conducted in a sieve tray-type absorption column. This tower contains 59 sieve trays, of which the top 45 trays contain herringbone cooling coils to remove heat of reaction/dilution and maintain low absorption temperatures.

The steam superheater is a clamp ring-type, floating-head, shell and tube, heat exchanger. It is able to produce up to 5775 kg/h of steam at 380°C and 4000 kPa.

A single-stage single-suction centrifugal pump is recommended to deliver 'red' product nitric acid from the base of the absorption tower to the product bleaching column.

The nitric acid storage tank proposed for this plant will provide a product storage capacity of one week supply in the event of a plant shutdown in the adjacent ammonium nitrate facility. The tank has a capacity of 1950 m^3 (representing 1450 tonnes of product acid).

Case Study — Amendments to Part I

Following the detailed equipment design, a few amendments need to be made to the information presented in Part I of the Design Project Report (Case Study). The two main areas needing amendment are detailed as follows.

1. Mass Balance around the Absorption Column The mass balance around this unit was refined due to the availability of better chemical equilibrium data (Ref. AI, see Chapter 9). This data predicted that the dissolved gas concentration in the 'red' product acid stream flowing from the column would in fact be of lower concentration than initially thought, and the dissolved gas component tended to be nitrogen monoxide and dioxide rather than nitrogen tetroxide. Therefore,

amendments need to be made to the inlet gas composition and flowrate, to the red product-acid flowrate and composition, and to the tail-gas stream. These changes are shown below on an hourly basis.

Absorber gas feed	Nitrogen + inerts	*27 842 kg*
	Oxygen	1 *502* kg
	Nitrogen monoxide	*117* kg
	Nitrogen dioxide	*2* 099 kg
	Nitrogen tetroxide	1 *302* kg
	Total	*32 862* kg
Product acid	Nitric acid	*7* 000 kg
	Water	*4 667* kg
	Nitrogen monoxide	*2 kg*
	Nitrogen dioxide	*116* kg
	Total	11 785 kg
Tail Gas	Nitrogen + inerts	*27 842* kg
	Oxygen	*256* kg
	Nitrogen monoxide	8 *kg*
	Nitrogen dioxide	*25* kg
	Nitrogen tetroxide	*5* kg
	Total	*28 136* kg

2. Equipment Costs Due to the availability of more complete information regarding the type of equipment required, several costs are revised.

Absorption column	As870 000
Steam superheater	As 25 000
Bleaching column delivery tank	As3500
Nitric acid storage tank	As390 000

Note Chapters 9 to 12 contain case study material for the detailed design of equipment for the nitric acid plant. There are no accompanying text notes. References are included at the end of each chapter. Details of the calculations performed and additional information are included in Appendices G to J.

CHAPTER 9
Case Study — Absorption
Column Design

Summary

AN ABSORPTION column is required to absorb the nitrogen oxide
components (NOx) from the gaseous reactor effluent by contacting
this gas with an aqueous medium. Three incoming streams are
handled by the column. The first is the gas stream of 32 570 kg/h with
12% (wt./wt.) nitrogen oxides. It enters at a temperature of 65°C and
a pressure of 960 kPa. The second stream contains weak-acid
condensate from the oxidation unit. It is 42% (wt.) aqueous nitric
acid solution, at 50°C and 970 kPa. The final stream is deionized
water at 7°C and 950 kPa.

The product specification requires 11 670 kg/h of 60% (wt.) nitric
acid solution, excluding dissolved nitrogen oxides. This specification
must be achieved while restricting the level of tail-gas emissions to
below 1000 ppm of nitrogen oxides. This acid product requires
approximately 2340 kg/h of deionized make-up water.

A sieve tray-type absorption column is proposed. The design
specifies a column 1.8 m diameter, approximately 32 m high, and
containing 59 sieve plates. Weak-acid condensate is added to tray 13
and make-up water is added at tray 59 (the top tray). Crossflow-type
trays are employed from plates 1 to 13, however, a decreased liquid
loading demands reverse flow-type trays for plates 14 to 59. The
operating pressure is approximately 950 kPa and the operating
temperature range is from 8°C to 65°C.

Cooling coils are required to remove unwanted reaction heat and
to reduce the absorber tray temperatures to favour absorption. Two
herringbone-type coils are included on each tray from plate 14
upward. The duty demanded from each cooling coil is approximately
3 kW.

A rigorous mathematical model was developed for this system using the correlations and data presented in Refs. Al and A2. The model enabled the sizing of the column and the optimization of tray spacing. It was run on the APPLE Macintosh (512K) personal computer.

The absorption column and its associated structures (platforms, ladders and railings) are expected to cost in the region of AS870 000. The final design comparesfavourably with similar units already found in industry.

Detailed calculations concerning the absorption column design are presented in Appendix G.

References Used: Al to Al 6

9.1 Introduction

This section discusses the design of a suitable absorption column for the production of nitric acid by the single-pressure process. A comprehensive design study determined the column mechanical details and physical characteristics, together with its anticipated performance. This chapter contains the operating criteria and product specifications, the solution method, and finally the results of the design calculations. Details of the design calculations and all associated data are included in Appendix G.

The absorption column is required to absorb nitrous reaction gases thus producing 80% (wt.) nitric acid product (dissolved impurity-free basis). This is achieved by the countercurrent absorption of the nitrogen oxide components from the reaction gas into a water/weak-acid media. The column specification requires an operating pressure of 950 kPa and an absorption temperature in the range of 10°C to 65°C.

The design must consider three feed streams and two product streams. The three inlet feed streams are 'strong' reaction gases, weak nitric acid solution and make-up water. Two outlet streams flow from the column. These are a 'lean' reaction gas (tail-gas) stream and 'red' product acid. Absorption of nitrous oxides increases as the temperature is reduced. This effect, together with the exothermic oxidation/absorption reactions, requires installation of an internal cooling circuit.

Only tray-type columns were considered because of the difficulty of incorporating an effective cooling'circuit into a packed column. Sieve trays (as opposed to bubble or valve-type trays) were preferred because of the ease of installing cooling coils and also their low unit cost. Details of tray selection are included in Appendix G.l.

9.2 The Design Method

The overall design approach consists of several design aspects or stages. The key elements are a computer-based mathematical model to determine the column diameter and number of trays, an iterative sieve tray design suggested in Ref. A3, and a mechanical design as specified in the Australian Standard: AS1 210 (Ref. A10). The process control scheme is specified based upon appropriate operating considerations (using Refs. A7 and A8). The important aspects of this design approach are outlined in the following sections. Details of calculations are included in Appendix G.

9.2.1 The Mathematical Model

The design method employs a mathematical model written in BASIC, which can be run on the APPLE Macintosh. This model was developed by incorporating modelling information appropriate to the system equilibria (available in Refs. AI and A2). A listing of this program, its documentation and its description are included in Appendix G.6. The main features are summarized below.

(a) A tray-by-tray approach based upon the specified feed compositions, feed temperatures and operating pressure.
(b) A provisional estimate concerning the sieve tray design and column dimensions. Refinements are made manually later.
(c) The program calculates any further oxidation of nitrogen monoxide in the void space prior to passing through a tray according to pressure, temperature, and partial pressure of nitrogen monoxide and oxygen.
(d) The equilibrium between nitrogen dioxide and nitrogen tetroxide is refined. Again this equilibrium is temperature dependent, and is sensitive to the partial pressure of nitrogen dioxide.

(e) The amount of nitrogen peroxide (i.e. dioxide and tetroxide) absorbed, as the gas stream passes through liquid on the plate, is calculated. This value is determined in accordance with the temperature, pressure, partial pressure, and tray efficiency. The tray efficiency is dependent upon liquid depth, hole size, gas velocity, liquid flowrate and tray spacing.

(f) The mass balances are recalculated and the program proceeds to the next plate. The same procedure from step (c) is applied.

(g) The program continues until all the acid required to be formed by the absorption is achieved. The number of trays required in the column is thus determined.

(h) The tail-gas composition and required make-up water are also calculated.

The main feature of this program is that the tray efficiency is calculated within the program. The tray efficiency is dependent upon the individual geometric properties of each tray. The program also performs an energy balance over each tray. The exothermic oxidation reactions are considered, and also the heats of dilution, heats of formation of acid and thesensible heat of the incoming and outgoing streams. Hence the cooling duty on each tray is determined. This model is used to find the optimum column diameter, the number of trays required, and also the cooling duty required on each plate.

Refer to Appendix G.2 for more details of the absorption column model and associated calculations.

9.2.2 Sieve-Plate Hydraulic Design

The sieve plates are designed in accordance with the recommendations made in Refs. A3 to A6. The iterative design approach was suggested in Ref. A3 (p.458). The procedure is summarized below.

(a) Calculate the maximum and minimum vapour and liquid flowrates for the turn-down ratio required.

(b) Obtain data for the physical properties of the system.

(c) Select a provisional plate spacing.

(d) Estimate the column diameter based upon flooding considerations.

(e) Decide upon the liquid flow arrangement.

(f) Specify an initial plate layout including downcomer area, active area, hole area, hole size, weir height.

(g) Check the weeping rate, if unsatisfactory return to step (f).

(h) Check the plate pressure drop, if unsatisfactory return to step (f).

(i) Check the downcomer backup, if too high then return to step (f).

(j) Decide plate layout details including calming zones and unperforated areas and check the hole pitch, if unsatisfactory return to step (f).

(k) Recalculate the percentage of flooding based upon chosen column diameter.

(I) Check for entrainment, if too high then return to step (f).

(m) Optimize the design by repeating steps (c) through (I) and determine the lowest cost design.

(n) Finalizethedesign, draw up the platespecification and sketch the plate layout.

This method, in conjunction with standard design recommendations, is both simple and effective in determining the final sieve-plate design. Details of the necessary calculations are presented in Appendix G.3.

9.2.3 Mechanical Design of Column

The mechanical design of the column incorporates three main stages. The first specification is for the materials of construction. Second, the shell and head thickness must be chosen in order to withstand the operating conditions and also extraneous forces. Finally, consideration of the construction and assembly. The design was performed according to the relevent Australian Standard, AS1 210: SAA Unfired Pressure Vessels Code (Ref. Al 0).

For this application, stainless steel 304L ('nitric acid grade') is the only practical material available. The shell and head thickness specification was based upon matching the operating and extraneous stresses encountered in this stainless steel against the recommended design stress (from appropriate codes). If the calculated stresses exceeded the design stress then a thicker plating was chosen. The optimum plate thickness was chosen according to the

above approach, while considering only the standard steel-plate thicknesses available through BH P in Australia (Ref. Al 1).

Construction considerations and the tolerances required were in accordance with recommendations in Refs. Al 0 and Al 2. Details of the calculations performed are presented in Appendix G.4.

9.2.4 Process Control Scheme

The process control scheme proposed for the column is based upon similar suggested layouts used for the control of distillation columns (Ref. A8). Modifications were made in accordance with the individual operating characteristics and requirements of this system. Final verification of the proposed scheme is achieved by consulting Refs. A7 and A9.

9.3 Important Operating Considerations

Before specifying the various inputs and unit requirements for the design, it is necessary to identify the most significant variables for this procedure. There are many important variables for good absorption in this particular system. The analysis of the unit is made even more complex by having to deal with three feed streams and satisfy design quality criteria on both of the two product streams. Fortunately the chemistry of the process is relatively simple, with only three nitrogen oxides present, i.e. NO, NO_2 and N_2O_4 (other nitrogen oxides are produced in negligible amounts under the operating conditions proposed). Analysis of the process chemistry and the effects of the relevant variables are contained in Appendix G.2 and Section 1.2.4. The main aspects are listed below.

(a) Formation of nitric acid occurs primarily by the reaction of nitrogen dioxide with water, although some acid forms from the tetroxide. The major reaction is shown below.

$$3NO_2 \text{ (g)} + H_2O \text{ (l)} \rightleftharpoons 2HNO_3 \text{ (aq)} + NO \text{ (g)-} \text{ REACTION 1}$$
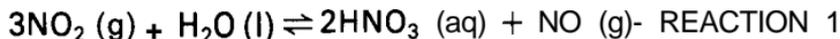
This is an exothermic reaction which is dependent on the partial pressure (therefore total operating pressure) of nitrogen dioxide.

DESIGN CONDITIONS

Code: AS1210
Design Pressure: 1200 kPa
Efficiency of joints: 85%
Shell Thickness: 16 mm
Head Thickness: 16 mm

OPERATING PARAMETERS

Weak Acid Feed (Plate 13): 5530 kg/h
Make-up Water Feed (Plate 59): 2340 kg/h
Gas Feed (column base): 32570 kg/h
Operating Pressure: 950 kPa
Operating Temperature: 10–60 °C

DESIGN NOTES

Stainless Steel 304L throughout the column (Nitric Acid Grade).

Fillet weld joining column cylinder to ellipsoid heads.

Liquid feed inlets via 50 mm (2") ducting

Liquid outlet via 75 mm (3" ducting).

Gas feed and outlet both 100 mm (4") ducting.

1.8 m
Ellipsoid Column Head = 700 mm
Manhole 450 mm
Mist Eliminator
Tray Baffle
Water Inlet
Cooling Coils
Baffle Height 500 mm
Plate Spacing 500 mm

Weak Acid Inlet
PLATE 13
PLATE 2
PLATE 1
NOx Gas Inlet
Manhole 450 mm
"Red" Product Acid Outlet

Figure 9.1
NOx
ABSORPTION COLUMN

By David JOHNSTON

FIGURE 9.1 NOx absorption column

(b) Secondary reactions concern the equilibrium of nitrogen dioxide with the tetroxide form, and the oxidation of nitrogen monoxide (formed in the production of acid) to nitrogen dioxide.

$$2NO_2 (g) \rightleftharpoons N_2O_4 (g) \qquad \text{-REACTION 2}$$
$$2NO (g) + O_2 (g) \rightleftharpoons 2NO_2 (g) \qquad \text{— REACTION 3}$$
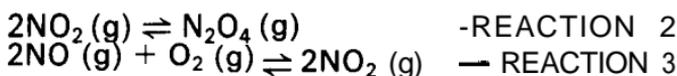
Reaction 3 is relatively slow and is regarded as the rate-determining step in the entire process. Other key reactions proceed virtually instantaneously. Both Reactions 2 and 3 are exothermic, and are dependent on the various partial pressures (operating pressure) and temperature. Lower temperatures favour the overall absorption chemistry. This factor, together with the exothermic nature of the reactions, determines the need for a cooling circuit within the column.

(c) The dissolution of both nitrogen monoxide and nitrogen dioxide into aqueous nitric acid is inversely proportional to their partial pressures. The amounts dissolved decrease with increasing temperature.

(d) In the mechanical considerations those factors of greatest significance are column diameter and plate spacing. These factors determine the average residence time of the gases in the space between the plates, and hence the degree of oxidation of nitrogen oxide (the rate-determining step).

(e) Tray design is also of critical importance. Tray area, weir height, and the size and spacing of tray holes are all important variables. These factors and the gas velocity determine the tray efficiency. Tray efficiency has a major effect on the overall column performance.

9.4 Design Constraints

There are several performance parameters required from this unit, including the production of 60% (wt.) nitric acid product (on a dissolved gas-free basis). This product must be obtained while ensuring the tail-gas emissions are kept below 1000 ppm.

The various design requirements are presented in Table 9.1, and a drawing of the equipment in Figure 9.1. The mass balance has been recalculated based upon better information regarding the

TABLE 9.1

Important design criteria for absorption column design

Operating Parameters	
Operating pressure	950 **kPA**
Operating temperature	**65°C** to 10°C
Column Inputs	
Reaction gas feed (at **65°C** and 960 **kPA**):	
Nitrogen and inerts	*27* 500 kg
Oxygen	1 500 kg
Nitrogen monoxide	120 kg
Nitrogen dioxide	2 150 kg
Nitrogen tetroxide	<u>1 *300* kg</u>
Total	*32 570* kg
Weak-acid condensate (at 50°C and 970 **kPa**)	
Nitric acid (42% wt.)	*5530* kg
Make-up water (deionized at **7°C** and 960 **kPa**)	
Water	*2340* kg
Output Requirements	
Product acid (60% wt.)	11 670 kg
	(NO, free basis)
Tail- gas emissions (at **10°C**)	**< 1000 ppm**

concentration of dissolved gases in the 'red' acid product (dissolved gases are ultimately stripped from the acid and returned as absorber feed).

The design procedure was undertaken based upon these amended values, with the computer model run many times in order to refine the solution and optimize the column size and tray design.

9.5 Absorption Column Specification

The results of the design calculations are presented in this section. Detailed calculationsare included in Appendix G.3 and G.4. The final design specification based upon these calculations is summarized in Table 9.2.

The design specification in Table 9.2 must be complimented by suitable construction tolerances. These suggested tolerances (from Refs. A10 and Al 2) are itemised below. They are based upon

TABLE 9.2

Summary of absorption column design specification

Sieve tray-type column	59 trays
Crossflow-type plates	Numbers 1-13
Reverse flow-type plates	Numbers 14–59
Weak-acid feed plate	Number 13
Make-up water feed plate	Number 59
Column inside diameter	1.8 m
Column outside diameter	1.832 m
Column height (total)	32 m
Plate spacing (equal for all plates)	500 mm
Operating pressure	9.5 bar
Design pressure	12 bar
Operating temperatures	10–65°C
Design temperature	100°C
Vessel thickness-shell	16 mm
-heads	16 mm
Internal corrosion allowance	5 mm
Materials of construction	SS304L ('nitric acid grade')
Design Code AS1210	
Legal construction; self supporting; no internal lining; no insulation; Class 1	

The detailed design drawing for the absorption column is presented in Figure 9.1

standard engineering practice and upon operational contraints.

(a) Overall vertical alignment of the tower from base to top is set at 20 mm.

(b) Bend line to bend line is 20 mm.

(c) Tower base out of level 5 mm (maximum) at outer edge of base plate.

(d) Plate spacing 500 $\pm$ 5 mm.

(e) Tolerance between adjacent trays is 10 mm.

(f) Height of both weirs on lower plates is 50 $\pm$ 5 mm, and on upper plates is 40 $\pm$ 5 mm.

(g) Downcomer area (weir positioning) $\pm$ 5%.

(h) Feed nozzle arrangements $\pm$ 5 mm.

9.6 Sieve Tray Specifications

The essential design and construction features of the absorption column are described in Section 9.5. It is now necessary to look more

closely at the sieve plates. Each plate represents a detailed item of design work.

The column contains two distinct sieve-tray hydraulic designs reflecting the change in liquid loading that takes place above the weak-acid feed plate. Plates at and below the weak-acid feed plate (plates 1 to 13) must handle a liquid flowrate of approximately 8 m^3/h. Those plates above the weak-acid feed plate (plates 14 to 59), however, have a significantly lower liquid loading of up to 3 m^3/h. This change in liquid loading necessitates a 'reverse-flow' type sieve plate featuring a central baffle.

The two tray geometries are shown in Figures 9.2 and 9.3. Figure 9.2 illustrates the 'cross-flow' type plate. The dimensions and features are listed in Table 9.3. Figure 9.3 demonstrates the 'reverse-flow' type plate, its dimensions and features are listed in Table 9.4.

TABLE 9.3

Cross-Flow   Type   Plates

Plate Number	1 to 13	Plate Material	SS304L
Plate i.d.	1.8 m	Downcomer Material	SS304L
Hole Size	5 mm	Plate Spacing	0.5 m
Hole Pitch	13 mm Δ	Plate Thickness	5 mm
Active Holes	10 200	Plate Pressure Drop	0.95 kPa

TABLE 9.4
Reverse-Flow   Type   Plates

Plate Number	14 to 59	Plate Material	SS304L
Plate i.d.	1.8 m	Downcomer Material	SS304L
Hole Size	5 mm	Plate Spacing	0.5 m
Hole Pitch	15 mm Δ	Plate Thickness	5 mm
Active Holes	9 200	Plate Pressure Drop	0.83 kPa

Operational flexibility is enhanced by securing the weir and downcomer in slides. These slides enable adjustment of weir height should the operating parameters change appreciably.

The plates from tray 1 to 13 are identical, while trays 14 through to 59 all utilize a common design (although different from those in the lower section).

The tray arrangement is made more complex due to the incorporation of cooling coils. In the original column specification
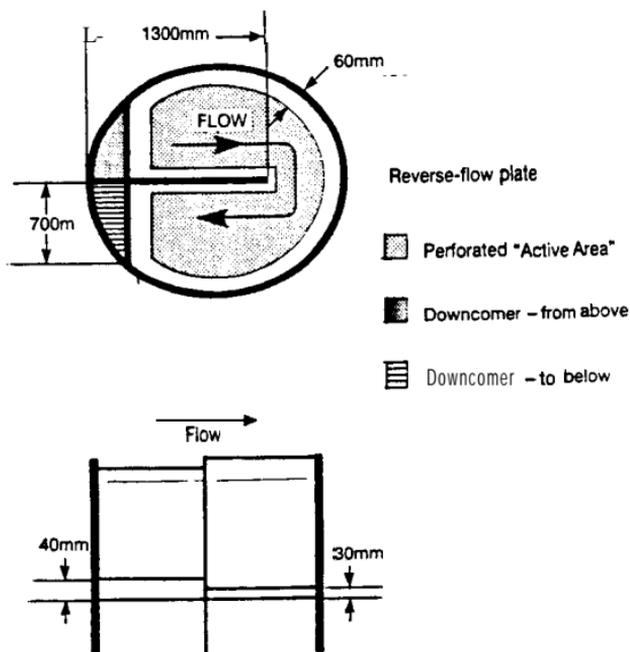
FIGURE 9.3   Sieve-plate specification: Plates above the weak-acid feed plate.
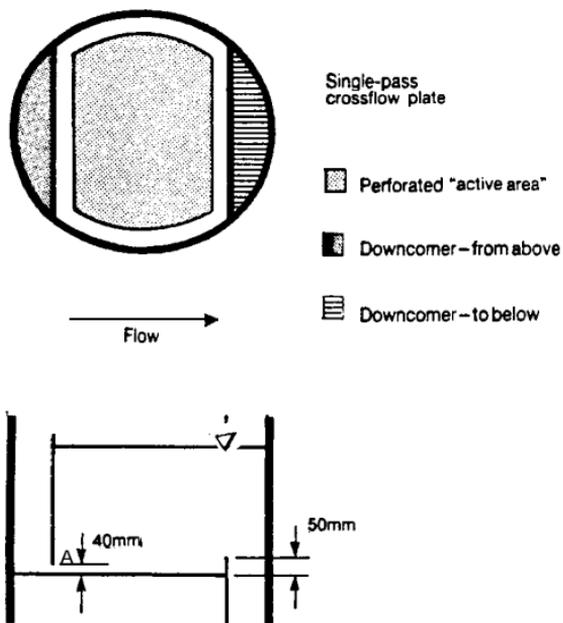
FIGURE 9.2   Sieve-plate specification: Plates at or below the weak-acid feed plate.

(Sections 4.2 and 7.4) it was anticipated that two cooling circuits would be required, one above the acid feed plate and one below. However, a more rigorous examination of the tray-by-tray heat balance (obtained from the computer-based model) suggests this may be significantly simplified. Plates at and below the weak-acid feed show an approximate balance in heat flow. The bottom plates are actually net consumers of heat. Therefore, it was decided to leave the bottom 13 plates free of cooling coils. This approach is advantageous because higher temperatures at this end of the column lower the concentration of dissolved nitrogen oxides in the acid solution.

The cooling duty on the column occurs on the top 46 plates. The energy balance provided by the mathematical model suggests a cooling duty ranging from 5750 W to 6500 W for these plates. This leads to the specification for herringbone-type cooling coils (see Figure 9.4). These reverse hairpin arrangements provide a large surface area per unit volume. They will be laid flat onto the sieve plates, one on either side of the plate baffle (see Figure 9.5). Details of the coil specification are given in Table 9.5.

9.7 Process Control Scheme

The process control scheme for the absorption column is presented in Figure 9.6. It was designed from the recommendations presented in the *HAZOP* analysis, the results of which are reported in Section 9.8. It features ratio control on the make-up water stream. The signals from flow transmitters on this line and on the gas input line are fed to the ratio controller, whereby the make-up water stream is adjusted.

Other control features include a pressure controller on the tail-gas outlet stream so that the column absorption pressure can be maintained at the design operating value of 950 kPa. A temperature transmitter on the tail-gas outlet stream provides the signal for control of the overall cooling-water flowrate. This is the temperature which is most useful in determining good absorption. The cooling circuit itself is fed from a common line (on which the overall flowrate is controlled). Small block valves on each of the tray cooling-coil feed lines enable flowrate regulation to each of the coils. These valves feature a removable top whereby a magnetic flowmeter may be
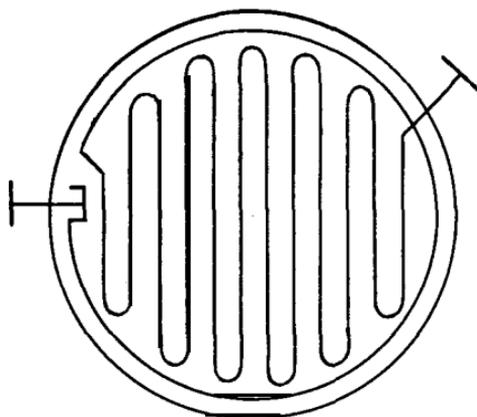
FIGURE 9.4 Herringbone cooling coils (two coils per sieve plate)
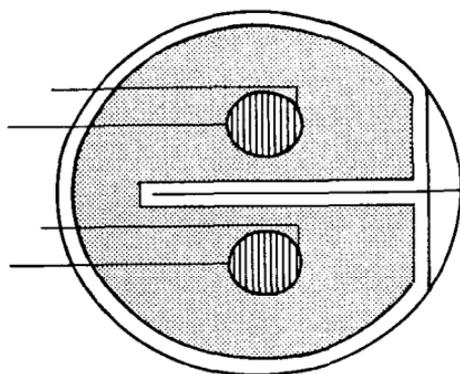
FIGURE 9.5 Coil positions on the sieve plate.

TABLE 9.5
Specification of tray cooling coils

Herringbone-type coils	
Two per plate (one either side of baffle)	
Outside diameter	*32* mm
Average heat-transfer coefficient	170W/(m²K)
Mean temperature difference	10°C
Effective length	10 m
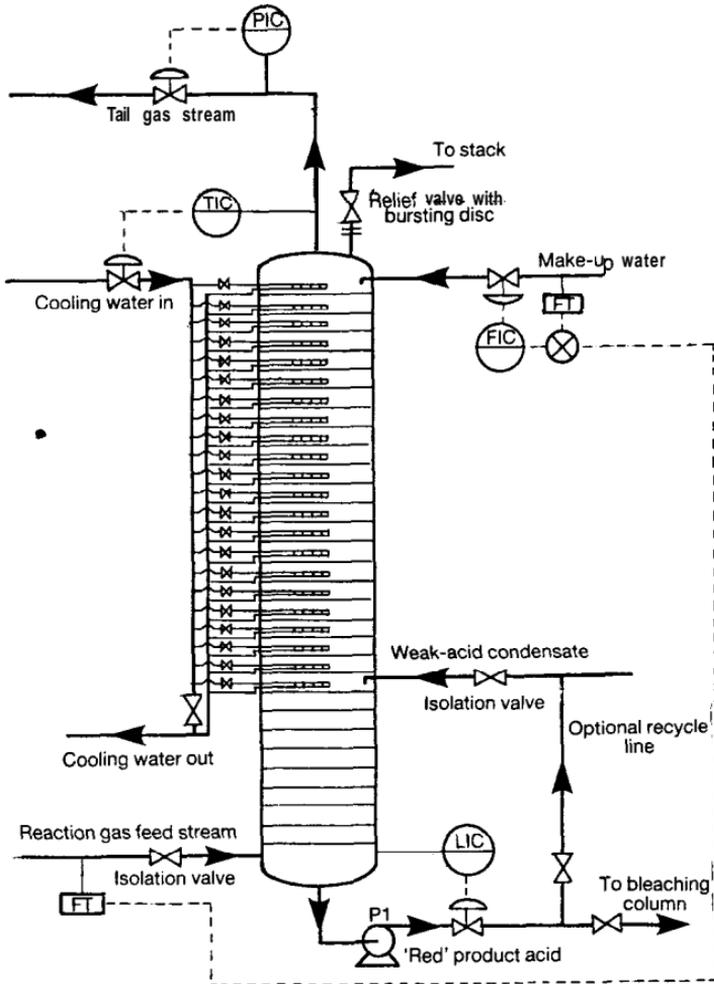
FIGURE 9.6 Control scheme for NOx absorption column.

inserted to read the flowrate. The valves need only be set initially and then periodically adjusted manually.

There is no automatic control on the flowrate of the gas inlet stream or weak-acid condensate stream, since both of these flows are predetermined by feed flowrates earlier in the process. Isolation valves and provision for spectacle blinds are included to enable the column to be isolated during shutdown periods.

The product-acid solution is withdrawn from the column using a level control valve on this line. The liquid level in the base of the

column must be maintained slightly above the level of the plate downcomer to prevent incoming gas from by-passing the sieve plates.

All controllers suggested for the absorption column feature *HIGH* and LOW alarms for good control.

The final safety requirement is a relief line with a relief valve protected by a bursting disc. This disc must be sized according to the Australian Standard: AS1 210.

9.8 Hazard and Operability Study

Having'specified the detailed design of the absorption column, the Hazard and Operability Study *(HAZOP)* can now be performed. This study is used to determine the process control requirements and also to gain an appreciation of the difficulties that may occur during operation of this unit. The HAZOP study comprises a detailed examination of each of the streams flowing into and out of the absorption column. All significant aspects of deviation for each of these streams are then considered under 'Key Word' headings. The HAZOP approach used here follows that suggested by the Institute for the Prevention of Accidents in their HAZOP publication. The results of the HAZOP are shown in Tables 9.6 to 9.11.

The HAZOP study was instrumental in determining the need for an adequate alarm system on each of the specified controllers. If liquid levels within the column are not well controlled, then either flooding (too much liquid) or plate by-passing bythegas (too little liquid) will occur. Both situations lead to a substantial reduction in absorption efficiency with large increases in emission levels. The other important control parameter was shown to be the temperature. If the temperature in the cooling-coil section rises, then there is an appreciable reduction in absorption. Control of temperature is important in the upper sections of the column because it is here that the greatest effect on emission levels occurs.

9.9 Discussion of Results

The optimum balance between column diameter and tray height is judged to be 1.8 m. This column diameter allows for a 20% margin in

TABLE 9.6

HAZOP study: Weak-acid condensate stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
NONE	No Flow	1. Pump failure.	Deficient quality product and high NOx tail gas emission levels.	a) Install LOW LEVEL ALARM on LIC at the base of the absorption column.
		2. Valve fails shut.	As for I Valve overheats.	Covered by a). b) Install kick-back on pumps
		3. Line fracture.	As for I and 2.	Covered by a) and b). c) Regular inspection and patrolling of weak-acid transfer lines and seals.
MORE OF	More Flow	4. Higher humidity in feed air.	Higher make but weaker product acid.	d) Install a HIGH LEVEL ALARM on LIC at the base of the absorption column.
	More Temperature	5. Higher feed rates causing larger heats of reaction.	Possible higher NOx emission due to lower absorption.	See Table 9. I 1.
	More Pressure	6. Isolation valve is closed in error while pump running.	Lines subject to full delivery pressure.	Covered by b) f) Perhaps worthwhile installing a pressure gauge upstream of the delivery pump.
		7. Thermal expansion on the isolation valve section (fire).	Possible line fracture or flange leakage.	g) Provide thermal expansion relief in the valved section.
LESS OF	Less Flow	9. Flange leakage or valve stub blanked but leaking	Decreased absorption. Lower product make.	Covered by a),c), and d).
	Less Temperature	10. Reaction gas temperature in oxidation unit lower.	Increased dissolved NOx concentrations in product acid.	See Table 9. I 1
OTHER	Maintenance	11. Equipment failure, flange leak, catalyst changeover in reactor, etc.	Process stops.	Ensure all pipes and fittings are constructed of the right materials and arc stress relieved.

TABLE 9.1

HAZOP study: Make-up water feed stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
NONE	No Flow	1. Pump failure.	High emission of NOx in tail gas.	a) Install LOW LEVEL ALARM on LIC at the base of the absorption column.
		2. Valve fails shut.	As for I. Valve overheats.	Covered by a). b) Install kick-back on pumps.
		3. Line fracture.	As for I and 2.	Covered by a) and b). c) Regular inspection and patrolling of weak-acid transfer lines and seals.
MORE OF	More Flow	4. Control valve fails open.	Dilute acid product is formed.	d) Install a HIGH LEVEL ALARM on LIC at the base of the absorption column.
	More Temperature	5. Higher feed temperature to the refrigeration unit. 6. Failure in the refrigeration unit.	Possible higher NOx emission due to lower absorption. As for 5.	See Table 9. II e) Ensure refrigeration unit is well maintained with adequate control.
	More Pressure	7. Isolation valve is closed in error whilst pump running. 8. Thermal expansion in the isolation valve section (fire).	Lines subject to full delivery pressure. Possible line fracture or flange leakage.	Covered by b). f) Perhaps worthwhile installing a pressure gauge upstream of the delivery pump. g) Provide thermal expansion relief in the valved section.
LESS OF	Less Flow	9. Flange leakage or valve stub blanked but leaking.	Decreased absorption. Higher operating cost in lost water.	Covered by a). c), and d)
	Less Temperature	10. Reaction gas temperature in oxidation unit lower.	Increased dissolved NOx concentrations in product acid.	See Table 9.1 I
OTHER	Maintenance	11. Equipment failure, flange leak, catalyst changeover in reactor. etc.	Process stops.	Ensure all pipes and fittings are constructed of the right materials and are stress relieved.

TABLE 9.8

HAZOP study: Gas-inlet stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
NONE	No Flow	I. Flow stopped upstream.	No absorption in column. Entire process stops as tail-gas Row stops.	a) Ensure liquid feeds to absorber and other process units shut down. b) Install LOW FLOW ALARM onto the FIC.
		2. Line blockage or the isolation valve shut in error.	As for I. Pressure buildup in pipe and secondary cooler.	Covered by a) and b). c) Install kick-back on upstream pumps and ensure pressure relief system is adequate.
		3. Line fracture.	As for I. Gases escape into the surroundings.	Covered by a) and b). d) Ensure regular patrolling of feed transfer lines. e) Plant emergency shutdown procedures.
MORE OF	More Flow	4. Increased feed	Possible reduction in absorption efficiency. May cause flooding.	f) Ratio control on the liquid feed streams should be sufficient. g) Install HIGH LEVEL ALARM on the FIC.
	More Pressure	5. Flooding.	Unit subject to high pressure, bursting discs may rupture, tail gas release.	Covered by c). h) Ensure correct sizing on pressure relief system.
		6. Isolation valve accidently closed 7. Thermal expansion in isolation valve section (fire).	As for 2. Line fracture or flange leakage.	Covered by b) and c). i) Provide for thermal expansion relief in the design of the isolation valve section.
	More Temperature	8. Insufficient cooling	Decreased absorption, higher pollution.	j) Ensure accurate temperature control on the internal cooling circuit.

TABLE 9.8 (Cont'd)

HAZOP study: Gas-inlet stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
LESS OF	Less Flow	9. Leaking inlet Range.	As for 3.	Covered by b) and d).
	Less Temperature	10. Overcooling.	Increased dissolved gases in acid (impurities).	Covered by j).
PART OF	High NOx Composition	11. Improved yield from reactor.	Higher tail-gas emission levels possible.	k) Manually increase make-up water flowrate.
OTHER	Maintenance	12. Catalyst change-over or upstream equipment failure . ..flange leakage etc.	Shutdown of entire process.	l) Ensure regular shutdown procedure instituted.

TABLE 9.9

HAZOP study: Gas-outlet stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
None	No Flow	1. No inlet gas flow. 2. Flooding in column. 3. PCV fails shut, line blockage or isolation valve closed in error. 4. Line fracture or flange failure.	No tail gas for expansion. Pressure build up in column and line As for I. As for 2. As for 1.	See Table 9.8. a) Install LOW LEVEL ALARM on PIC. b) Install pressure relief valve with bursting disc. Covered by b). c) Install HIGH LEVEL ALARM on PIC. Covered by a) d) Institute regular inspection of all transfer lines.
MORE OF	More Flow	5. Increased gas feed at inlet. 6. Decreased NOx absorption.	Transfer line subject to higher pressures. As for 5. Tail-gas emission levels up.	Covered by b) and c). Covered by b) and c). d) Look to altering make-up water feed rate in response.
	More Temperature	7. Higher feed gas or liquid inlet temperature.	Decreased absorption and higher NOx emissions.	e) Install HIGH LEVEL ALARM on TIC. See Tables 9.6 and 9.7.
	More Pressure	8. All of 5, 6, and/or 7. 9. Thermal expansion in isolation valve section (fire). 10. PCV fails shut or isolation valve shut in error.	As for 5, 6, and 7. Line fracture or flange leakage. As for 3.	Covered by b) and c). f) Install thermal expansion relief in isolation valve section. Covered by b) and c).
LESS OF	Less Flow	I I. Leaking flange or valve stub not blanked and leaking. 12. Flooding.	Less tail gas for expansion and release of NOx to atmosphere. As for 2.	Covered by a) and d). Covered by a) and b).
MORE THAN	Liquid Carryover	13. The entrainment device ineffective.	Condensation in down- stream lines (corrosion).	Replace entrainment device.
OTHER	Maintenance	14. Equipment failure, flange leak, catalyst changeover in reactor, etc.	Process stops.	Ensure all pipes and fittings are constructed of the right materials and are stress relieved.

TABLE 9.10

HAZOP study: Liquid-outlet stream

Key Word	Deviation	Possible Causes	Consequences	Action Required
NONE	No Flow	1. No liquid inlet from either make-up water or acid condensate.	See Tables 9.6 and 9.7	
		2. Flooding in column.	Increase in column pressure.	a) Covered by control and alarms specified in Tables 9.6, 9.7 and 9.8.
		3. LCV fails shut.	Liquid level in column increases adding to flooding problems.	Covered by a). b) Install HIGH LEVEL ALARM on LIC.
		4. Line fracture.	Discharge of acid into the surroundings. Loss of feed to the stripping column.	c) Regular patrolling and inspection of transfer lines. d) Install suitable alarms to stripping column to indicate loss of flow.
MORE OF	More Flow	5. LCV fails open.	Gas begins to bypass the plates causing higher NOx emissions.	e) Install LOW LEVEL ALARM on LIC.
	More Temperature	6. Higher inlet temperatures.	Less dissolved NOx in acid but higher NOx tail-gas emissions.	Covered in Tables 9.6, 9.7 and 9.8
	More Pressure	7. LCV fails shut or isolation valve closed in error.	Line subject to full surge or delivery pressure.	Covered by a).
LESS OF	Less Flow	8. Leaking flange or valve stub not blanked and leaking.	Loss to surroundings.	Covered by d) and e)
	Less Temperature	9. Lower inlet stream temperatures or over capacity from cooling circuit.	Higher concentrations of dissolved NOx in product acid.	See Tables 9.6, 9.7 and 9.8.
MORE THAN	NOx Dissolved	IO. Lower stream temperature.	Higher downstream operating costs. Same as 9.	See Tables 9.6, 9.7 and 9.8.
OTHER	Maintenance	I I Equipment failure, flange leak, catalyst changeover in reactor, etc.	Process stops.	Ensure all pipes and fittings are constructed of the right materials and are stress relieved

TABLE 9. I I

HAZOP study: Cooling-water circuit

Key Word	Deviation	Possible Causes	Consequences	Action Required
NONE	No Flow	1. Pump failure.	High emissions of NO_x in tail gas.	a) Install HIGH LEVEL ALARM on TIC on the tail-gas outlet line to indicate high emissions.
		2. Valve fails shut.	As for 1. Valve overheats.	Covered by a). b) Install kick-back on pumps.
		3. Line fracture.	AS for I and 2.	Covered by a) and b). c) Regular inspection and patrolling of cooling-water circuit lines and associated seals.
MORE OF	More Flow	4. Control valve fails open.	Product acid is at lower temperature, therefore higher dissolved NO_x.	d) Install a LOW LEVEL ALARM on TIC on the tail gas outlet line.
	More Temperature	5. Higher feed temperature to the refrigeration unit.	Possible higher NO_x emissions due to lower absorption.	Covered by a).
		6. Failure in the refrigeration unit.	As for 5.	e) Ensure refrigeration unit is well maintained with adequate control.
	More Pressure	7. Isolation valve is closed in error while pump running.	Lines subject to full delivery pressure.	Covered by b). f) Perhaps worthwhile installing a pressure gauge upstream of the delivery pump.
		8. Thermal expansion in the isolation valve section (fire).	Possible line fracture or flange leakage.	g) Provide thermal expansion relief in the valved section.
LESS OF	Less Flow	9. Flange leakage or valve stub blanked but leaking.	Decreased absorption. Low quality product and high emissions.	Covered by a), c), and d).
	Less Temperature	IO. Higher duty from refrigeration unit.	Increased dissolved NO_x concentrations in product acid.	Covered by a), c), and d)
OTHER	Maintenance	I. Equipment failure, flange leak, catalyst changeover in reactor, etc.	Process stops.	Ensure all pipes and fittings are constructed of the right materials and are stress relieved.

vapour rates (vapour velocities 4.0–4.5 m/s) before weepage problems are encountered. This 20% tolerance gives some operating flexability while maintaining a credible column size.

Notably it was the weepage limit that ultimately determined the choice of column diameter. It was necessary to provide gas velocities high enough to prevent significant weepage of liquid through the plate holes. Appendix G contains a comparison of the results for various column diameters, and also includes the weepage calculations.

The choice of column size is always a compromise between conflicting costs. The capital cost of larger units versus the cost of satisfying the appropriate emissions regulations. The desired acid production can be achieved using fewer plates if the partial pressure of nitrogen oxide components in the gas stream is increased. This may be achieved in two ways. Either increase the concentration by raising the ammonia to air feed ratio at the reaction stage, or increase the system pressure. The former approach also produces a higher nitrogen oxides emission level, while the alternative requires a greater stress tolerance on the shell structure. There will always be some assessment required of these factors if an optimum design is to be achieved.

The sieve-plate design is limited by the need for cooling coils. The tray design must ensure that the cooling coils sit easily in the layer of liquid. Therefore, the minimum weir height available was 40 mm. Despite this initial constraint a satisfactory tray design was developed.

It was decided to use dionized water in the cooling circuit, even though maximum heat transfer might be better achieved using other refrigerants. The advantages of other coolants are offset by the higher cost of the refrigerant and the risk of introducing a foreign liquor into the product acid. If a cooling coil fails inside the column, dionized water as the cooling medium will not poison the acid. Problems would occur due to large heats of dilution being released and possibly from product flooding, but the dilute acid produced would not be lost and could be recycled later to increase the acid strength.

The height of the column, the need for a large amount of piping in the cooling circuit, and the need to periodically adjust the cooling-coil flow valves, all suggest that it would be advantageous to install a parallel platform running alongside the column. This platform could

provide lateral support for the column and support of the column pipe work. It would also enable safe maintenance and regulation of the cooling circuit.

9.10 Assessment of the Design Method

The overall design obtained wasverysuccessful. With theavailability of comprehensive chemical equilibria data it was possible to develop an accurate mathematical model to aid the design calculations. The sieve tray and mechanical column design were both straightforward and appeared to be efficient.

The accuracy of this model appears to be very good. Comparison of the results with actual industrial column applications (see Ref. A2) suggests the model is indeed highly accurate. The limiting component in the model is the accuracy of the equilibria data, but as the majority of these empirical correlations are still used after more than 40 years their reliability should be almost guaranteed.

The limitations of the model are known, but they are certainly not easy to correct. First, there will be a temperature profile across each sieve plate. Although this profile may not represent a large variation it is appreciated that temperature is a sensitive operating parameter. The second limitation on the model is the assumption of 'plug flow' of the vapour through the void space between plates. This assumption may not be far wrong, but conditions close to the liquid surface would be better approximated by perfect mixing. The model adopts a simple approach to the problem, but it is particularly detailed in its attempts to represent actual conditions, The iterative sieve-tray design method suggested by Coulson and Richardson (Ref. A3) was also quite easy to implement. Whether the ease found in using this method was merely fortuitous is unknown, but certainly an effective design was obtained.

The shell and head design was in accordance with the design codes in Australian Standard: AS1 210. The assessment of expected pressures against the design value represented a simple and efficient method of determining the required metal thickness in each case.

9.11 Revised Absorption Column Costing

The costing for this unit has been revised from that presented in

Section 6.5 because of the detailed design information now available. The costing was done in accordance with the recommendations of Mulet, Corripio and Evans (Ref. Al 3). Their method utilizes correlations for the cost of a simple carbon-steel structure. Factors are then applied to account for the cost of other materials, the inclusion of trays, for operating pressure, and for incidentals such as ladders and railings. The cooling circuit was costed on the basis of heat-transfer area using similar correlations. Details of the cost estimation calculations are included in Appendix G.5.

The final cost estimate for this absorption column and associated structures was determined to be As870 000. This is significantly more than the first estimate (As250 000), however, the new cost reflects the need for a larger column than originally considered.

9.12 Conclusions

The absorption column design represents a compromise between mass transfer factors and economic considerations. The final design specification is for a column of 1.8 m diameter, approximately 32 m high, and containing 59 sieve trays.

Weak-acid condensate is fed to tray number 13, and make-up water is directed to the top tray (number 59).

Cooling coils are required to provide low temperatures for improved absorption. These coils must remove the considerable amount of reaction heat released in the absorption process. Herringbone-type coils are recommended for trays 14 to 59 only, with two coils (32 mm o.d. of 10 m effective length) required per tray.

The final judgement on the absorption column shows that it compares well with columns found in industry. The column at the KNC nitric acid facility in Kwinana is approximately 25 m high and satisfies about half the flowrate proposed for this application.

The absorption column and associated structures (platforms, ladders and railings) is expected to cost in the region of As870 000.

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CHAPTER 10
Case Study — Steam
Superheater Design

Summary

STEAM MAY be produced by cooling the nitrous process gases leaving the ammonia reactor/converter. These gases leave the reactor at 645°C and 1035 kPa, and at a flowrate of 31 706 kg/h.

The steam specification stipulates the need for superheated steam at 380°C and 4000 kPa. This medium-pressure product is of sufficient quality for the plant steam-turbine and ammonia superheater, with the remaining portion to be sold to another plant. A heat balance over the entire steam-production circuit concludes that this steam product may be produced at the rate of 5775 kg/h. This result determines the required heat duty for the steam superheater as 585 kW.

A shell-and-tube heat exchanger is recommended to produce superheated steam. This type of exchanger provides easy access to the heat exchange surface for platinum recovery (resulting from catalyst carry-over from the reactor). Thermal expansion at these operating temperatures requires the use of a floating head-type exchanger, with an internal clamp ring-type being preferred for reasons of shell clearance and high operating pressures.

The required heat-transfer area of 19.5 m^2 is based on an overall heat-transfer coefficient of 102 $W/(m^2 K)$. The best exchanger geometry for this application includes six internal baffles, one shell-side pass and two tube-side passes. The shell is fabricated from standard carbon steel piping of nominal pipe size 30, schedule number 80. The 112 tubes required are each 1.83 m long and 38.1 mm (1.5 in.) o.d. (BWG 12). The tubes must be fabricated from stainless steel type 250 for reasons of temperature tolerance.

The design calculations highlighted the shortcomings of the Kern method of exchanger design. The Kern method fails to account for shell-side inefficiencies such as bypassing, leakage, crossflow losses, and window losses. This leads to a marked overestimate of the shell-side heat-transfer coefficient and shell-side pressure drop. The Bell method is recommended to correct these deficiencies.

Details of the calculations associated with the steam superheater design are presented in Appendix H.

The steam superheater is costed at As25 000.

References Used: El to E11.

10.1   Introduction

Sufficient energy is released in this process, through the highly exothermic ammonia oxidation reaction, to provide enough high-level recoverable heat for two main areas. These are detailed below.

(a) Heat energy to tail gas for delivery to a gas expansion unit. This tail-gas expander provides the bulk of the shaft energy requirements for feed-air compression.

(b) Steam production.

It is the second of these two areas that is of interest in this chapter. Heat is delivered for steam production in many separate stages. Preheat is first supplied to the low-pressure deionized water prior to deaeration. Higher pressure product (HP boiler feedwater) is further preheated to around 100°C for supply to the waste-heat boiler. The waste-heat boiler is then able to vapourize the high-pressure deionized and deaerated water for final delivery to the steam superheater.

Medium-pressure steam is produced in this process to provide some plant heating/energy requirements. An economic advantage is the income derived from the export of excess steam. The steam specification corresponds to requirements for in-house application at the steam turbine, and also in response to the needs of the adjacent ammonia plant. The final steam specification is a superheated medium-pressure product at 380°C and 4000 kPa.

In order to determine the design constraints, it is necessary to

perform both mass and energy balances across the steam superheater. The balances for this unit are unchanged from those results presented in Section 7.4. The inlet temperatures and pressures for both streams and their flowrates are known. Given the required steam specification, the remaining unknowns can be calculated. The full calculation details are included in Appendix F.2.6. The important findings from those calculations are presented Table 10.1.

This chapter contains the specification for the steam superheater to achieve the required design. Aspects to be considered include the heat-transfer requirements from the unit, and the mechanical and construction aspects of the design. It is noted that the unit must conform to the appropriate design codes (Australian or equivalent).

10.2 Summary of Design Method

The design of the steam superheater has three stages, each with their own design approach requirements. The key to a good design will be the mutual satisfaction of all three aspects. The design calculation considerations are grouped according to:
(a) thermal design;
(b) mechanical design;
(c) materials of construction.

TABLE 10.1

Steam superheater duty parameters

Reaction Gas	
Flowrate in/out	*31 706* kg/h
Temperature in	645°C
Temperature out	*595°C*
Pressure in	1035 kPa
Saturated Steam	
Flowrate in/out	*5775* kg/h
Temperature in	250°C
Temperature out	380°C
Pressure specification	*4000* kPa
Total heat duty on the unit	2108 MJ/h (586 kW)

10.2.1 The Kern Method

Kern (Ref. El) is a well established, highly regarded, authority in the area of heat-exchanger design. These design philosophies are followed for the preliminary design work. The three aspects of design are all encompassed within the design method suggested by Kern. 'This method is based upon experimental work with standard tolerances and will give a reasonably satisfactory result in most cases' (Ref. E2, p. 545).

The Kern method requires a preliminary specification of the type of heat exchanger required including a suggested internal geometry. This preliminary sizing may be performed using a 'typical' heat-transfer coefficient for the two exchanger fluid media present. Having formulated the preliminary design an overall coefficient is then calculated by first evaluating a tube-side heat-transfer coefficient, and then a shell-side heat-transfer coefficient. Finally, fouling is considered so that the overall heat-transfer coefficient approaches a more realistic design value. The preliminary design is then adjusted and the calculations repeated.

The tube-side heat-transfer coefficient may be calculated in two ways. The first uses the Dittus-Boelter equation for heat-transfer in a turbulent environment. This equation is given below.

$$N_{Nu} = C\ N_{Re}^{0.8}\ N_{Pr}^{0.33}\ (\mu/\mu_{wall})^{0.14} \text{ —— EQUATION 10.1}$$

where

N_{Nu} = Nusselt number = $h_i D_e/k_f$;
C = constant = 0.021 for gases;
N_{Re} = Reynolds number;
N_{Pr} = Prandtl Number = $C_p\ \mu/k_f$;
k_f = thermal conductivity of the reaction gas = 0.062 W/(m K);
$h,$ = internal (tube-side) heat-transfer coefficient.

This correlation is widely accepted. All the variables relate to the fluid properties inside the tubes. As a check a second correlation recommended by Kern is used. It employs the heat-transfer factor (j_h) which is a function of Reynolds number. This equation is shown below.

$$h_i = (k_f/D_i)\ [j_h\ N_{Re}\ N_{Pr}^{0.33}\ (\mu/\mu_{wall})^{0.14}] \text{ —— EQUATION 10.2}$$

The variables shown are consistent with Equation 10.1. To obtain a conservative design, the lower value obtained from using these two equations will be used in further design calculations.

The shell-side heat-transfer coefficient in the Kern method is calculated using Equation 10.2. An equivalent diameter is calculated which is representative of the shell-side fluid passage geometry. Equation 10.2 therefore becomes:

$$h_o = (k_f/D_e) [j_h N_{Re} N_{Pr}^{0.33} (\mu/\mu_{wall})^{0.14}] \text{ — EQUATION 10.3}$$

In Equation 10.3 all variables relate to the properties of the shell-side fluid.

Finally, fouling factors 'typical' of the two fluid media are taken from Table 12 in Ref. EI (p. 645) and converted into their metric equivalents. All factors are then combined into the final equation for determination of the overall heat-transfer coefficient:

$$I/U_o = I/h_i + I/h_o + [[D_{to}\ln(D_{to}/D_{ti})]/(2k_s)] + F_i + F_o \text{ —}$$
EQUATION 10.4

where
U_o = overall heat-transfer coefficient;
h_i = internal (tube-side) heat-transfer coefficient;
h_o = outside (shell-side) heat-transfer coefficient;
D_{to} = outside diameter of tubes;
D_{ti} = inside diameter of tubes;
k_s = thermal conductivity of tube material;
F_i = tube-side fouling factor;
F_o = shell-side fouling factor.

The new overall coefficient, calculated from Equation 10.4, may be used for a revised design. The pressure drop over both the shell and tubes must also be calculated and will determine whether the design is satisfactory. It will also indicate the most appropriate changes to the geometry in the preliminary design.

The tube-side presure drop is calculated using the equation in Ref. E2 (p. 542).

$$AP, = N_p [8 j_f (L_t/D_{ti}) + 2.5] (\rho u_t u_t/2) \text{ — EQUATION 10.5}$$
where

AP, = pressure drop across tubes;
N_p = number of tube-side passes;
$_{if}$ = tube-side friction factor;
L_t = length of tubes;
D_{ti} = internal diameter of tubes;
ρ = density of fluid;
u_t = velocity of fluid through tubes.

The shell-side pressure drop is calculated using the equation in Ref.
E2 (p. 548).

$$AP, = 8\ j_f\ (D_{si}/D_e)\ (L_t/l_b)\ (\rho\ u_s\ u_s/2) \text{ — EQUATION 10.6}$$
where
AP, = pressure drop across shell;
if = shell-side friction factor;
D_{si} = internal diameter of tubes;
D_e = shell equivalent diameter;
L_t = length of tubes;
l_b = baffle length;
ρ = density of shell-side fluid;
u_s = velocity of fluid through shell.

Details of the associated calculations are presented in Appendices
H.4 to H.8.

10.2.2 The Bell Method

The Kern method is inaccurate for calculating the shell-side heat-
transfer coefficient and shell-side pressure drop (Ref. E2, p. 545).
Kern makes no account of the effect of bypss and leakage in the shell-
fluid passage. For these reasons the Bell Method (Ref. E9) is
employed to serve as a check on the preliminary design.
   The revised shell-side heat-transfer coefficient is given by the
equation shown below.

$$h_s = h_o\ F_n\ F_w\ F_b\ F_l \text{ — EQUATION 10.7}$$
where
h_s = new shell-side heat-transfer coefficient;

h_o = ideal heat-transfer coefficient given by Kern method;
F_n = tube-row correction factor;
F_w = window-effect correction factor;
F_b = bypass correction factor;
F_l = leakage correction factor.

Each correction factor is determined using a series of correlations and figures in Ref. E2 (pp. 553-569). The overall heat-transfer coefficient is recalculated from Equation 10.4.

 The shell-side pressure drop is also recalculated by adjusting for the above factors. The new shell-side pressure drop is given by the equation below.

$$AP, = (2\ AP,)\ + [AP,\ (N_b - 1\ )] + (N,\ AP,) \quad \text{— EQUATION 10.8}$$

where

$AP,$ = new shell-side pressure drop;

$AP,$ = pressure drop at exchanger ends;

$AP,$ = pressure drop in crossflow zone;

N_b = number of shell cross baffles;

$AP,$ = pressure drop in window zone.

Correlations in Ref. E2 (pp. 553-569) are again used to determine the component pressure drop values.

 Details of the associated calculations are presented in Appendices H.9 to H.ll.

10.2.3 Mechanical Sizing

The exchanger shell and tube sizings and exchanger tolerances are calculated according to the British design codes. These codes were preferred to the Australian codes because the main reference in the sizing of this excanger (Ref. E2) used the British Standards. Those British Standards used include:
BS 3274 (1960): Tubular Heat Exchangers for General Purposes.
BS 3606 (1978): Specification for Shell Tubes in Heat Exchangers.
BS 5500 (1985): Unfired Fusion Welded Pressure Vessels.
The required shell thicknesses are calculated according to the formula suggested in American Standard ASME VIII (see Appendix H).

Calculations relating to the sizing of the exchanger nozzle inlets and outlets are presented in Appendix H.12.

For the full design calculations and documentation refer to Appendix H, Sections H.2 to H.ll.

10.3 Design Selection Factors

Several factors must be considered in order to arrive at a preliminary exchanger design. These factors are considered in the order of the decision-making process.

10.3.1 Exchanger Type

A shell-and-tube type heat exchanger is recommended because it is a versatile exchanger often used in similar applications. The main factors prompting this decision are the large surface area provided in a small volume, a good shape for higher pressure operations, and the ease of cleaning. The final factor is that design and fabrication methods are well established. This enables the design specification to be as close to the optimum as is practical (without the aid of commercial computer-based design packages).

A floating head-type shell-and-tube heat exchanger is recommended for this application because of the need to provide capacity for thermal expansion of the tube bundle. The floating head also enables easy withdrawal of the tube bundle for cleaning purposes. This factor may be very advantageous, not because the streams are subjected to fouling, but because of the possibility that platinum carryover from the reactor will be deposited on the walls of the tubes. Platinum recovery is improved by providing easy access.

Finally, a clamp ring-type internal floating-head exchanger is selected. This split-flange design reduces the large clearances otherwise needed in other internal floating-head exchangers. External floating-head exchangers were rejected because these may only be used with shell pressures up to 20 bar (this application is 40 bar). Beyond 20 bar there is a danger of leakage through the head gland (Ref. E2).

10.3.2 Choice of Flow Mode

The reaction gas stream is specified on the tube-side for several
reasons.
 (a) It is the most corrosive of the two streams.
 (b) Any fouling tendancy will be greatest from this stream.
 (c) It is the higher temperature stream and therefore heat loss will be
minimized by flow through the tubes rather than the shell.
 (d) It has a much greater flowrate, and typically if the lowest flowrate
stream is allocated to the shell-side a better heat-transfer
performance is achieved.
The choice is not completely obvious because the higher pressure
fluid is usually better allocated to the tubes rather than the shell.
However, considering the considerable advantages to be gained
from the four areasabove (see (a) to (d)), the tube-side isconsidered
the best option for the nitrous process gas.

10.3.3 Materials Selection

The selection of appropriate materials of construction poses several
problems. First, the reaction gas stream is not corrosive. The nitrogen
oxide components are not themselves corrosive, and the
temperatures are high enough to prevent any chance of nitric acid
formation and condensation. The oxygen content in this gas stream is
inherently low since most of the oxygen in the feed air isconsumed in
the ammonia oxidation reaction. This might suggest that carbon steel
would be suitable for the exchanger tubes. However, temperature is
the limiting constraint for materials selection in this particular
application. Carbon steels heated for prolonged periods at
temperatures above 455°C may be subject to carbon segregation,
which is then transformed into graphite (Ref. E5, p. 23.5). When this
occurs the structural strength of the steel is reduced.
   Higher temperatures necessarily require chromium in the alloy
selected (Ref. E5, p. 23.59). An alloy of 9% to 12% chromium is
suitable for temperatures up to 730°C, and an alloy of 14% to 18%
chromium extends the service to 850°C. Selection of the higher
chromium content is not really necessary as the maximum
temperature envisaged is only 650°C. Therefore, the cheapest and

most practical material is a ferritic steel such as stainless steel type
250. This grade of stainless steel is readily available, being a standard
grade manufactured by BHP in Australia. It also satisfies the
Australian Standard AS1 548: Steel Plates for Boilers and Pressure
Vessels. The complete composition, and the mechanical and physical
properties of this steel are given in Ref. E7 (p. 29). This grade is
manufactured to satisfy the properties demanded in the Australian
Standard AS1 204.

The shell-side steam does not demand any special corrosion-
resistance properties from the material. The steam has already been
deaerated and deionized during pretreatment. The maximum
anticipated temperature is only 380°C. This permits the choice of
simple carbon steel for the exchanger shell.

10.3.4 Shell and Tube Sizing

The preliminary exchanger sizing was performed with several
requirements to be satisfied. First, larger tube sizes are preferred to
enable easier platinum recovery (platinum catalyst being deposited
from the reactor). The recommendations in Ref. E5 for tubes in boiler
applications suggest a tube size greater than 25 mm (1 in.) external
diameter. The second constraint is that the heat-exchanger area will
not be large due to the relatively small heat duty and the large mean-
temperature difference involved (see Appendix H.3). This suggests a
small tube length, within practical limits.

The preliminary tube spacing, baffle spacing, tube pitch, etc.,
conform to the recommendations made in Ref. E2 (Chapter 12). A
triangular tube pitch was chosen due to higher heat-transfer rates
than for the equivalent square pitch. Two tube-side passes are
recommended to maintain good gas velocities through the tubes,
although larger tube diameters will be required. One shell pass
should be adequate, providing internal cross baffles are used. Six
internal baffles are chosen in order to provide good hydraulic
conditions for heat-transfer and to provide lateral support for the tube
bundle.

The preliminary design was revised after determination of the
overall heat-transfer coefficient and the pressure drops involved.
These calculations are contained in Appendix H.

TUBE SIDE:		TUBES:
Fluid Handled	Process Gas	38.1 mm (1.5") O.D.
Flow Rate	31706 kg/h	32 mm (1.25") I.D.
Pressure	1035 kPa	Standard 2.8 mm thickness
Temperature	645 °C to 595 °C	112 tubes of 1.83 m (6') lengths
Head Material	SS250	Tube Material SS250
		Sealing strips each 5 vertical rows

SHELL SIDE:		SHELL
Fluid Handled	Saturated Steam	762 mm (30") O.D.
Flow Rate	5775 kg/h	700 mm I.D.
Pressure	4000 kPa	ie. nominal size 30,
Temperature	250 °C to 380 °C	schedule no. 60.
Shell Material	Carbon Steel	6 cross baffles
		25% vertical cut

TRIANGULAR TUBE PITCH

2.8 mm

O.D. 38.1 mm

47.5 mm

Figure 10.1
STEAM SUPERHEATER

By David JOHNSTON

FUNCTION.
To superheat saturated steam.
To cool nitrous gases exiting reactor.

OPERATION:
Continuous.

TYPE.
Internal floating head with clamp ring.
586 kW duty.
Exchange Area – 19.5 square metres.
1 shell side pass / 2 tube side passes
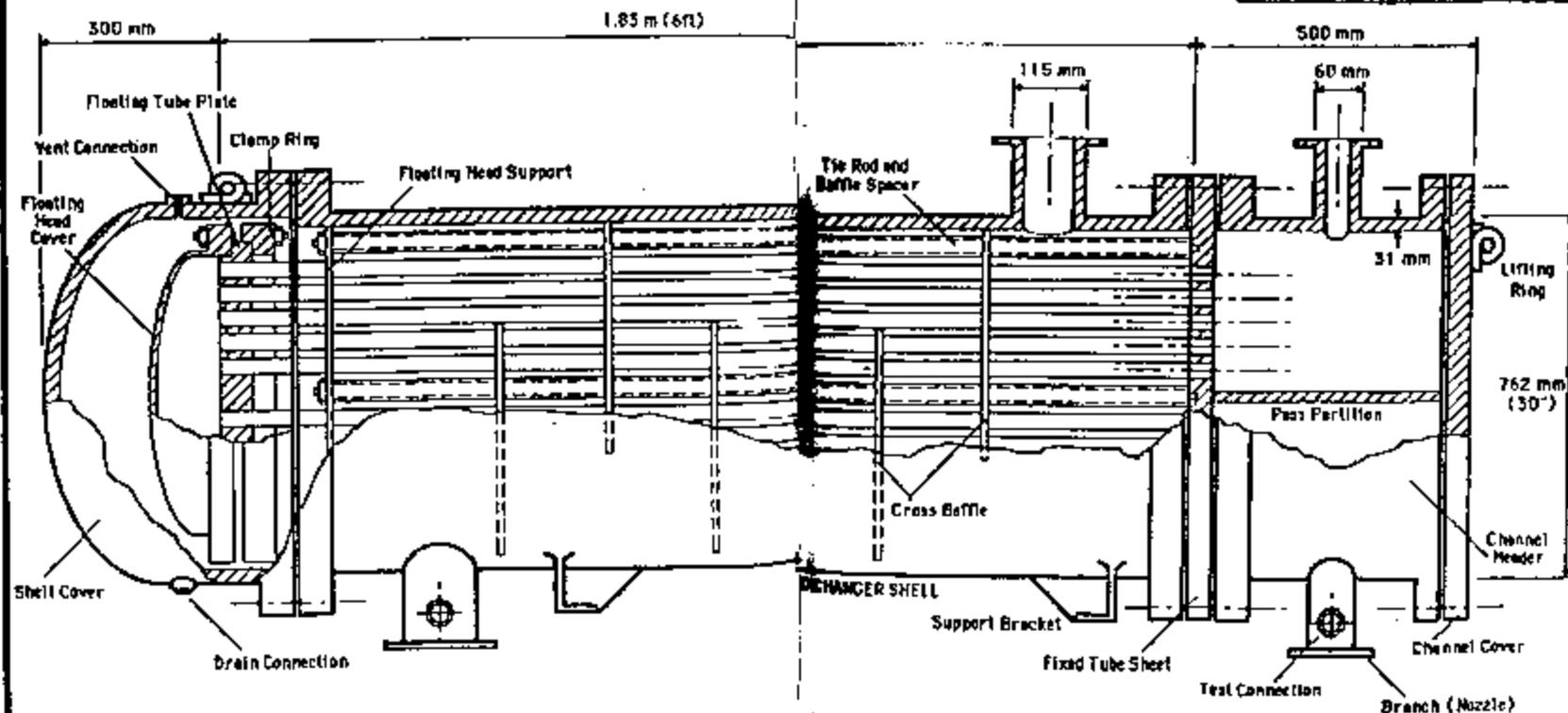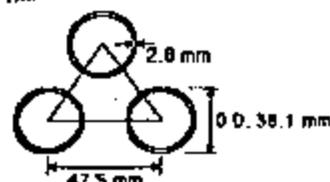
FIGURE 10.1   Steam superheater

TABLE 10.2

Specification for steam superheater design

Shell and Tube Heat Exchanger
  Clamp-ring (split-flange design) floating-head type
  1 shell pass, 2 tube passes
  Area of heat-transfer surface 19.5m²

Shell	-762 mm (30 in.)o.d.
	-700 mm i.d.
	-Nominal pipe size 30, schedule number 80
	--Shell thickness 3 1 mm
	—Constructed from carbon steel
	—6 internal cross-flow baffles
	—Baffle clearance tube to baffle 0.8 mm (1/32 in.)
	.. baffle to shell 5 mm (3/16 in.)
	-In/outlet branch, nominal pipe size 2, schedule number 80
Tubes	-112 tubes (SS304L); 38.1 mm (1.5 in.) o.d.; 32 mm i.d.; BWG 12
	-Tube thickness 2.8 mm; length 1.83 m (6 ft)
	-Triangular tube pitch 48 mm
	-Sealing strips required; 1 every 3 vertical tube rows
	-In/outlet branch nominal pipe size 4, schedule number 80

Tube-side heat-transfer coefficient	570 W/(m²K)
Shell-side heat-transfer coefficient	195 W/(m² K)
Tube-side fouling factor	0.002 (m² K)/W
Shell-side fouling factor	0.001 (m² K)/W
Overall heat-transfer coefficient	102 W/(m² K)
Tube-side pressure drop	26 kPa
Shell-side pressure drop	6 kPa

The engineering drawing for this unit is shown in Figure 10.1.

10.4 Design Specification

The full design specification for the steam superheater is given in Table 10.2, and a detailed drawing presented in Figure 10.1.

10.5 Process Control

The process control scheme suggested for the steam superheater is shown in Figure 10.2. This flowscheme features a control valve on the saturated-steam inlet line. A temperature controller taking its signal from the steam outlet line ensures the flow is regulated to maintain the superheater temperature of 380°C. Steam pressure is
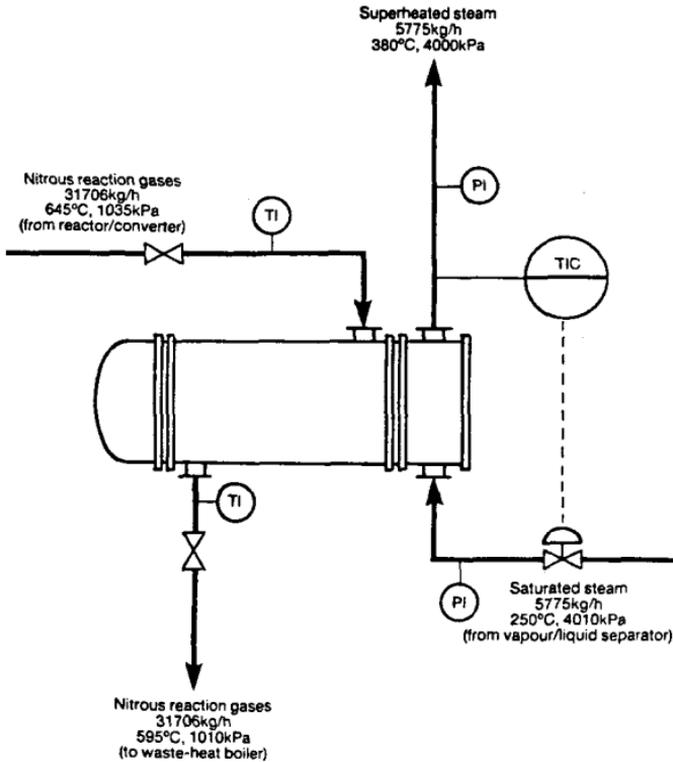
FIGURE 10.2 Steam superheater process-control scheme.

controlled prior to entry into the unit at the boiler-feed delivery stage. A pressure indicator on both inlet and outlet steam lines enables this parameter to be adequately monitored.

The nitrogen oxide reaction gas stream cannot be directly controlled from the steam superheater. Instead the flowrate, temperature and pressure are predetermined by the reactor feed conditions. No direct control exists on this stream as far as the production of steam is concerned, both inlet and outlet lines possess isolation valves for plant shutdown. These lines would be blanked before any platinum recovery work was attempted on the steam superheater. Inlet and outlet linesalsofeature temperature indicators, consistent with the policy of constant monitoring of this parameter throughout the process.

10.6 Design Method Evaluation

The most significant conclusion from the design calculations was the confirmation of suggested shortcomings in the Kern design method. The shell-side heat-transfer coefficient determined using the Kern method was 545 W/(m² K). However, upon consideration of loss factors, including window losses, bypassing, leakage and tube configuration losses, the revised shell-side heat-transfer coefficient by the Bell method was close to 195 W/(m² K). This difference caused the overall design heat-transfer coefficient to be reduced from 150 W/(m² K) to 102 W/(m² K).

The other parameter also needing revision was the shell-side pressure drop. This was reduced from 50 kPa (Kern method) to around 6 kPa. The Kern method is subject to overestimates of the pressure drop for the reasons already discussed.

If this 33% reduction in the overall heat-transfer coefficient and the 90% reduction in pressure drop had not been discovered, this would have caused a significant underestimate in the design to occur. This example should therefore serve to highlight the need for a more rigorous design approach to shell-side evaluation than that proposed by Kern. The Bell method proved quite adequate in this regard.

10.7 Revised Cost Estimation

The steam superheater costing, revised from that presented in Section 6.5, is included in Appendix H.13. This costing employs the correlations of Corripio, Chrien and Evans (Ref. E4). These correlations enable a preliminary figure to be determined for a standard carbon-steel exchanger based upon the heat-transfer area. This preliminary figure is then adjusted to allow for factors including choice of construction material, operating pressure, and exchanger type/configuration. The final cost determined from the revised costing of the steam superheater is As25 000.

10.8 Conclusions

The need for a clamp-ring, internal floating-head, shell and tube-type heat exchanger for use as a steam superheater is determined. The

unit considered is not large, being consistent with the small heat duty required. The main design features include 1 shell-side pass and 2 tube-side passes, with 6 internal cross-flow baffles. Total heat-transfer area required was found to be 19.5 m^2 based upon a calculated overall heat-transfer coefficient of 102 $W/(m^2 K)$.

The shell is constructed from carbon steel and will be fabricated from standard pipe of nominal size 30, schedule number 80. The 112 tubes required are 1.83 m (6ft) lengths and standard BWG 12. The tubes are made from stainless steel type 250 as recommended in the Australian Design Code AS1548: Design of Boilers and Pressure Vessels.

Some design problems became apparent when using the Kern method. The method resulted in overestimation of the shell-side heat-transfer coefficient and shell-side pressure drop. If the unit specification has proceeded based only on the Kern results the exchanger would have been unable to achieve the desired level of heat-transfer. The Bell method is recommended to avoid these problems.

The cost of the steam superheater is estimated to be As25 000.

References -Steam Superheater

El. D.Q. Kern, *Process Heat Transfer,* McGraw-Hill, New York (1950).

E2. J.M. Coulson and J.F. Richardson, *Chemical Engineering,* Volume 6, Chapter 12, Pergamon Press, UK (1983).

E3. W.M. Rohsenhow and J.P. Hartnett (Eds), *Handbook of Heat Transfer,* McGraw-Hill, New York (1973).

E4. A.B. Corripio; K.S. Chrien, and L.B. Evans, 'Estimate Costs of Heat Exchangers and Storage Tanks via Correlations', *Chemical Engineering,* 25 January, pp. 125-l 27 (1982).

E5. R.H. Perry and D. Green (Eds), *Perry's Chemical Engineers' Handbook,* 6th Edition, McGraw-Hill, New York (1984).

E6. G.F.C. Rogers and Y.R. Mayhew, *Thermodynamic and Transport Properties of Fluids,* 3rd Edition, Basil Blackwell, Oxford (1982).

E7. *BHP Company Limited,* 'Hot Rolled Products' (1984).

E8. *Firth-Vickers,* Publication 101, 'Rust, Acid and Heat-Resisting Steel', Staybrite Works, Sheffield (1966).

E9. K.J. Bell, 'Exchanger Design: Based on the Delaware Reserch Report', *Petroleum Chemistry,* Number 32, October, C26 (1960).

El 0. *Australian Standard* AS1 210: SAA Unfired Pressure Vessels Code (1982).

EI 1. Anon., 'Economic Indicators' *Chemical Engineering,* 16 April, p. 7 (1985).

CHAPTER 11
Case Study — Bleaching-Column Feed Pump Specification

Summary

A PUMP is required to deliver 'red' nitric acid product from the base of the NOx absorption column to the product bleaching column. It must transport 9.5 m³/h of this liquid product against a differential head of 450 kPa.

The pump specified to perform this task is a single-stage single-suction centrifugal pump. It employs a closed impeller, mechanical seals, and is constructed of stainless steel 304L. The small power load of approximately 3.5 kW is provided by an electric motor (2 pole) drive employing a horizontal shaft. The design casing pressure is sufficient to give a 30% safety factor over the total head required. Mechanical seals are preferred to a packed-box arrangement for their superior ability to contain the corrosive and toxic acid liquor.

Discussions were held with an industrial supplier of pumps in order to find a suitable pump for this application. An AJAX E32-20 model from AJAX pumps of Booragoon, Western Australia, was deemed suitable. The cost of this pump (from nomographs) is approximately AS3500.

Details of the pump calculations, specification, and manufacturer's literature are included in Appendix I.

References Used: P1 to P8.

11 .1 Introduction

This chapter considers the sizing, specification and selection of the pump required to deliver 'red' nitric acid from the NOx absorption column to the bleaching column. The final product acid is prepared in

the bleaching column by stripping out the dissolved nitrous gases with a secondary air stream. The requirements of the bleaching-column delivery pump are determined by the mass and energy balance for the absorption column (see Section 7.4 and Chapter 9). The relevent variables are presented below.

Pump Specification Data

'Red' Nitric Acid

Flowrate	11 800 kg/h
Temperature	65°C
Density at 65°C	1250 kg/m³
Viscosity at 65°C	1.9×10^{-3} Pa s
Partial pressure of nitric acid	16 kPa
Partial pressure of water	80 kPa

11.2 Design Method

In order to specify a pump to meet the required operating conditions it is necessary to calculate the forces opposing the pumping of this fluid, i.e. the pressure or head which makes pumping necessary. Calculations were performed for normal flow conditions and also for design conditions 20% above the normal. The pump specification is based on the larger design values.

The selection of the pump should be considered in conjunction with the design of the complete piping system. The total head required is the sum of the dynamic head due to friction losses in the piping, fittings, valves and process equipment, and any static head due to differences in elevation.

The dynamic losses can only be estimated by first sizing the pipe diameter of the line between the absorption column and the bleaching column. This is performed using recommended liquid velocities (Ref. P1 , p. 163) in conjunction with the known volumetric flowrate. The area calculated can be translated into a standard pipe diameter. The dynamic losses are then estimated by two methods. The first employs Genereaux's formula (Ref. P1 , p. 160);

$$\Delta P_{line} = 4.07 \times 10^{10} \, G^{1.84} \, \mu^{0.16} \, \rho^{-1} \, D^{-4.84}$$

where

ΔP_{line}	= dynamic pressure drop per metre of line;
G	= mass flowrate of flowing liquid;
μ	= fluid viscosity;
ρ	= fluid density;
D	= pipe diameter.

Alternatively, the friction loss can be estimated using the formula and friction factor (F_p) chart presented in Ref. P3 (p. 143). In this case the friction factor is a function of Reynolds number (N_{Re}).

Friction loss per metre = $4 \, F_p \rho \, u^2 / D$

where

F_p	= friction factor from chart (= f(N_{Re}));
ρ	= liquid density;
U	= fluid linear velocity;
D	= pipe diameter.

The friction loss in pipe fittings such as orifices, joints, elbows and isolation valves is calculated by converting these items into a so-called 'equivalent length'. This conversion uses standard loss values (Ref. P1, p. 164) and expresses them as an 'equivalent pipe length'. In this way the dynamic losses can be evaluated using the above equation for the expected dynamic pressure loss per metre. The pressure drop across a control valve is a function of the valve design. Sufficient pressure drop must be allowed when sizing the pump to ensure satisfactory operation of the control valve over the full range of flow required. The control valve should be sized for a maximum flowrate 30% above the normal stream flowrate (Ref. P1, p. 156).

Finally the total suction and discharge heads are evaluated by summing the various components. This allows determination of the major pump variables for specification, these are:

(a) differential head -simply the difference between discharge and suction heads;

(b) net positive suction head — the difference between suction pressure and the vapour pressure of the liquid;

(c) volume flowrate.

Having selected the type of pump, other factors such as optimum

pump speed and mechanical geometry are determined from standard practice recommendations (Refs. P2, P4, P5). The pump speed is evaluated using the dimensionless 'specific pump speed' number (N_s).

$$N_s = (NQ^{0.5})/(\Delta Hg)^{0.75}$$

where
N_s = specific pump speed;
N = actual cycle speed of pump;
Q = volumetric flowrate;
ΔH = differential head;
g = acceleration due to gravity.

Certain pumps will operate more efficiently in certain ranges of the specific speed, and an optimum pump speed range can be determined. The pump power is calculated using the formula (Ref. P2, p. 6.5):

$$P_p = \Delta HQ\rho/(3.67 \times 10^{\prime\prime})$$

where
P_p = pump power required;
ΔH = differential head;
Q = volumetric flowrate;
ρ = liquid density.

These calculations are checked by consulting manfacturers' literature concerning pumps (Refs. P7 and P8). Graphs presented in the literature enable a final sizing of the pump, including factors such as rotational speed, impeller diameter, power input and minimum net positive suction head.

The last consideration is the problem of cavitation. The net positive suction head (NPSH) determined in the calculations, must be sufficiently high to prevent spontaneous vapourization of liquid in the event of a drop in suction pressure.

Full details of the pump design calculations and the choices in pump specifications are included in Appendix I. A summary of these calculations supporting the final pump specification is presented in Figure 11 .1 and Table 11 .1 .
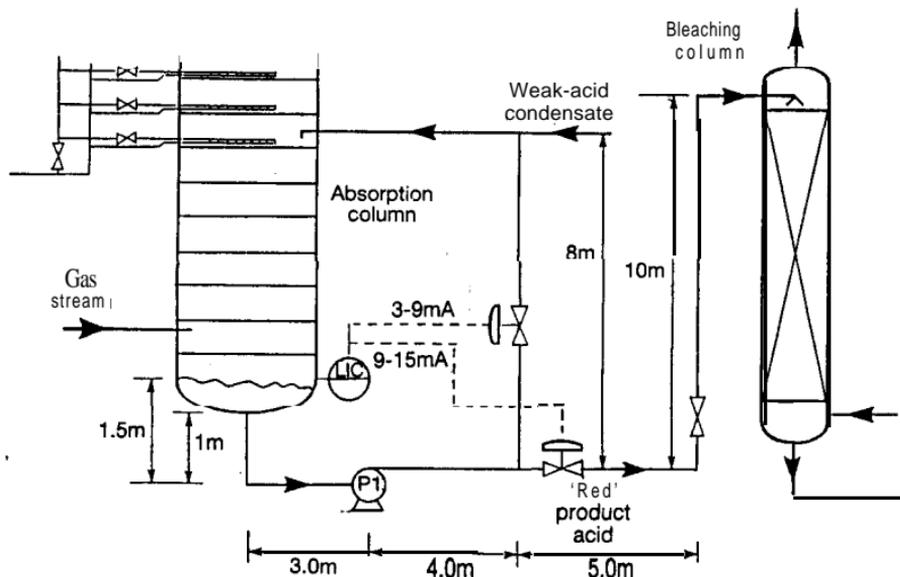
FIGURE 11.1 Flowsheet for pump specification.

11.3 Pump Specification

Following calculation and consideration of alternative pumps (see Appendix I), the final pump specification is presented in Table 11.2. For full details of the pump, its drive construction, and dimensions, refer to the sections of the AJAX manufacturers catalogue and details included in Appendix I.

11.4 Discussion

The demands of this hydraulic system are best satisfied by the use of a single-stage centrifugal pump. These pumps are widely used throughout the chemical processing industry and offer several key advantages over other types as listed below:

(a) simple construction, therefore low cost;
(b) non-pulsating flow (desirable for bleaching column feed);
(c) low maintenance costs (Ref. P2, p. 6.7);
(d) quiet operation.

TABLE 11.1

Summary of pump specification calculations

Pump and Line Calculation Sheet

Fluid	'Red' nitric acid
Temperature ('C)	65
Density (kg/m³)	1250
Viscosity (Pa s)	0.0019
Normal flow (kg/s)	3.28
Design maximum flow (kg/s)	4.12

Suction Calculations			*Discharge Calculations*		
Line size (mm)	40.0		Line Size (mm)	38.0	
Flow	Normal	Maximum	Flow	Normal	Maximum
Velocity (m/s)	2.3	3.4	Velocity (m/s)	2.3	3.4
Friction loss (kPa/m)	2.2	3.2	Friction loss (kPa/m)	2.2	3.2
Line length (m)	4.0	4.0	Line length (m)	22.1	26.2
Line loss (kPa)	8.8	12.8	Line loss (kPa)	50.0	84.0
Entrance loss (kPa)	15.0	15.0	Distributor orifice loss (kPa)	15.0	22.0
			Control valve (kPa)	140.0	200.0
Static head (kPa)	20.0		Static head (kPa)	132.0	
Equipment maximum pressure (kPa)	950.0	950.0	Equipment maximum pressure (kPa)	950.0	950.0
			Discharge pressure (kPa)	1287.0	1388.0
Suction pressure (kPa)	946.0	944.0	Suction pressure (kPa)	946.0	944.0
Vapour pressure (kPa)	96.0	96.0	Differential pressure (kPa)	341.0	445.0
Net positive suction head (kPa)	850.0	848.0	Control valve (% dynamic loss)	68.0	

TABLE 11.2

Bleaching-column feed pump specification

AJAX E32-20

Type	Centrifugal
Number of stages	1
Single/double suction	Single
Operating efficiency	40%
Vertical/horizontal mounting	Horizontal
Rotational speed	2900 rpm
Impeller type	Closed
Impeller diameter	195 mm
Construction materials	SS304L
Casing	Circular
— Design pressure	1500 kPa
— Design temperature	80°C
Drive	Electric, 440V, 50 c/s, 3 phase
Motor	2 pole, 4 kW
Seal type	Mechanical, external flush
Maximum flow	9.5 m³/h
Differential pressure-maximum	34 m (60% nitric acid)
-normal	26 m (60% nitric acid)
Power requirement	3.5 kW

this type of application, and achieve good overall efficiencies when working within their operational limits (Ref. P2, p. 6.7).

A horizontal shaft and mounting is preferred for this unit because the difficulties in lubricating vertical shafts make this an unnecessary option. The shaft seals are vital to prevent leakage of the highly corrosive and highly toxic nitric acid. Mechanical seals are preferred because they show no visible signs of leakage and are regarded as having superior performance compared to the older packing-type seals (Ref. P2, p. 6.9). Stainless steel 304L is designated as the material of construction required to safely handle 60% nitric acid.

The low power requirement is ideally suited to an electric motor drive, with the standard 440V (50 cycles per second), 3 phase connection. AJAX confirmed this choice, advocating a 2-pole motor of the above characteristics to meet the rotational speed of 2900 rpm. This is advantageous because motors of 4 poles or more greatly increase the unit cost.

A single-stage model was chosen because the low flowrates and relatively small differential head required do not warrant the added expense of a multiple-stage unit. Single suction is also a prerequisite

in that double-suction models are unable to handle corrosive liquids. A closed impeller is selected as these are cheap, offer good service in

11.5 Conclusions

The pump selected for the transport of 'red' nitric acid from the absorption column to the product bleaching column is a single-stage single-suction centrifugal-type pump. An AJAX E32-20 pump is recommended for this application. The application requires the pump to satisfy a maximum flowrate of 9.5 m³/h with a maximum differential head of 450 kPa.

This pump uses a closed impeller, mechanical seals, and is constructed of stainless steel 304L. The small power load of approximately 3.5 kW is supplied by an electric motor drive employing a horizontal shaft. The design casing pressure is sufficient to give a 30% safety factor above the total head required. Mechanical seals are preferred to a packed box arrangement due to their superior ability to contain the corrosive and toxic acid liquor.

The cost of this pump has been determined from industrial suppliers to be approximately AS3500, (this approximate price was also obtained from nomographs).

References — Pump Selection

P1. J.M. Coulson and J.F. Richardson, *Chemical Engineering,* Volume 6, Chapter 11, Pergamon Press, Oxford (1983).

P2. R.H. Perry and D. Green (Eds), *Perry's Chemical Engineers' Handbook,* 6th Edition, McGraw-Hill, New York (1984).

P3. J.A. Robertson and CT. Crowe, *Engineering Fluid Mechanics,* 2nd Edition, Houghton-Mifflin Company, Boston (1976).

P4. J.H. Doonlin, 'Select Pumps to Cut Energy Cost', *Chemical Engineering,,* 17 January, pp. 137-l 42 (1977).

P5. G.J. DeSantis, 'How to Select a Centrifugal Pump', *Chemical Engineering,* 22 November, pp. 163-l 68 (1976).

P6. Anon., 'Economic Indicators', *Chemical Engineering,* 17 April, p. 7 (1985).

P7. *AJAX PUMPS* Technical Data: 'E' Range Back-Pull-Out Centrifugal Pumps. Performance Curves: 'E' Range Back-Pull-Out Centrifugal Pumps.

CHAPTER 12
Case Study — Nitric Acid Storage-Tank Design

Summary

A TANK is required to store the one week production capacity from the nitric acid plant. This storage buffer allows the plant to continue operation for up to one week in the event of an unforeseen shutdown in the adjacent ammonium nitrate plant. The tank is a fixed cone-roof cylindrical-type design,

The tank is specified to have a capacity of 1950 m^3. This figure is the sum of one week production of nitric acid (1500 m^3) plus an extra 450 m^3. This extra 450 m^3 capacity will be the normal tank operating level and is available for product sales to external markets. This represents approximately 20 standard road-tanker loads. The dimensions of the tank represent a standard specification available through the Denver Company (United States). The tank internal diameter is 15.2 m and the tank height required is 10.7 m. The internal and external pressure loads require a wall, base and roof plate thickness of 16 mm (a standard plate thickness available through BHP, Australia) to meet the Australian design code for pressure vessels (AS1 210). This thickness gives a 100% safety factor over the maximum anticipated stresses.

Safety features for the tank include a pressure relief-valve system on the tank roof, to be opened when draining or adding to the tank contents. A bursting disc system also allows for pressure release in the event of pressure build-up. A manhole provides access to the tank for internal maintenance. The product inlet line is standard nominal pipe size 1.5, schedule number 40S. This inlet is sized for the maximum production flowrates. The product outlet line is of standard nominal pipe size 6, schedule number 120. This line is sized such that a standard-size road tanker may be filled in 15 minutes.

Details of the calculations associated with the tank design are presented in Appendix J.

The cost of this vessel is estimated (from a correlation) to be As390 000.

References Used: T1 to T5.

12.1   Introduction

A storage tank for product nitric acid is a necessity on this plant. The tank should have the capacity to store one week of full acid production to allow for continued supply in the event of unscheduled shutdowns in the adjacent ammonium nitrate plant. This requires a minimum tank capacity of 1500 m³. However, it is recommended to increase the tank capacity so that an inventory of 450 m³ of product acid is always available within the tank for outside sales. This extra volume is equivalent to 20 standard road-tanker loads. The tank must be constructed of stainless steel type 304L ('nitric acid grade'), the specification of this material is given in Appendix D. The design data required for this unit are specified below.

Nitric Acid Tank- Design Data

Design tank capacity	1950 m³
Design temperature	80°C
Design pressure	Pressure due to liquid head
Working pressure	120kPa
Nitric acid density	1250 kg/m³
Material of construction	SS304L
Design tensile strength	108 MPa
Joint efficiency	80%

The design tank capacity is estimated to be 1950 m³. This corresponds to a standard tank size as supplied by the Denver Company (United States). The normal tank operating level will be approximately 500 m³ to give the extra tank capacity as a reserve volume. This also ensures that a minimum of capital cost (in the form of product acid in the tank) is unused.

The design temperature represents the upper limit that acid may be

fed to the tank from the process. The working pressure represents the sum of atmospheric pressure and acid vapour pressure at the design temperature.

12.2 Design Method

The tank dimensions are determined according to standard tank geometries as used by the Denver Company (United States). Tank shell thickness is sized according to the limitations imposed by the Australian design standard for pressure vessels (AS1 210; Ref. T1). The tank contents are non-flammable but highly toxic and corrosive, appropriate safety features are recommended. Details of the calculations are given in Appendix J.

12.3 Tank Specification

The final tank specification (based on the calculations in Appendix J) is shown in Table 12.1. The tank is represented diagramatically in Figure 12.1.

A standard metal staircase and railing skirts the outer edge of the tank providing access to the tank roof. A manhole in the tank roof provides access for internal repairs. Discrete inlet and outlet lines are required to feed into the base of the tank. A pressure relief valve is

TABLE 12. I

Nitric acid storage-tank specification

Vertical cylindrical-type **tank**	
Fixed conical roof	
Total tank capacity	**1940** m^3
Normal operational capacity (maximum)	**1500** m^3
Tank inside diameter	15.2 m
Tank height	**10.7** m
Tank wall thickness	16 mm
Material of construction	SS304L
Inlet and outlet at tank base	
Inlet line: Nominal pipe size 1.5. schedule number **40S**	
Outlet line: Nominal pipe size 6, schedule number 120	
Manhole, bursting disc, pressure relief valve on roof	
No insulation	
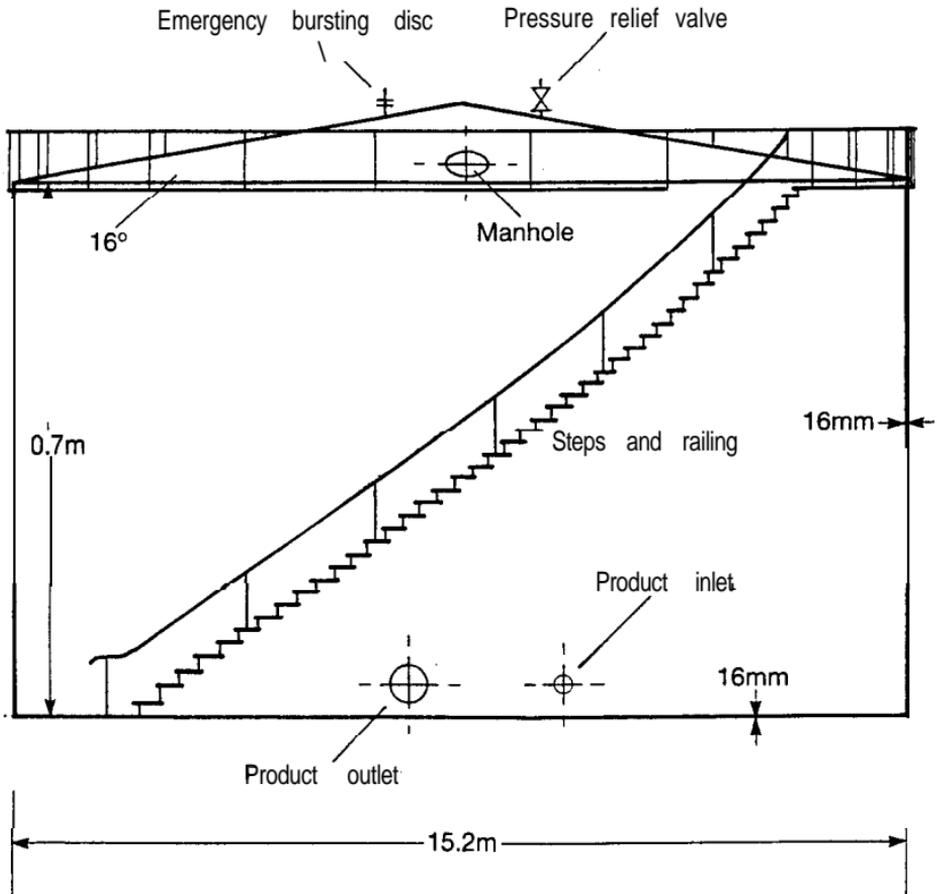
FIGURE 12.1   Nitric acid storage tank.

attached to the roof. This valve is opened automatically when pumping product to, or withdrawing product from, the tank. The valve is shut when pumping stops so that vapour losses from the tank are contained. A bursting disc on the roof also provides emergency pressure relief for an unforeseen pressure build-up within the tank.

12.4 Conclusions

The nitric acid storage tank required for the plant has a capacity of approximately 1950 m³. This represents approximately 2600 tonnes

of product acid. The tank will normally contain only about 700 tonnes to satisfy outside product sales. The remaining capacity is in reserve in case the adjacent ammonium nitrate plant shuts down and cannot accept nitric acid feed. This provides a one week production buffer for the plant. The final specification requires a tank 15.2 m diameter, height 10.7 m, and a plate thickness of 16 mm.

The cost of the tank has been calculated (Appendix J) from correlations (Ref. T4) to be approximately As390 000.

References -Tank Design

T1. *Australian Standard* AS1 210: SAA Unfired Pressure Vessels Code (1982).
T2. *BHP Company Limited,* 'Hot Rolled Products' (1984).
T3. J.M. Coulson and J.F. Richardson, *Chemical Engineering,* Volume 6, Chapter 11, Pergamon Press, Oxford (1983).
T4. R.H. Perry and D. Green (Eds), *Perry's Chemical Engineers' Handbook,* 6th Edition, McGraw-Hill, New York (1984).
T5. Anon., 'Economic Indicators', *Chemical Engineering,* 17 April, p. 7 (1985).

Final Comments

At this stage of the design project the student probably writes an appropriateconclusion and submits the report. In an industrial situation there are (at least) one thousand and one tasks still to do. The next stage would be a thorough and accurate costing of the entire project and submission of detailed quotations for all aspects of equipment, design, construction, etc. If the project is to proceed, company policy decisions will have to be made by senior management. The decisions made in the design stages are frequently modified as the project proceeds and the calculations are revised. The iterative design process continues. The start-up and commissioning stages are considered in the design process, as are procedures for plant shut-down and detailed maintenance programmes. Plant modifications should only be performed with reference to the final design study. In this respect plant design is a truly dynamic activity and a *final* solution is never obtained.

A student design project can only provide an approximation to industrial design work. However, it should be possible to consider the nature of the design process and various aspects realistically. Student design projects are restricted only by time. In industry the same constraint applies with the additional consideration of economics.

Action Review the design project, identify aspects that should he performed in greater detail if more time was available.

How does this academic exercise differ from an industrial design study?

APPENDICES

Mote References for Appendices A to F can be found in Section 2.5. References for Appendices G to J can be found at the end of Chapters 9 to 12 respectively.

Contents   of   Appendices

APPENDIX A
Data for Section 1.2

TABLE A. 1

Physical and chemical properties of nitric acid

Composition (wt% HNO_3)	Density (kg/m^3)	Freezing point $(°C)$	Roiling point $(°C)$	Specific heat (at 20°C) $(kJ/kg\ K)$	Partial pressure (Pa at 20°C) HNO,	H_2O
0.0	998.2	0	100.0	4.19		2333
10.0	1054.3	- 7	101.3	3.73		2266
20.0	1115.0	-17	103.4	3.39		2026
30.0	1180.0	-36	107.0	3.18		1760
40.0	1246.3	-30	112.0	3.01		1440
50.0	1310.0	-20	116.4	2.85	27	1053
60.0	1366.7	-22	120.4	2.64	120	653
70.0	1413.4	-41	121.6	2.43	387	347
80.0	1452.1	-39	116.6	2.22	1400	120
90.0	1482.6	-60	102.0	1.97	3600	27
100.0	1512.9	-42	86.0	1.76	6000	0

Maximum boiling point azeotrope occurs for 68.8 wt% HNO, at a boiling point of 122°C.
For 100% HNO,:

Heat energy of formation (at 25°C)	=	- 173.35 kJ/mole
Free energy of formation (at 25°C)	=	79.97 kJ/mole
Entropy (at 25°C)	=	155.71 kJ/mole
Heat of fusion	=	10.48 kJ/mole
Heat of vaporization (at 20°C)	=	39.48 kJ/mole

TABLE A.2

Side reactions for the oxidation of ammonia

Side reactions	Heat of reaction $(\Delta H_R$; J/mole)	
$4NH_3\ (g) + 3O_2\ (g) \rightleftharpoons 2N_2\ (g) + 6H_2O\ (g)$	- 3 16 832	Reaction 7
$2NH_3\ (g) + 2O_2\ (g) \rightleftharpoons N_2O\ (g) + 3H_2O\ (g)$	- 275 780	Reaction 8
$4NH_3\ (g) + 6NO\ (g) \rightleftharpoons 5N_2\ (g) + 6H_2O\ (g)$	- 451 296	Reaction 9
$2NO\ (g) \rightleftharpoons N_2\ (g) + O_2\ (g)$	- 90 309 -	Reaction 10
$4N0\ (g) + 3O_2\ (g) + 2H_2O\ (1) \rightleftharpoons 4HNO_3\ (aq)$	- 144 334	Reaction 11
$4NO_2\ (g) + O_2\ (g) + 2H_2O\ (1) \rightleftharpoons 4HNO_3\ (aq)$	- 87 226 -	Reaction 12
$2N_2O_4\ (g) + O_2\ (g) + 2H_2O\ (1) \rightleftharpoons 4HNO_3\ (aq)$	- 58 609	Reaction 13

Data from Reference PT 1 (p. 174).

APPENDIX B
Data for Section 2.3

B.I Kwinana Nitrogen Company Proprietary Limited

KWINANA NITROGEN Company Pty Ltd (KNC) is an organization now wholly owned by Wesfarmers Ltd and operated by CSBP & Farmers Ltd. KNC owns and operates three plants on their Kwinana site, strategically located between BP (Refinery) Kwinana and CSBP and Farmers Ltd's Kwinana complex. KNC's first plant takes either natural gas or 'off gas' containing a high percentage of hydrogen from the BP oil refinery, and processes it to produce sufficient ammonia for use in the nitric acid (plant 2) and ammonium nitrate (plant 3) production at KNC and for compound fertilizer production in CSBP's fertilizer complex.

Nitric acid is produced by KNC predominantly for use as a raw material in the production of ammonium nitrate (both explosive and fertilizer grades), less than 1,000 tonnes of nitric acid is sold to outside buyers each year.

The nitric acid plant was commissioned in 1968 and produces between 60,000 and 80,000 tonnes per year of a 58% product by weight, dependent upon the demand for ammonium nitrate. Current annual production is around 72,000 tonnes. Maximum production capacity is approximately 275 tonnes per day.

The plant employs a single-pressure catalytic process for the oxidation of ammonia to nitrous gases. A single converter/reactor is used containing nine layers of platinum/rhodium gauze catalyst and a platinum catchment gauze. This consists of a palladium mesh situated below the catalyst to arrest platinum losses to the downstream process. Nitrous gases are passed through a sieve tray-type absorption column. The column contains internal cooling coils

to remove the heat of reaction generated when the oxides are absorbed in water. The cooling coils use chemically treated demineralized water as the coolant to avoid corrosion problems (the type 304L stainless steel used for construction in much of the plant is highly susceptible to chloride attack).

The initial oxidation reaction is highly exothermic and it is an efficient process, being self sufficient in the steam requirements. Waste heat is recovered and the steam produced is used to drive the air compressor.

The authors wish to thank *CSBP and Farmers Ltd* for permission to reproduce this information.

B.2 Export Shipping Proposal -Summary

Currently Alcoa of Australia Limited imports almost 600,000 tonnes of 50% caustic soda solution, mainly from Europe and North America,, for its alumina production operations in Western Australia. This demand constitutes 20 deliveries per year of 30,000 tonnes each (3 ships involved). The ships unload their caustic soda at the Fremantle and Bunbury ports.

Many of these ships depart for foreign ports without a cargo. For example, some of the European ships are sometimes routed to Singapore where they pick up palm oil for the trip back to Europe. However, these ships nearly always leave Western Australia with empty holds. Therefore, shipping charges out of WA for a chemical cargo would be only a fraction of that charged if a ship had to be provided for exclusive use (see Table B.9).

The ships holds which contain the caustic are constructed of mild steel. Although mild steel is not the ideal material for resisting corrosion, the cost of using stainless steels (together with their chloride corrosion problems) prohibit their use. Unfortunately nitric acid strongly attacks mild steel at a rate which makes it totally unacceptable for use in storage tanks and holds. Therefore in order to use one of the caustic-carrying ships for transporting nitric acid, a corrosion-resistant material would need to be found. Fortunately poly vinyl chloride (PVC) is ideal for this application and such a coating would be both inexpensive and easy to apply.

TABLE B. 1

Australian annual production data for nitric acid (100% basis)

Period	Annual production (tonnes)
1976	147 559
1977	166 056
1978	152 610
1979	163 566
1980	179 654
1981	182 219
1982	179 858
1983	187 812
1984	192 342
1985	200 576
1986[a]	89 774

"First six months of 1986 only.
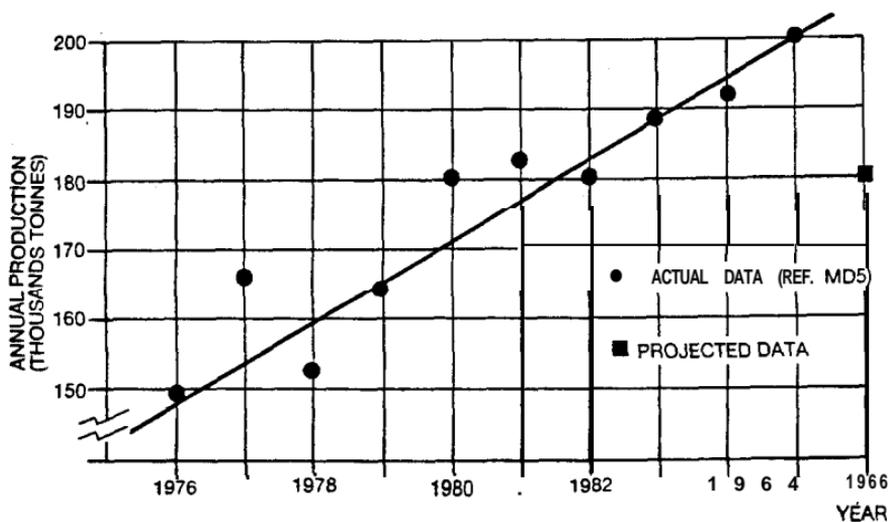
FIGURE B.1 Australian production trend for nitric acid.

APPENDIX B

TABLE B.2

Monthly Australian production of nitric acid (100% basis)

Year	Month	Production (tonnes)
1985		
	April	11468
	May	17440
	June	17 823
	July	15020
	August	15344
	September	18 931
	October	21 507
	November	15 872
	December	16 066
1986		
	January	19 003
	February	16 197
	March	15905
	April	16 954
	May	12 046
	June	9 667

Year	Production (tonnes)
1983-84	189 167
1984–85	196 843
1985-86	192 514

Data from Reference MD5 (p.8).

TABLE B.3

Australian export figures for nitric acid (100% basis):
Year ending June 1986 (Ref. MD7)

EXPORTS			
COMMODITY Country to... State of origin	Quantity (kg)	Value (A\$)	Unit Value (A\$/kg)
NITRIC ACID			
Fiji			
New South Wales	2163	1267	0.59
Indonesia			
Queensland	1480	5945	4.02
New Zealand			
New South Wales	102 240	26 860	0.26
Papua New Guinea			
New South Wales	326 830	146 610	0.45
Victoria	45	312	6.93
Queensland	57 330	24 570	0.43
Re-exports	21	408	19.43
Sub-total exports	384 226	171 900	0.45
Vanuatu			
Re-exports	350	877	2.51
Total to all countries:			
New South Wales	431 233	174 737	0.41
Victoria	45	312	6.93
Queensland	58 810	30 515	0.52
Re-exports	371	1285	3.46
Total Exports	490 459	206 849	0.42

TABLE B.4

Australian import figures for nitric acid (100% basis):
Year ending June 1986 (Ref MD7)

IMPORTS

COMMODITY Country from State of lodgement	Quantity (kg)	Value (A$)	Unit value (A$/kg)
NITRIC ACID			
Germany, Federal Republic			
New South Wales	313	2194	7.01
Victoria	798	6477	8.12
Australia	1111	8671	7.81
Japan			
Western Australia	8	103	13.80
Switzerland			
New South Wales	5	56	11.10
United Kingdom			
Victoria	755	8769	11.62
Western Australia	6720	5811	0.86
Australia	7475	14 580	1.95
United States of America			
Victoria	6469	13 035	2.02
Total from all countries:			
New South Wales	318	2250	7.08
Victoria	8083	29 622	3.66
Western Australia	6728	5915	0.88
Total Imports	15 129	37 787	2.50

TABLE B.5

Australian   producers   of   nitric   acid

Company	State	Product acid
ACF & Shirley Fertilizers Ltd	Queensland	58%
Adelaide and Walleroo Fertilizers Ltd	South Australia	60%
Consolidated Fertilizers Ltd		
(via Eastern Nitrogen Ltd)	New South Wales	57%
Cresco Fertilizers Ltd	Victoria	57%
Du Pont Australia Ltd	Queensland	65%
ICI Australia Ltd	Victoria	70%
Kwinana Nitrogen Company Pty Ltd	Western Australia	58%
State Authority	Victoria	95%

Total 1985/86 financial year production of 193 000 tonnes.
Data from References MD2 and MD4.

TABLE B.6

World   annual   production   data   for   nitric   acid   (100%   basis)

Period	Annual production (tonnes)
1976	26 800 000
1977	28 100 000
1978	30 000 000
1979	31 200 000
1980	31400 000
1981	32 750 000
1982	33 100 000
1983	33 600 000

(DATA   FROM   REF.   MD4)
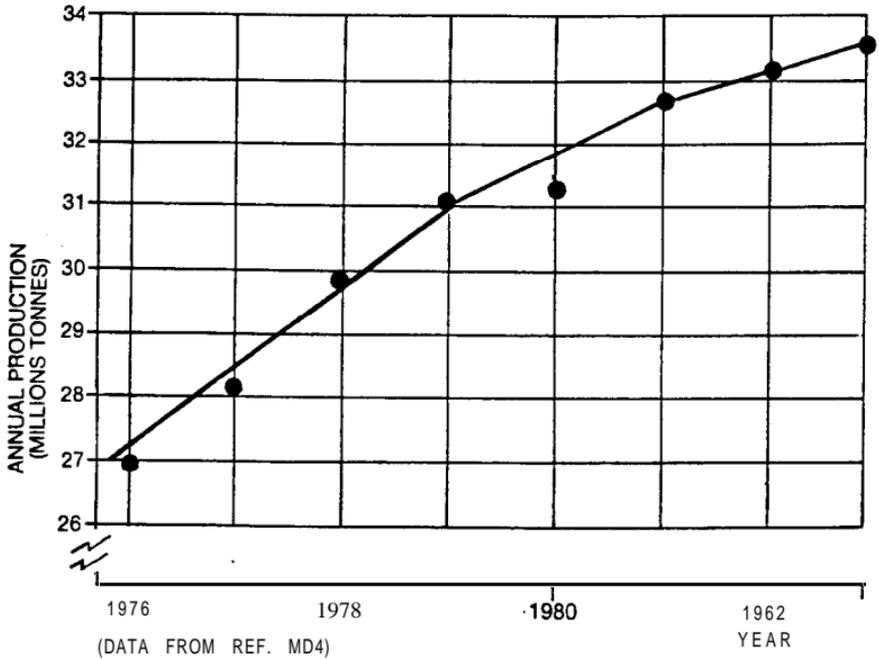
FIGURE B.2   World   production   trend   for   nitric   acid.
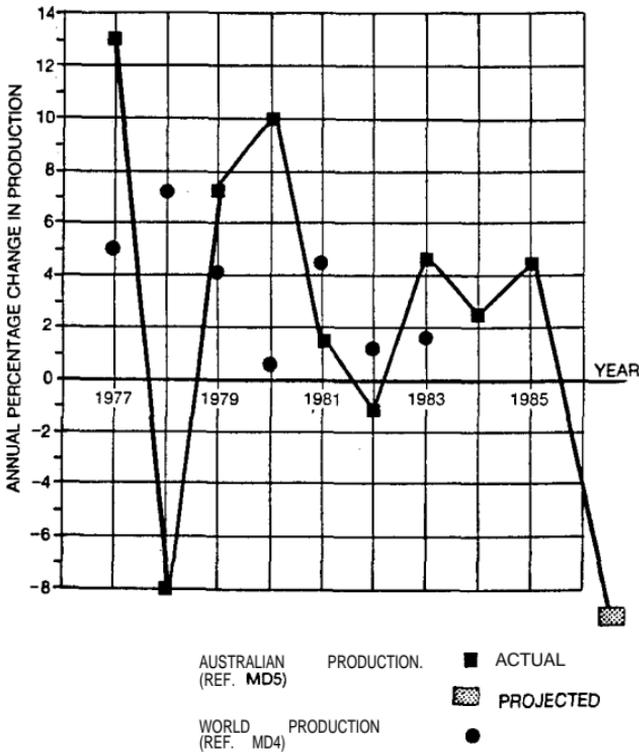
FIGURE B.3   Percentage change in production of nitric acid (100% basis),

TABLE B.7

Major importers of nitric acid in South-East Asia

country	Quantity imported annually [a] (tonnes)
China	15 000
Indonesia	3000
Malaysia	1000
New Zealand	600
Pakistan	500
Philippines	1.500
South Korea	6500
Singapore	2000
Thailand	2500
Total	32 600

[a]Data compiled from Reference MD4 (pp. 107.108) relating to 1980 figures.

TABLE B.8

Current prices for nitric acid and ammonia feed (world market)

	Grade	Foreign Price	A$/tonne
Nitric Acid	30°Be, 38°Be, 40°Be, 42°Be	196 US$/ton	339
	94% 98%	280 US$/ton	487
	100% basis	195–280 US$/ton	339487
	100% basis	62 000-83 000 yen/tonne	637-852
	100% basis	198287 f/tonne	449-648
Ammonia	Anhydrous	98–105 US$/ton	170–183
	Anhydrous	120 000–140 000 yen/tonne	1232-1438
	Anhydrous	59-66 f/tonne	134–150

Data from References MD8 and MD9.
Prices exclude shipping costs (see Table B.9).
Exchange rates used (Ref. MD10) as at 11 October 1986:
   1 A$ = 0.6332 US$
   1 A$ = 97.380 Japanese yen
   1 A$ = 0.44 17 British £ (pounds sterling)
Import/export prices are shown in Tables B.3 and B.4.

TABLE B.9

Shipping and port handling charges

IMPORT

	Cost	
Source	(US$/tonne)	(A$/tonne)
USA	200	316
Japan	207	327
U K	150	237

EXPORT

Destination	Cost[a] (A$/tonne)
Singapore	150

All port charges (import or export) are A$25/tonne.
Customs costs are A$60/container (1 container = 30 tonne).
Delivery charges from Fremantle Wharf are A$1 IO/container (Perth metropolitan area
only).
[a]Calculated at 11 October 1986.

APPENDIX C
Data for Section 3.2

C.I The Two Main Weak-Acid Production Processes — Summary

C.1 .1 The Single-Pressure Process

THE SINGLE-PRESSURE process shown in Figure 3.2 was developed to take full advantage of operating pressure in enabling equipment sizes to be reduced throughout the process. A single compression step is used to raise the pressure through the entire process sufficiently to favour absorption. Operating pressures range from 800 kPa as used by the Sumitomo Chemical Company Ltd (Ref. PT2; p.164), to 1100 kPa as used in the C&I Girdler single-pressure process (Ref. PT3; p.177).

Increased consumption of ammonia due to the higher pressure operation can be minimized by increased ammonia oxidation and complete ammonia/air mixing, and uniform flow distribution across the gauze. The higher oxidation temperature results in an increased consumption of platinum and rhodium and the need to rework the gauze every five to seven weeks (Ref. PT3; p.177). The higher temperature and the favourable pressure effect make possible a greater recovery of energy from the process.

The process begins with the vaporization of ammonia at 1240 kPa and 35°C using process heat. Steam is then used to superheat the ammonia to 180°C. Filtered air is compressed by an axial compressor to an interstage level and then, following cooling, by a centrifugal compressor to a discharge pressure of 1090 kPa. A portion of the air is diverted for acid bleaching, the remainder is circulated through a jacket surrounding the tail-gas preheater and then used for ammonia

oxidation. The heated air and the ammonia vapour (10.3% ammonia by volume) are then mixed and passed through the platinum/rhodium gauze, where the heat of reaction (producing nitric oxide) raises the temperature to between 927°C and 937°C.

The reaction gas flows through a series of heat exchangers, energy is recovered either as high-pressure superheated steam or as shaft horsepower from the expansion of hot tail gas. Approximately 70% of the oxidation to nitrogen dioxide occurs as the gas passes through the energy recovery train and is cooled to 185°C. After further cooling to 63°C in the primary cooler/condenser, separation of approximately one third acid product as 42% strength nitric acid is achieved. The remaining gas reaches a 43% oxidation conversion to nitrogen dioxide, with approximately 20% dimerization. The gas is combined with bleached air containing additional nitrogen peroxide, it then passes through an empty oxidation vessel and the secondary cooler. In cooling to 66°C, the gas provides heat to a recirculating hot water system used for vaporizing the ammonia. The gas entering the absorber is 95% oxidized to nitrogen peroxide, and 27% dimerized.

In the absorber deionized water is added to the top tray, and weak acid from the low-pressure condenser is added to a tray corresponding to its strength. Down-flowing acid and up-flowing acid alternately mix as the chemical steps of action formation and nitric oxide oxidation take place with the liberation of heat. There are three operational zones in the absorber, these are the lower zone cooled with plant cooling water, the middle zone cooled with chilled water, and the upper zone which is essentially adiabatic. High efficiency of heat removal in the middle and lower zones is particularly important due to its effect on the oxidation and dimerization reactions. For this design project, chilled water at 7°C is used and the tail-gas exit temperature is approximately 10°C. Acid from the bottom of the absorber is bleached at 1010 kPa with partially cooled compressed air. The bleach air, containing nitrogen peroxide stripped from the acid, is then added to the main gas stream before entering the oxidation vessel. The cold gas is warmed by heat exchange with the hot compressed bleached air, and then heated to the expander inlet

temperature of 620°C by two exchangers in the recovery train. The expander recovers 80% of the required compressor power. Expanded tail gas at 300°C flows through an economizer, providing heat to high-pressure boiler feed water and to low-pressure deionized deaerator make-up water. Subsequently tail gas is exhausted to the atmosphere at 66°C and less than 1000 ppm of nitrogen oxides.

The chilled water (7°C) for the absorber is supplied by an absorption refrigeration unit, using heat recovered from the compressor intercooler as the energy source. Heat for ammonia vaporization, as previously noted, is available at 35°C and is recovered from the secondary gas cooler. The system uses circulating condensate as the energy transfer medium.

C.1.2 The Dual-Pressure Process

The dual-pressure process shown in Figure 3.1 was developed to take advantage of two factors:
  a) low-pressure ammonia oxidation;
  b)   high-pressure absorption or acid formation.
In addition to the higher conversion, the lower catalyst gauze temperature (associated with the low-pressure ammonia oxidation) results in a much lower rate of platinum deterioration (Ref. PT3; p.175). Both advantages are maximized at the lowest pressure. In contrast, however, absorption is best performed at the highest pressure. The low-pressure ammonia oxidation is usually performed in the pressure range of 101.3 kPa (Du Pont: Ref. PTI 5; p.3) to 344 kPa (The Compaigne Francaisde L'Azote-COFAX: Ref. PT3; p.175). High-pressureabsorption is usually performed in theoperating range of 800 kPa (Du Pont: Ref. PT15; p.9) up to 1010 kPa (COFAX: Ref. PT3; p.175).

The process begins with the vaporization of ammonia at 550 kPa and 7°C, followed by superheating to 76°C using heat from the compressed bleached air. Filtered air is compressed in an axial compressor to 350 kPa and is mixed with the superheated ammonia vapour (1 O-l 1% ammonia by volume) prior to entering the converter/reactor.

In the converter, the gases react over the platinum/rhodium gauze catalyst. The gases leaving the gauze at 330 kPa and 865°C flow

through a series of heat exchangers for recovery of energy, either as high-pressure superheated steam or shaft horsepower from expansion of hot tail gas. Approximately 40% of the oxidation to nitrogen dioxide occurs in the gas as it passes through the energy recovery train and is cooled to 135°C (exit from the tail-gas preheater). After further cooling to 45°C in the medium-pressure condenser, and separation of 20% of the acid product as 30% strength nitric acid, the gas reaches 50% oxidation to nitrogen peroxide with approximately 20% dimerization.

The gas is combined with bleach air containing additional nitrogen peroxide and is compressed in a centrifugal nitrous-gas compressor to 1025 kPa. The exit temperature of 224°C is achieved due to the combined heat effects of the compression, the further oxidation to 80% nitrogen peroxide, and the virtual disappearance of the dimer. The compressed gas flows through an empty oxidation chamber, a high-pressure boiler feedwater economizer, and a low-pressure deionized water economizer, and thus is cooled to 95°C. Residence time in the system and the effect of increased pressure result in at least 95% oxidation to nitrogen peroxide, but the dimerization is low due to the temperature level. The gas is then cooled to the dew point (50°C) for entry into the absorber. The dimerization increases to 48%, adding significantly to the heat removed prior to the absorber. The system uses circulating condensate as the energy transfer medium.

The absorber is essentially the same as that described for the single-pressure process. Tower size and specific design details can be optimized to the particular process conditions and may vary

TABLE C. 1

Process comparison at the ammonia conversion stage

Parameter	Single-pressure process	Dual-pressure process
Pressure (kPa)	1056	336
Ammonia concentration (Vol.%)	10.3	10.0
Temperature of air/ammonia mixture (°C)	249	866
Temperature of nitrous gas leaving (°C)	937	865
Average conversion efficiency (%)	94.5	96.0
Number of converters	1	2
Gauze diameter (mm)	2032	4343
Mass gauze (kg)	62.2	62.2

considerably with the temperature of the chilled water. Chilled water at 15°C is used and the outlet gas temperature is 18°C. Refrigeration for the chilled water is provided by the ammonia vaporizer which operates at 7°C.

Weak acid from the bottom of the absorber is let down to 330 kPa for bleaching with air from the axial compressor. This air, with nitrogen peroxide stripped from the acid, flows to the suction of the nitrous-gas compressor together with the main nitrous gas stream from the condenser.

C.2 Discounted Cash Flow (DCF) Analysis of Plant Cost Differences — Summary

A discounted cash flow analysis is performed on those capital cost and production cost figures reported in Ref. PT1 (plant capacity of 280 tonne of 60% nitric acid). Although the plant cost data relates to a US plant, the figures still indicate the trend.

DCF Data and Conditions

USA plant capacity: 280 tonne/day of 60% nitric acid.

Single-pressure process, capital cost	As8 million
Dual-pressure process, capital cost	AS9.1 million
Difference in capital costs	AS1 .1 million

The difference in total production costs for identical capacity plants is reported as being AS2.15 per tonne, before taxes and depreciation, in favour of the dual-pressure process.

The initial principal represents the difference in capital costs. It will be invested at a 20% rate of interest; the interest will be subject to company tax of 49%. The value of this compounding sum is shown in Table C.7 under the column headed 'Principal'.

The operating profit after subtracting company tax will be invested as it becomes available at this same rate of interest. It is assumed that

operating profits also increase by the current annual 3% price rise assumed for nitric acid. The investment will accumulate each year and the value of this investment is shown in the column 'Accumulated profit' of Table C.7.

TABLE C.2

Comparison of the temperature limits

Parameter	Single-pressure process	Dual-pressure process
Upper-exit gauze ("C)	937	866
Lower-exit tail-gas preheater ("C)	184	135

TABLE C.3

Comparison of the high-level recoverable heat energy

Parameter	Single-pressure process	Dual-pressure process
Heat energy available:		
Sensible heat (kJ/h)	157 946	150 253
Heat of reaction (kJ/h)	22 115	13 327
Heat loss (kJ/h)	529	529
Total available (kJ/h)	179 532	163 051

TABLE C.4

Comparison of the compression power (actual shaft power)

Parameter	Single-pressure process	Dual-pressure process
Compression requirement (kJ/h)	63 360	61 524
Power from expander (kJ/h)	50 400	46 800
Steam turbine make-up (kJ/h)	12 960	14 724

TABLE C.5

Comparison of production costs

	Unit price (A$)	Unit quantity	Single-pressure process		Dual-pressure process	
			Quantity per tonne	Cost/tonne (A$)	Quantity per tonne	Cost/tonne (A$)
Raw Material						
Ammonia	191	tonne	0.2866	54.74	0.2821	53.15
Operating Costs						
Cooling water	0.020	cubic metres	112.6	2.350	101.6	2.152
Deionized water	0.500	cubic metres	1.063	0.555	1.160	0.349
Condensed steam	0.470	tonne	0.350	-0.165	0.608	− 0.286
Steam exported	13.90	tonne	0.460	− 6.360	− 0.342	− 4.747
Power	0.030	kWh	7.540	0.239	4.740	0.150
Direct labour (L)	12.60	per hour	0.053	0.669	0.053	0.669
Catalyst						
Platinum metal	12.84	gram	0.1880	2.414	0.0848	1.008
Rhodium metal	28.52	gram	0.0264	0.752	0.0064	0.181
Recovery, refining, etc.				1.282		0.320
Total Direct Costs (USA)				56.476		53.218
Indirect Costs						
Supervision (S) @ A$18.85 per hour			0.026	0.522	0.026	0.522
Payroll overhead @ 100 (L + S)				1.191		1.191
Maintenance, plant overheads, etc., @ 8% investment				5.151		5.824
Interest on catalyst inventory @ 15%				0.395		0.052
Total Indirect Costs (USA)				7.259		7.590
Total Production Costs (USA)				63.735		60.808
Total Indirect Costs (Europe)				7.170		7.337
Total Production Costs (Europe)				63.646		60.555

Data from Reference PT3 (p. 182) relating to 1979.

TABLE C.6

Inter-relationship of operating pressure, catalyst gauze
temperature and conversion efficiency

Operating pressure (atmospheres)	Gauze temperature (°C)	Conversion (%)
1	790–850	97-98
3.5	870	95-96
8	920	95-96
10.5	940	9495

TABLE C.7
Discounted cash flow (DCF) on production profits

Year	Principal (A$)	Operating profit (A$/year)	Accumulated profit (A$/year)
0	1 100000		
1	1 207 800	97 300	97 300
2	1 325 200	100 300	207 100
3	I 456 100	103 300	330 700
4	1 598 800	106 400	469 500
5	I 755 500	109 600	625 100
6	1 927 600	112800	799 200
7	2 116500	116200	993 700
8	2 323 900	119700	1210 800
9	2 551 600	123 300	1 452 800
10	2 801 700	127 000	1 722 200
II	3 076 200	130 800	2 021 800
12	3 377 700	134 700	2 354 700
13	3 708 700	138 800	2 724 200
14	4 072 200	143 000	3 134 100
15	4471200	147 200	3 588 500
16	4 909 400	151700	4 091 800
17	5 390 500	156 200	4 649 100
18	5918800	160 900	5 265 600
19	6 498 900	165 700	5 947 300
20	7 135 700	170 700	6 700 900
21	7 835 100	175 800	7 533 300
22	8 602 900	181 100	8 452 700
23	9 446 000	186 500	9 467 600

APPENDIX D
Data for Section 4.2

Summary of Properties of Stainless Steel Type 304L

Stainless steel type 304L is an extra-low-carbon stainless steel. Grade 304L can be used for applications involving service temperatures up to 450°C, and under certain conditions up to 800°C. The following data were obtained from Ref. G8 (p.B10).

Composition (%)

C	Si	Mn	Cr	Ni
0.03 max	0.2-l .0	0.5-2.0	18.0-l 9.5	10.0-l 1.5

Physical Properties

Similar to normal stainless steels.

Mechanical Properties

Normal heat-treated condition, rapid cooling from 1050°C:

0.2% Proof Stress (minimum)	20 kg/mm²
Ultimate tensile stress (minimum)	50 kg/mm*
Elongation, minimum (5.65,/A)	40%
Reduction of area	65%
Izod impact	11.1 m kg
Fatigue limit (10^7 reversals of stress, unbroken)	27 kg/mm²

Heat Treatment

This steel may be re-softened to remove the effect of cold work by rapid cooling from 1050°C.

Welding Properties,

Good, may be welded without subsequent heat treatment.

APPENDIX E
Data for Section 6.5

E.I Equipment Cost Calculations

E.I .1 Heat Exchangers

Sample calculation

1. Application -Ammonia vaporizer

   Heat exchange area = $83m^2$

   Details: 2 passes on tube side, carbon steel construction, fixed head.

   Employing the correlations presented in Ref. CE9 to estimate the purchase cost based upon heat-exchange area:

   Base cost (C,) = $\exp[8.202 + 0.01506 \ln(A) + 0.06811 (\ln(A))^2]$

   = US$14 700 (Base cost for carbon steel)

   Applying the various design indices:

   Design-type factor (F_D) = $\exp[-0.9003 + 0.09606 \ln(A)]$
   = 0.62

   Design pressure factor (F_P) = $1.2002 + 0.0714 \ln(A)$
   = 1.5

   Materials factor (F_M) = 1 .0

   Therefore, the US exchanger cost is approximately:

   Exchanger cost = $C_B F_D F_P F_M$
   = $14700 \times 0.62 \times 1.5 \times 1.0$
   = US$13 670

   These correlations are based upon 1982 United States figures and must be adjusted for both inflation and currency differences. The M & S equipment index asspecified in Ref. CE8; p.7, is used:

   Final exchanger cost = 13 670 $(786.0/745.6)(1/0.61)$
   = As23 625

This appears to be a reasonable figure based upon costing
experience obtained in industrial employment.

Other Applications

2. Application -Ammonia superheater
   Heat-exchange area = 30 m²
   Base cost = US$8450
   Design-type factor = 0.56
   Design pressure factor = 1.4
   Materials factor = 1 .0
   US cost = US$6900
   Final Australian cost = A$1 1 400

3. Application — Reactor/exchanger section
   Heat-exchange area = 72 m²
   Base cost = US$13 500
   Design-type factor = 0.61
   Design pressure factor = 1.5
   Materials factor = 1 .0
   us cost = US$1 2 500
   Final Australian cost = A$20 800

4. Application-Steam   superheater
   Heat-exchange area = 15 m²
   Base cost = US$6100
   Design-type factor = 0.52
   Design pressure factor = 1.4
   Materials factor = 1 .0
   us cost = US$4400
   Final Australian cost = A$7500

5. Application -Waste-heat boiler
   Heat-exchange area = 110 m²
   Base cost = US$17 600
   Design-type factor = 0.64
   Design pressure factor = 1.5
   Materials factor = 1 .0
   us cost = US$1 7 300
   Final Australian cost = A$29 000

6. Application -Tail-gas preheater
   Heat-exchange area = 89 m²
   Base cost = USS15 400
   Design-type factor = 0.62
   Design pressure factor = 1.52
   Materials factor = 1 .0
   US cost = USS14 600
   Final Australian cost = As24 300

7. Application — Cooler/condenser
   Heat-exchange area = 97 m²
   Base cost = USS16 200
   Design-type factor = 0.63
   Design pressure factor = 1.53
   Materials factor = 2.4 (for SS304L)
   US cost = USS37 600
   Final Australian cost = As62 500

8. Application — Secondary cooler
   Heat-exchange area = 140 m²
   Base cost = US$20 775
   Design-type factor = 0.65
   Design pressure factor = 1.55
   Materials factor = 2.4 (for SS304L)
   US cost = US$50 600
   Final Australian cost = As84 200

E.1.2 Absorption/Stripping   Columns

The costs of the two sieve-plate columns (absorption column and
product bleaching column) are estimated according to the
recommendations presented in Ref. CE9 (p.389). The columns are
costed according to their diameter, operating pressure, and the
number of sieve trays.
   The absorption column operates at about 10 bar and is estimated
(Ref. PT2) to be 2m in diameter with 50 trays. At US$1850/tray, the
cost isapproximately USS95 500. Including the cost of providing two
separate cooling circuits in the column, the cost increases to

US$1 10 000. This figure relates to 1983 costs; adjusting for inflation and currency differences, the final Australian purchase cost is approximately A$1 55 000.

The product bleaching column also operates at 10 bar, but is only 1 m in diameter, and has 10 trays. The final Australian purchase cost is estimated to be As25 000.

Absorption column = A$1 55 000
Product bleaching column = As 25 000

E.1.3 Compressor

The air compressor is costed according to the nomographs presented in Ref. CE9 (p.560 and 561). The nomographs match the compressor cost against either power rating or capacity.

The first stage of compression uses an axial compressor with steam turbine drive. It has a capacity of 511 m^3/min (36 000 kg/h). Using Ref. CE9 (Figure 13.47, p.560), the estimated cost is US$300 000. This figure relates to 1979, when updated to 1986 costs and adjusted for currency differences the final Australian purchase cost is As490 000.

The second stage of compression uses a centrifugal compressor with a gas expander drive. It has a duty of approximately 3400 brake horsepower. Using Ref. CE9 (Figure 13.50, p.561), the estimated cost is US$550 000 (1979). The final Australian cost is adjusted to be As951 000.

The compressor intercooler is costed as a heat exchanger as shown in Section E.1 .1. The final Australian cost is estimated to be As20 500.

1 st stage axial compressor	= As490 000
Compressor intercooler	= As 20 500
2nd stage centrifugal compressor	= As951 000

E.1.4 Reactor and Oxidation Unit

These two units are costed as pressure vessels using Ref. CE1 0. The reactor operates at 10 bar and approximately 940°C. It is constructed of mild steel and is refractory lined. The final Australian cost was estimated to be As46 000. Including the cost of the exchanger

section of the reactor, the total Australian purchase cost was As66 800

The oxidation unit operates at 10 bar and 140°C. It is constructed of SS304L. The final Australian purchase cost is estimated to be As34 000.

Total reactor cost = AS66 800
Oxidation unit   = As34 000

E.1.5 Refrigeration Unit

The refrigeration unit is costed according to the cooling duty of 500 kW. The cost is taken from Ref. CE9 (Figure B7, p.886) and is estimated to be USS33 000. This figure may be adjusted to give a final Australian purchase cost of As55 000.

 Refrigeration unit = As55 000

TABLE E. l

Estimation of extra fixed capital cost components

Assorted costs	Factor of total purchase cost
Piping	0.40
Instrumentation	0.20
Electrical	0.10
Ancillary buildings	0.16
Utilities	0.50
Site development	0.20
Total	1.56

TABLE E.2

Engineering   costs

cost	Factor of total physical plant cost
Design and engineering	0.25
Contractor's fees	0.05
Contingencies	0.10
Total	0.40

TABLE E.3

Working-capital cost estimation

Item	Quantity allowed	Cost (A$m)
Catalyst	One reactor charge	0.55
Ammonia feed	One weeks feed (340 tonnes)	0.10
Nitric acid product	One weeks product (2000 tonnes)	1.73
Maintenance stores	10% of annual maintenance charge	0.05
Wages and salaries	Two weeks	0.02
Total		2.45

TABLE E.4

Estimation of Australian production costs (per tonne of 100% acid produced)

Parameter	Unit price (A$)	Unit quantity	Quantity required	Cost (A$)
Direct Costs				
Raw Material:				
Ammonia	300	tonne	0.2866	85.98
Operating Costs:				
Cooling water	0.20	cubic metres	112.7	2.25
Deionized water	0.50	cubic metres	1.064	0.53
Condensed steam	0.50	tonne	− 0.350	-0.17
Steam exported	15.00	tonne	− 0.456	− 6.85
Power**				1.05
Direct labour (L)	12.00	hour	0.0529	0.64
Catalyst:				
Platinum	27.00	gram	0.1880	4.70
Rhodium	60.00	gram	0.0264	1.58
Recovery, refining, etc.				1.20
Sub-total Direct Costs:				89.15
Indirect Costs				
Supervision (S)	20.00	hour	0.026	0.52
Payroll overhead @ 100(L + S)				1.05
Maintenance, plant overheads, etc., @ 8% of investment				5.15
Interest on catalyst inventory @ 15%				1.20
Sub-total Indirect Costs:				7.92
Total Production Costs (Western Australia, 1986)				97.07

**Power requirements approximately 1300 kW/day.

SEC Industrial Tariffs —— Supply charge   21.91 ¢/day
First 165 kW/day   14.4 ¢
kW/day thereafter   11.5 ¢
Total power cost is approximately $7.50/hour or $1.05/tonne acid

TABLE E.5

Additional annual operating expenses

Item	Value (A$)
Insurance (at 1% of fixed capital)	135 000
Depreciation (at 10% of fixed capital)	1 350000
Maintenance (at 5% of fixed capital)	675 000
Administration costs	500 000
Sales and overseas marketing	1 000000
Laboratory (at 10% of direct labour)	35 000
Total	3 695000

TABLE E.6

Capital and interest repayment schedule

Year	Interest (A$)	Gross owing (A$)	Payment (A$)	Net owed (A$)
1	3 375 000	16875000	5 020 000	11 885000
2	2 963 750	14818750	5 020 000	9 798 750
3	2 449 688	12 248 438	5 020 000	7 228 438
4	1 807 109	9 035 547	5 020 000	4015 547
5	1003 887	5 019 434	5 020 000	—

Finance Arrangements
The project is financed by capital borrowed at 25% annual interest. The initial principal borrowed corresponds to the maximum calculated capital-cost estimation of A$13.5 million. Five equal repayments of A$5.02 million are required.

APPENDIX F
Calculations for Section 7.4

F.I Overall Mass Balance Sample Calculation

Specification: Plant producing 280 tonnes per day of 60% wt. nitric acid.

Basis: 100% acid and hourly production (7000 kg of 100% acid per hour).

Other data and process information
1. Conversion to nitrous oxides is 95% (remaining 5% to nitrogen gas).
2. Nitrogen in air is essentially inert in the process.
3. Converter feed is 10.3% ammonia by volume.
4. Ammonia feed: 0.2866 kg/kg 100% nitric acid (Ref. PT3).
5. Tail gas leaving the absorber at 10°C is saturated with water.
6. Air feed at 60% humidity (0.012 kg/kg dry air at 25°C).
7. All minor air components are insignificant to the calculations and are therefore considered as inert.

Initial Known Data,
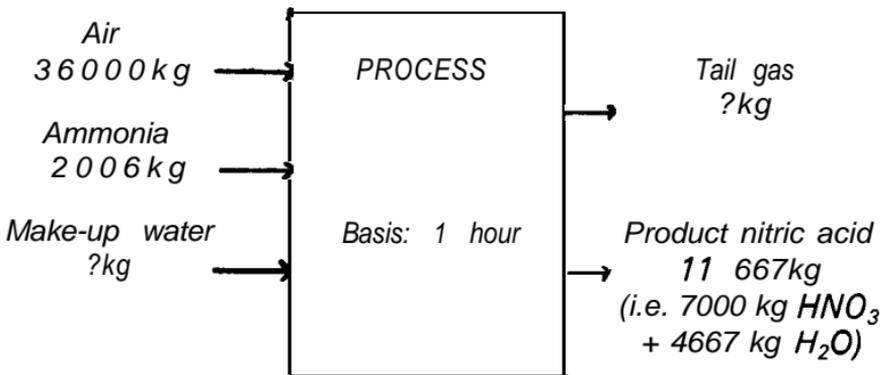
Method

Specification of the plant capacity and the above assumptions do not allow immediate solution of the overall mass balance. There is one more variable than there are known values/streams. Hence, an iterative approach must be used to solve the mass balance.

An initial assumption is made that the tail-gas stream is composed of nitrogen only. By an elemental nitrogen balance, the first estimate for the flowrate of the tail-gas stream is made. This enables the mass (and then moles) of hydrogen in the tail-gasstream to bedetermined. A mole balance on elemental hydrogen determines the moles of hydrogen added in the make-up water. This figure is used to determine the unknown feed stream of make-up water. Hence, all feed streams are defined.

With all inputs known, the tail-gas stream is re-determined and the calculation repeated until suitable accuracy is obtained. Two iterations proved quite sufficient. The final results are presented in Section 7.4, and a sample calculation is detailed as follows.

Step 1 *Determine nitrogen mole balance*
Inputs

Air	Nitrogen	1950
Ammonia	Nitrogen	117
Make-up water	Nitrogen	
Total	Nitrogen	2067

outputs

Product	Nitrogen	111
Tail gas	Nitrogen	?
	Nitrogen (as NO_x)	< 1
Total	Nitrogen	2057

Hence the moles of nitrogen gas in the tail gas = $(2067 - 1 - 111)/2$
$$= 977.5 \text{ kmol}$$

Step 2 *Calculate mass of tail gas (assume 3% oxygen)*
Mass of tail gas = mass of nitrogen in tail gas + inerts + 3% oxygen
$$= (978 \times 28) + 432 + [0.03(978 \times 28)]$$
$$= 28\ 638 \text{ kg}$$

Step 3 Calculate amount of water in the tail gas
Tail gas leaving the absorber is at 10°C and is saturated.
Absolute humidity = .0.011 kg water/kg bone dry air.
Therefore, mass of water in tail gas = humidity × mass of tail gas

$$= (0.011 \times 28\ 638)/(1 + 0.011)$$
$$= 312 \text{ kg}$$

Therefore, moles of water in tail gas = mass of water/molecular weight

$$= 312/18$$
$$= 17.3 \text{ kmol}$$

Step 4 Perform mole balance for hydrogen

Inputs

Air	Hydrogen (in water vapour)	44
Ammonia	Hydrogen	354
Make-up water	Hydrogen	?
Total	Hydrogen	664

outputs

Product	Hydrogen (in acid)	111
	Hydrogen (in water)	518
Tail gas	Hydrogen (in water vapour)	35
Total	Hydrogen	664

Moles of make-up water = $(664 - 354 - 44)/2$
$$= 133 \text{ kmol}$$
Mass of make-up water = 133×18
$$= 2394 \text{ kg}$$

Step 5 Re-calculate mass of tail gas from overall mass balance
Mass of tail gas = (36 000 + 2006 + 2394)   7000   4667
$$= 28\ 733 \text{ kg}$$

Return to step 3 until desired level of accuracy is achieved (only two iterations   required).

Final Solution:
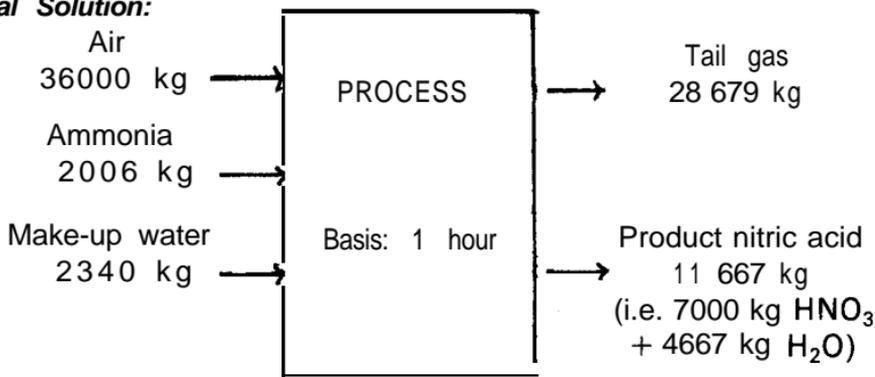
A full elemental mole balance can now be performed, based on this final solution. The final results are detailed in Section 7.4.

F.2 Unit Mass and Energy Balance Calculations

The process mass and energy balances are based on the stoichiometry claimed for the single-pressure process as described in Ref. PT1

All thermodynamic data for air, the reaction gas mixture, and feed ammonia are taken from Ref.TD1 (Section 3). Heat capacities for the various gas mixtures are calculated from the correlations in Ref.TD2. (Table E.1, p.538). Steam tables in Ref.TD3 are also used. Nitric acid properties are taken from Ref.TD4 (p.D-126 and D-77). Reaction equilibrium data are obtained from Refs. PT1 and PT2.

Requirement: 280 tonnes per day of 60% wt. nitric acid

Basis of calculations: 1 hour

The suggested feed ratios (Ref. PT1, p.182) are:
(a) 0.2866 kg ammonia per kg 100% acid;
(b) 5.15 kg air per kg 100% acid (82.5% reactor feed and 17.5% secondary air for product bleaching).

Make-up water (calculated from overall hydrogen balance):

$= (18/2)$ [Hydrogen in (product acid $+$ tail gas $-$ ammonia $-$ air)]

$= (18/2) [4667 (2/18) + 7000 (1/63) + 286 (2/18) - 354$
$- 428 (2/18)]$

$= 2338$ kg

Feeds

1. 2006 kg ammonia.

2. 36 000 kg air: 29 700 kg to reactor and 6300 kg to bleaching column.

3. 2338 kg make-up water to absorber.

Data:

Component	$\Delta H°_f$ (kJ/kmol)		C_P (kJ/kmol K)
$H_2O(v)$	-241	830	34.34
$H_2O(l)$	— 285	840	75.4
NO	90	370	29.75
NO_2	33	800	37.8
N_2O_4	26	300	41.65
HNO_3	— 206	570	110.0
O_2		—	29.8
N_2		—	29.16

Composition of air (mass fractions):

0.7491 nitrogen;	0.2270 oxygen;
0.0120 inerts;	0.0119 water.

Nomenclature used in calculations

P	Pressure	kPa
Q	Volumetric flowrate	m³/hour
F	Mass flowrate	kg/hour
T	Temperature	°C
C_P	Heat capacity	kJ/(kg K)
H	Enthalpy	kJ/kg
n	Moles	kmol
ρ	Density	kg/m³
Index	Describes ratio of reaction heat:sensible heat	—

F.2.1 Compressor

A two-stage compressor with intercooler is used to supply air for the process. Air requirements total 36 000 kg.

Stage one is low-pressure compression using an axial compressor. It takes air feed at approximately 35°C and compresses it to 180°C and 310 kPa.

Stage two of compression is a high-pressure process employing a centrifugal compressor. It takes air from the intercooler at 45°C and 300 kPa, and discharges it at 232°C and 1090 kPa.

The theoretical power for the total compression stage (with intercooler) is described by the formula (Ref. TD5; p.580):

$$\text{Theoretical power} = P_1 Q_1 \ln(P_2/P_1)$$
$$= P_1 F_{Air}/\rho \ln(P_2/P_1)$$
$$= 99(36000/1 .178)\ln(1090/99)$$
$$= 7\ 260\ 000\ kJ$$

Assuming the efficiency of the compressor is 65% (from Ref. TD5; p.581, Figure 14-I 1):

$$\text{Actual shaft power required} = \text{Theoretical power/Efficiency}$$
$$= 7\ 260\ 000/0.65$$
$$= 11\ 170000 kJ$$

A detailed design for the compressors would be required to determine the inlet and outlet velocities for the steam in the steam turbine, and the tail gas in the tail-gas expander. In this process, tail-gas expansion typically provides about 80% of the required energy with an efficiency in the production of shaft energy of less than 25%.

A detailed design for the compressors would be required to determine the inlet and outlet velocities for the steam in the steam turbine, and the tail gas in the tail-gas expander. In this process, tail-gas expansion typically provides about 80% of the required energy with an efficiency of expansion to shaft energy transfer of approximately 36%. The steam turbine provides the remaining 20% of the required energy with an efficiency in the production of shaft energy of less than   25%.

$$\text{Intercooler heat duty} = F_{air}\ C_P\ (T_{in} - T_{out})$$
$$= 36000 \times 1.05\ (180 - 45)$$
$$= 5103000 kJ$$

$$\text{Cooling water flowrate} = \text{Heat duty}/[C_P\ (T_{out} - T_{in})]$$

$$= 5\ 103\ 000/\ [4.2\ (40 - 20)]$$
$$= 60\ 750\ kg$$

F.2.2 Ammonia Vaporizer

The ammonia vaporizer receives liquid ammonia from the adjacent plant at $-15°C$ and 1240 kPa and vaporizes it at 35°C using warm water. The warm-water loop circulates water from the ammonia vaporizer to the secondary cooler. Water enters the ammonia vaporizer at 80°C and exits at about 50°C.

Enthalpy of ammonia at 1240 kPa: $-15°C$ -850 kJ/kg
                                  35°C 525 kJ/kg

Heat duty to vaporize ammonia $= (H_{out} - H_{in})\ F_{Amm}$
$$= [525 - (-850)]\ 2006$$
$$= 2\ 758\ 250\ kJ$$

Heat capacity for water $= 4.2$ kJ/(kg K).

Required circulation rate from the warm-water loop:

F_{ww} $=$ Heat duty$/[C_P\ (T_{in} - T_{out})]$
$$= 2\ 758\ 250/\ [4.2\ (80 - SO)]$$
$$= 21\ 900\ kg$$

F.2.3 Ammonia Superheater

The ammonia superheater takes the saturated ammonia vapour at 35°C and superheats it to 177°C for mixing with air downstream. Superheated steam at 380°C and 4000 kPa is the heating medium.

Average heat capacity for ammonia vapour in the range 35°C to 177°C is 2.25 kJ/(kg K).

Heat duty to superheat ammonia $= F_{Amm}\ C_P\ (T_{out} - T_{in})$
$$= 2006 \times 2.25\ (177 - 35)$$
$$= 640\ 920\ kJ$$

Enthalpy of superheated steam vapour at 4000 kPa: 380°C 3165 kJ/kg
                                            250°C 2800 kJ/kg

Heat of condensation at 250°C and 4 000 kPa: 1714 kJ/kg

Steam requirements $=$ Heat duty$/[\Delta H_{cond} + (H_{380} - H_{250})]$
$$= 640\ 920/[1740 + (3165 - 2800)]$$
$$= 305\ kg$$

F.2.4 Reactor Feed Mixer

Preheated air at 262°C and 1090 kPa is mixed with superheated ammonia at 177°C and 1240 kPa.

Reactor feed mixture

% Ammonia = $[2006/(2006 + 29\ 700)] \times 100$
          = *6.3%*
% Nitrogen = *$(0.7491 \times 29\ 700/31\ 706) \times 100$*
          = *70.2%*
% Oxygen   = *$(0.227 \times 29\ 700/31\ 706) \times 100$*
          = *21.3%*
% Inerts   = *$(0.012 \times 29\ 700/31\ 706) \times 100$*
          = *1.12%*
% Water   = *$(0.0119 \times 29\ 700/31\ 706) \times 100$*
          = *1.11%*

Heat   capacity: Ammonia at 177°C   2.38 kJ/(kg K)
                   Air at 262°C   1.05 kJ/(kg K)

Reactor feed temperature

Energy required to heat ammonia   =   Energy loss by air
$$F_{Amm}\ C_P (T_{out} - T_{in}) = F_{air}\ C_P (T_{in} - T_{out})$$
$$2006 \times 2.38\ (T_{out} - 177) = 29\ 700 \times 1.05\ (262 \quad T_{out})$$
$$T_{out} = 250\,°C$$

F.2.5   Reactor/Converter

The feed mixture oxidises over the platinum catalyst, thus converting the ammonia into nitrogen monoxide (NO) with a yield of 95%. The remaining 5% forms nitrogen gas and thus remains essentially inert to the reaction in all subsequent process units.
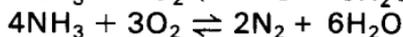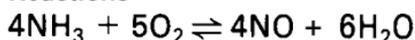
Reactions
$$4NH_3 + 5O_2 \rightleftharpoons 4NO + 6H_2O$$
$$4NH_3 + 3O_2 \rightleftharpoons 2N_2 + 6H_2O$$
Mass of nitrogen = Nitrogen in air + Nitrogen from reaction
              = *$[0.7491 \times 29700] + [0.05\ (2006/17)\ 14]$*
              = *22 331* kg
Mass of inerts = constant = 356 kg

Mass of nitrogen monoxide = 0.95 $(2006/17)$ 30

$$= 3363 \text{ kg}$$

Mass of water = Water in air + Water from reaction

$$= 353 + [0.95 (3 \times 2006/17) (18/2)]$$

$$= 3539 \text{ kg}$$

Mass of oxygen by difference $= 31706 - (22\,331 + 356 + 3363$

$$+ 3539)$$

$$= 2117 \text{ kg}$$

Reactor mixture

% Ammonia = 0%

% Nitrogen $= (22\,331/31\,706) \times 100$

$$= 70.4\%$$

% Oxygen $= (2117/31\,706) \times 100$

$$= 6.68\%$$

% Inerts $= (356/31\,706) \times 100$

$$= 1.12\%$$

% Water $= (3569/31\,706) \times 100$

$$= 11.2\%$$

% N 0 $= (3363/31\,706) \times 100$

$$= 10.6\%$$

Platinum-gauze temperature is 935°C.

Highly exothermic reaction yields approximately 14 000 kJ/kg ammonia.

Average heat capacity of the reaction mixture (1.22 kJ/kg K) is calculated by multiplying the pure gas values by their weight%.

Reaction mixture temperature $= T_{in} + [(AH_{,,,,}, F_{Amm})/(C_P F_{Tot})]$

$$= 250 + [(14\,000 \times 2006)/(1,22 \times 31\,706)]$$

$$= 976°C$$

The reaction mixture then enters the heat exchange section of the reactor providing energy to preheat the air feed and also providing the final preheat stage for the tail gas before it enters the expander. The design temperature for the reaction gas outlet is 645°C.

Total heat given out to the reactor internal exchangers is the sum of further reaction heat and also sensible heat in the reaction gases cooling from 972°C to 643°C. Reaction heat for this stage is reported in Ref. PT1 as 0.054 times the sensible heat. Average heat capacity for the reaction mixture in this temperature range is 1.27 kJ/(kg K).

Total heat transferred to exchangers = Index F_{Tot} C_P $(T_{out} - T_{,,})$
= *1.054 × 31 706 × 1.27 ×*
(976 − 645)
= 14047977 kJ

Heat to preheat air = F_{Air} C_P $(T_{out} - T_{in})$
= *29 700 × 1.05 (262 − 232)*
= *935 550* kJ

Heat to preheat tail gas = Total heat − Heat to preheat air
= *14 047 977 − 935 550*
= 13 112 427 kJ

The design temperature for the inlet tail gas to the expander is 620°C, hence the inlet temperature of the tail gas into the reactor can be calculated using an average heat capacity (in the range 620°C to *200°C)* of 1 .19 kJ/ (kg K).

Heat to preheat tail gas: 13 112 427 ≈ 1 .19 × 28 678 (620 − T_{in}) kJ
Therefore, inlet tail-gas temperature = 235°C.

F.2.6 Steam Superheater

Oxidation continues as the reaction gases cool, with the equilibrium increasingly favouring the formation of nitrogen dioxide and nitrogen tetroxide at the expense of nitrogen monoxide. The trend is initially gradual, but strengthens as the gases pass through the tail-gas preheater and cooler/condenser. During reaction in the superheater, approximately 5% of the nitrogen monoxide reacts with oxygen to form nitrogen dioxide. No nitrogen tetroxide is formed at this stage. The nitrogen, water, and inert gas components remain unchanged.

Reaction $2NO + O_2 \rightleftharpoons 2NO_2$
Mass of nitrogen monoxide = 0.95 × 3363
= *3195* kg
Mass of nitrogen dioxide = *0.05* (3363/30) *46*
= *258* kg

Mass of oxygen by difference = 31 706 − (22 331 + 356 + 3569
$$+ 3195 + 258)$$
$$= 1997 \text{ kg}$$

Superheater exit-gas mixture

Nitrogen	=	$(22\ 331/31\ 706) \times 100$
	=	70.4%
Oxygen	=	$(1997/31\ 706) \times 100$
	=	6.30%
I nerts	=	$(356/31\ 706) \times 100$
	=	1.12%
Water	=	$(3569/31\ 706) \times 100$
	=	11.2%
NO	=	$(3195/31\ 706) \times 100$
	=	10.1%
NO_2	=	$(258/31\ 706) \times 100$
	=	0.81%

The steam production section comprising the steam superheater and waste-heat boiler is designed to lower the reaction-gas temperature from 645°C to 280°C (further oxidation in these vessels will also continue to produce reaction heat). Steam of sufficient quality is to be produced for export to the adjacent ammonia and ammonium nitrate plants. Medium-pressure steam (at 380°C and 4000 kPa) is preferred.

The average heat capacity of the reaction gases in this temperature range is 1 .19 kJ/(kg K). Reaction heat is 0.143 times the sensible heat.

Energy for steam raising	=	Sensible heat + Reaction heat
	=	1 .143 (Sensible heat)
	=	Index F_{Tot} C_P (T_{out} − T_{in})
	=	$1.143 \times 31\ 706 \times 1.19 (645 − 280)$
	=	15740825 kJ

The high-pressure boiler-feed water is preheated to 96°C before entering the steam production circuit.

Steam raising capacity
= Energy required to change from liquid (96°C) to vapour (380°C)
= Heat available/$[(H_{380} − H_{250}) + \Delta H,,, + (C_P (250 − T_{in}))]$
= 15 740 825/$[(3165 − 2800) + 1714 + (4.2 (250 − 96))]$
= 5775 kg

Steam-superheater duty = $F_{steam} (H_{380} - H_{250})$
$$= 5775 \ (3165 - 2800)$$
$$= 2107875 kJ$$

Reaction-gas exit temperature $= T_{in} - [\text{Heat duty/Index } C_P$
$$F_{Tot})]$$
$$= 645 - [2\ 107\ 875/(1.143 \times$$
$$1.19 \times 31\ 706)]$$
$$= 596\,^{\circ}C$$

F.2.7 Waste-heat Boiler

The equilibrium continues to shift as the temperature of the reaction-gas mixture falls. Eventually a further 15% of the nitrogen oxide has reacted to form nitrogen dioxide, and 3% of this nitrogen dioxide reacts to form nitrogen tetroxide.

Reactions $2NO + O_2 \rightleftharpoons 2NO_2$
$$2NO_2 \rightleftharpoons N_2O_4$$
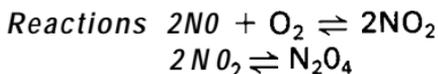
Mass of nitrogen monoxide = 0.85×3195
$$= 2716 \text{ kg}$$
Mass of nitrogen dioxide $= 0.97 \times [258 + (0.05 \ (3195/30)$
$$\times 46)]$$
$$= 9\ 6\ 3 \quad k\ g$$
Mass of nitrogen tetroxide $= 0.03 \ (963/46) \ (92/2)$
$$= 30 kg$$
Mass of oxygen by difference $= 31\ 706 - (22\ 331 + 356 + 3569$
$$+ 2716 + 965 + 30)$$
$$= 1739 kg$$

Waste-heat boiler exit-gas mixture
Nitrogen $= (22\ 331/31\ 706) \times 100$
$$= 70.4\%$$
Oxygen $= (1739/31\ 706) \times 100$
$$= 5.48\%$$
I nerts $= (356/31\ 706) \times 100$
$$= 1.12\%$$

Water	$= (3569/31\ 706) \times 100$
	$= 11.2\%$
NO	$= (2716/31\ 706) \times 100$
	$= 8.57\%$
NO_2	$= (963/31\ 706) \times 100$
	$= 3.04\%$
N_2O_4	$= (30/31\ 706) \times 100$
	$= 0.09\%$

Reaction gases from the steam superheater enter the waste-heat boiler at 595°C. The average heat capacity of the reaction gases in thistemperature range is I .I 9 kJ/(kg K). Reaction heat is 0.143 times the sensible heat.

Heat duty of the waste-heat boiler = Heat provided by reaction gases

$$= \text{Index } C_P\ F_{Tot}\ (T_{in} - T_{out})$$
$$= 1.143 \times 1.19 \times 31\ 706 \times (595 - 280)$$
$$= 13584548\text{kJ}$$

The circulation rate through the waste-heat boiler (WHB) and the temperature of the inlet water are unknown. Two equations describe the system. Equation F1 is the heat balance across the waste-heat boiler and Equation F2 is the net heat balance on the vapour/liquid separator.

$$\text{Heat duty}_{WHB} = [F_{WHB}\ \Delta H_{vap}] + [F_{WHB}\ C_P\ (T_{out} - T_{in})]$$
$$13\ 584\ 548 = [F_{WHB}\ 1714] + [F_{WHB}\ 4.2\ (250\quad T_{in})]...\text{Equation F1}$$

Net heat in from WHB + Heat from ammonia superheater condensate
= Heat to raise feed-water temperature
$$[F_{WHB} - 5470] \times [I\ 714 + 4.2\ (T_{,,} - 96)] + [305 \times 4.2\ (250 - T_{in.})]$$
$$= [5470 \times 4.2\ (T_{in} - 96)] \qquad\qquad ...\text{Equation F2}$$

These two equations are solved simultaneously giving a water flow through the waste-heat boiler of 6000 kg at an inlet temperature of 117°C.

F.2.8 Platinum Filter

The platinum filter is simply a gold/palladium alloy net that traps platinum from the reactor catalyst gauze as it is carried in the reaction gases. The filter works because the gold/palladium forms a complex with the platinum. Although there is no heat transfer in this unit the reaction does proceed, altering the composition of the reaction gases and raising their temperature. The small residence time in the filter only allows about 3% further oxidation of nitrogen monoxide to nitrogen dioxide. About 0.4% of the nitrogen dioxide subsequently oxidizes to nitrogen tetroxide.

Reactions $2NO + O_2 \rightleftharpoons 2NO_2$
              $2NO_2 \rightleftharpoons N_2O_4$
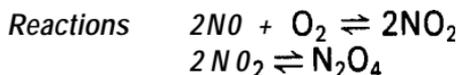
Mass of nitrogen monoxide $= 0.97 \times 2716$
                              $= 2634$ kg

Mass of nitrogen dioxide $= 0.996 [963 + (0.05 (2615/30) 46)]$
                              $= 1083$ kg

Mass of nitrogen tetroxide $= 30 + [0.004 (1083/46) (92/2)]$
                              $= 34$ kg

Mass of oxygen by difference $= 31\,706 - (22\,331 + 356 + 3569 + 2634 + 1083 + 34)$
                              $= 1699$ kg

Platinum filter exit-gas mixture

Nitrogen $= (22\,331/31\,706 \times 100$
            $= 70.4\%$

Oxygen $= (1699/31\,706) \times 100$
            $= 5.36\%$

Inerts $= (356/31\,706) \times 100$
            $= 1.12\%$

Water $= (3569/31\,706) \times 100$
            $= 11.2\%$

NO $= (2634/31\,706) \times 100$
            $= 8.30\%$

NO_2 $= (1083/31\,706) \times 100$
            $= 3.42\%$

N_2O_4 $= (34/31\,706) \times 100$
            $= 0.11\%$

The heat of reaction released is composed of the 15 820 kJ/kg nitrogen monoxide reacted and 6060 kJ/kg nitrogen dioxide reacted. Average heat capacity of the reaction gases in this temperature range is 1 .1 7 kJ/(kg K). The mass flowrate is unchanged.

Exit temperature of reaction gases
$= [(\text{mass } \Delta H_{react1}) + (\text{mass } \Delta H_{react\,2})]/(C_P F_{Tot})] + T_{in}$
$= [(82 \times 15\ 820) + (6060 \times 6)]/[1.17 \times 31\ 706] + 280$
$= 315°C$

F.2.9 Tail-gas Preheater

The tail-gas preheater is the second stage of preheat for the tail gas. It uses sensible heat and reaction heat from the process reaction gases to preheat the tail gas (ultimately for expansion).

The oxidation reaction also occurs inside the tail-gas preheater. The equilibrium shifts strongly in favour of the nitrogen dioxide and nitrogen tetroxide, such that 25% of the remaining nitrogen monoxide forms nitrogen dioxide, and 7% of the nitrogen dioxide forms nitrogen tetroxide.

Reactions $2\,N\,O + O_2 \rightleftharpoons 2NO_2$
                   $2\,N\,O_2 \rightleftharpoons N_2O_4$
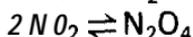
Mass of nitrogen monoxide	$= 0.75 \times 2634$
	$= 1975\text{kg}$
Mass of nitrogen dioxide	$= 0.93\ [963 + (0.05\ (2634/30) \times 46)]$
	$= 1083\text{kg}$
Mass of nitrogen tetroxide	$= 34 + [0.07\ (1947/46)\ (92/2)]$
	$= 110\ \text{kg}$
Mass of oxygen by difference	$= (31\ 706) - (22\ 331 + 356 + 3569 + 1975 + 1083 + 110)$
	$= 2282\ \text{kg}$

Platinum filter exit-gas mixture

Nitrogen	$= (22\ 331/31\ 706) \times 100$
	$= 70.4\%$
Oxygen	$= (2282/31\ 706) \times 100$
	$= 7.2\%$

Inerts $= (356/31\ 706) \times 100$
                $= 1.12\%$
Water $= (3569/31\ 706) \times 100$
                $= 11.2\%$
NO $= (1975/31\ 706) \times 100$
                $= 6.23\%$
NO_2 $= (1083/31\ 706) \times 100$
                $= 3.4\%$
N_2O_4 $= (11\ 0/31\ 706) \times 100$
                $= 0.35\%$

The mass flowrate of tail gas is known from the overall mass balance, but the tail-gas preheater balance can only be performed if the inlet temperature of the tail gas is known, This must be calculated from the specified design conditions for the absorber, and from the heat exchange at the primary tail-gas warmer. Tail gas leaves the absorber at 10°C (design specification) and it passes through the primary tail-gas warmer where it exchanges heat with the secondary air (for the product-acid bleaching column). The air enters at 232°C and leaves at 80°C. Assuming the tail gas has approximately the same heat capacity as air (a realistic assumption), then:

Stage 1 tail-gas warming:

$$\text{Temperature} = T_{in} + [(F_{air}/F_{Tgas})\ (T_{in} - T_{out})]$$
$$= 10 + [(6300/28\ 678)\ (232 - 80)]$$
$$= 43°C$$

The second exchange in the primary tail-gas warmer is with boiler-feed water. This water enters at 103°C and leaves at 96°C.

Stage 2 tail-gas warming:

$$\text{Temperature} = T_{in} + [(C_{PW}/C_{PTgas})\ (T_W/F_{Tgas})\ (T_{in} - T_{out})]$$
$$= 43 + [(4.2/1.02)\ (5470/28\ 678)\ (101 - 96)]$$
$$= 47°C$$

Therefore, tail gas enters the tail-gas preheater and is expected to increase its temperature to 235°C (from calculations for the reactor). The average heat capacity for the tail gas in the temperature range of 45°C to 235°C is 1.05 kJ/(kg K).

The average heat capacity for the reaction gases in the temperature range 315°C to 150°C is 1.17 kJ/kg K). The reaction heat is

approximately 0.140 times the sensible heat in this temperature range.

Reaction-gas exit temperature
$$= T_{in} - [(C_{PTgas}/C_{PRgas}) (F_{Tgas}/F_{Rgas}) (T_{out} - T_{in})/Index]$$
$$= 315 - [(1.02/1.17) (28\ 678/31\ 706) (235 - 47)/1.140]$$
$$= 185°C$$

F.2.10 Cooler/Condenser

The cooler/condenser further cools the reaction gases so that water condenses with some nitric acid. The design concentration for the nitric acid produced from the cooler/condenser is 42% by weight. It is assumed that all the water vapour is condensed. The other design specification for this unit is that of the gases leaving the cooler/condenser to the oxidation unit; 43% of the nitrogen monoxide has reacted to form nitrogen dioxide, and 20% of this gas has dimerized to nitrogen tetroxide.

It is important to note that since the cooler/condenser is the first of the vessels to be constructed of stainless steel, all subsequent cooling water must be taken from the deionized-water circuit.

Reactions
$$2NO + O_2 \rightleftharpoons 2NO_2$$
$$2NO_2 \rightleftharpoons N_2O_4$$
$$3NO_2 + H_2O \rightleftharpoons 2HNO_3 + NO$$
$$4NO + 3O_2 + 2H_2O \rightleftharpoons 4HNO_3$$
$$4NO_2 + O_2 + 2H_2O \rightleftharpoons 4HNO_3$$
$$2N_2O_4 + O_2 + 2H_2O \rightleftharpoons 4HNO_3$$
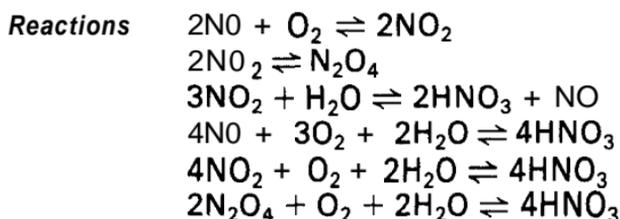
Calculations are performed in the same manner as for previous units. The reaction gas mixture is depleted to 26 176 kg as a result of the nitric acid removal.

Reaction-gas mixture exit from the cooler/condenser
Nitrogen $= (22\ 331\ /26\ 176) \times 100$
               $= 85.3\%$
Oxygen $= (399/26\ 176) \times 100$
               $= 1.52\%$

Inerts $= (356/26\ 176) \times 100$
                $= 1.36\%$
Water $= 0\%$
NO $= (696/26\ 176) \times 100$
                $= 2.66\%$
NO_2 $= (1240/26\ 176) \times 100$
                $= 4.74\%$
N_2O_4 $= (1154/26\ 176) \times 100$
                $= 4.41\%$

The weak acid produced by condensation of the reaction gases is approximately 42% wt. nitric acid, with 58% water. The 5531 kg of weak acid produced is sent to an appropriate tray in the absorption column.

The design exit temperature for the reaction gases leaving the cooler/condenser is 60°C. The average heat capacity of the reaction gases in the temperature range 185°C to 60°C is 1.13 kJ/(kg K). The heat of reaction/condensation/solution at this stage is approximately 0.442 times the sensible heat. For the heat balance calculations, the acid formed is assumed to be vapour until the exit from the cooler/condenser.

Cooler/condenser heat duty $= \text{Index } C_P F_{Tot} (T_{in} - T_{out})$
                              $= 1.442 \times 1.13 \times 31\ 706 \times (185 - 60)$
                              $= 6\ 458\ 000 \text{ kJ}$

Based on deionized cooling water being available at 20°C, and leaving the cooler/condenser at 40°C:
Cooling water flowrate $= \text{Heat duty}/[C_P (T_{out} - T_{in})]$
                              $= 6\ 458\ 000/[4.2 \times (40 - 20)]$
                              $= 77\ 000 \text{ kg}$

F.2.11 Oxidation Unit

The reaction gas mixture leaving the cooler/condenser enters the oxidation unit, here it mixes with the secondary air stream (this air stream has bleached the product acid and hence contains 460 kg

nitrogen tetroxide). The additional oxygen provided by this air stream enables rapid oxidation of most of the remaining nitrogen monoxide and some of the nitrogen dioxide formed.

Reactions $2 NO + O_2 \rightleftharpoons 2NO_2$
$2 NO_2 \rightleftharpoons N_2O_4$
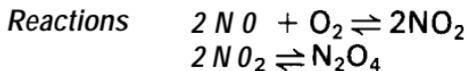
The calculations proceed as previously illustrated, the reaction gas mixture has increased from 26 176 kg to 32 936 kg with the addition of the secondary air stream.

Exit reaction gas from oxidation unit

Nitrogen   = *(27 050/32 936) × 100*
             = *82.1%*
Oxygen   = (1381132 936) × 100
             = *4.19%*
I nerts   = *(432/32 936) × 100*
             = *1.31%*
Water   = *(75/32 936) × 100*
             = *0.23%*
NO   = *(122/32 936) × 100*
             = *0.37%*
NO_2 = *(2203/32 936) × 100*
             = *6.69%*
N_2O_4 = *(1676/32 936) × 100*
             = *5.08%*

The oxidation reactions contribute a total of 2 965 100 kJ of energy according to the stoichiometry of the process. Assuming the temperature of the initial gas mixture is proportional to the mixture ratio, with 26 176 kg of reaction gas at 60°C and 6760 kg air at 50°C, then the equilibrium temperature prior to reaction is 58°C.

The temperature of reaction gas exiting the oxidation unit:

T_{out} = $T_{in} + \Delta H_{react}/(C_P F_{Tot})$
            = *58* + [2 965 1 00/(1.1*0* × 32 936)]
            = *140°C.*

F.2.12 Secondary Cooler

The final design composition for the nitrogen oxides in the reaction gases is 5% nitrogen monoxide, 95% nitrogen dioxide, with 27% dimerization to nitrogen tetroxide. This may be achieved by using the circulating warm-water loop to cool the reaction gases.

Reactions $2NO + O_2 \rightleftharpoons 2NO_2$
$2NO_2 \rightleftharpoons N_2O_4$
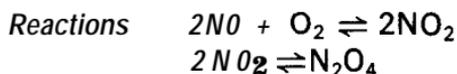
The composition of the exit reaction gas from the secondary cooler is calculated as before.

Secondary cooler exit-gas composition
Nitrogen   = *(27 050/32 936)* × *100*
             = *82.1%*
Oxygen   = $(1428/32\ 936)$ × 100
             = *4.33%*
Inerts   = *(432/32 936)* × *100*
             = *1.31%*
Water   = *(75/32 936)* × *100*
             = *0.23%*
NO   = *(102/32 936)* × *100*
             = *0.31%*
NO_2 = *(2164/32 936)* × *100*
             = *6.57%*
N_2O_4 = $(1685/32\ 936)$ × *100*
             = *5.12%*

The design temperatures for the warm-water loop are 50°C (at inlet) and 80°C (outlet). The flowrate at the ammonia vaporizer is calculated to be 21 900 kg.
The secondary-cooler outlet reaction-gas temperature
$= T_{in} - [(C_P F_{ww} (T_{out} - T_{in}))/(C_P F_{Tot})]$
= *140* − $[(4.2 \times 21\ 900\ (80 - 50))/(1.10 \times 32\ 936)]$
= *65°C.*

F.2.1 3 Absorber

All streams into the absorber are now known, these are:
 (a) reaction gas (of known temperature and composition);
 (b) weak acid (of known temperature and composition);
 (c) make-up water (at design temperature of 7°C).

The composition and temperature of the 'red' product acid stream is also known, based on 460 kg nitrogen tetroxide being absorbed into the acid.

Reactions $N_2O_4 \rightleftharpoons 2NO_2$ (initial dimerization)
             $2NO + O_2 \rightleftharpoons 2NO_2$
             $3NO_2 + H_2O \rightleftharpoons 2HNO_3 + NO$

Combined reaction $4NO + 2H_2O + 3O_2 \rightleftharpoons 4HNO_3$
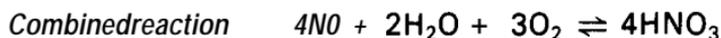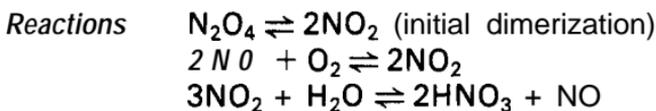
The composition of the tail-gas stream from the absorber can now be calculated. Calculations are based on an overall mass balance, and also on a mole basis for each element. The assumption made is that the tail gas is saturated with water upon leaving the column. Calculations were performed using an electronic programmable calaculator.

Tail-gas composition
Nitrogen $= (27\ 050/28\ 677) \times 100$
         $= 94.3\%$
Oxygen $= (850/28\ 677) \times 100$
         $= 2.96\%$
Inerts $= (432/28\ 677) \times 100$
         $= 1.51\%$
Water $= (286/28\ 677) \times 100$
         $= 1.00\%$
NO $= 0\%$
NO_2 $= (32/28\ 677) \times 100$
         $= 0.11\%$
N_2O_4 $= (27/28\ 677) \times 100$
         $= 0.9\%$

The tail-gas composition confirms the design specification for nitrogen oxide emissions of less than 1000 ppm. The emission is approximately 960 ppm. Western Australian emissions standards specify that emission of nitrogen oxides must not exceed 0.2 g per cubic metre of gas (STP). The emission from the absorber of 960 ppm is less than 0.1 g per cubic metre of gas (STP).

The design specification for the absorber requires that the tail gas leaves the absorber at 1 0°C, make-up water to be supplied at 7°C to the top plate, and inclusion of two independent sets of cooling coils in the column. The first set of coils runs in the top half of the column. The inlet temperature for these coils is 7°C and the design exit temperature is 20°C. The second set of coils occupy the bottom half of the column using cooling water from the normal cooling-water circuit. Therefore, the inlet temperature is 20°C, and the outlet temperature is 40°C.

The required flowrate through each set of cooling coils can be calculated if the heat duty is known. The assumption that the top set of coils occupies the top half of the column, and the bottom coils in the bottom half of the column, can only be tested after further detailed design work on the absorber (see Chapter 9 and Appendix G).

Using the heat capacity data and enthalpy data listed at the beginning of this appendix, it is possible to perform the energy balance. The minor components in the tail-gas stream are grouped together in the calculation under the heading 'other gases'.

Heat duty

$$= [n_{out} \ H^\circ_{out}] - [n_{in} \ H^\circ_{in}]$$

$$= \text{nitrogen} + \text{other gases} + \text{nitric acid} + \text{water (l)} + \text{water (g)} + NO + NO_2 + N_2O_4$$

$$= [(27 \ 050/28) \ 29.16 \ (10 - 65)]$$
$$+ [(1628/30) \ 30 \ (10 - 65)]$$
$$+ [(7000 - 2323)/63][(110 \ (45 - 65)) - 206 \ 570] + [(2323/63) \ 110 \ (45 \quad 60)]$$
$$+ [(879/1 \ 8) \ (285 \ 840)] + [(1 \ 459/1 \ 8) \ 75.4 \ (45 - 7)] + [(3208/18) \ 75.4 \ (45 - 60)]$$
$$+ [(211 /18) \ (241 \ 830)]$$
$$+ [(102/30) \ (- 90 \ 370)]$$
$$+ [(2132/46) \ (- 33 \ 800)]$$

+ [(1425/92)(− 26 500)]

= (-1 550500) + (-73040) + (-15559500) + (13989200)
+ (2 834 800)
+ (-307300) + (-1 566600) + (-410500)

= − 2 643 400 kJ

Therefore, the cooling coils must remove just over 2.5 million kJ every hour.

Top-coil section

Heat duty for top-coil section $= [T_{tout} - T_{tin})/(T_{bout} - T_{tin})]$ Heat duty

$= [(20 - 7)/(40 - 7)]\ 2\ 643\ 400$

$= 1\ 041\ 000\ kJ$

Required flowrate in top-coil section $=$ Heat duty/$[C_P (T_{tout} - T_{tin})]$

$= 1\ 041\ 000/[4.2\ (20 - 7)]$

$= 19\ 100\ kg$

Bottom-coil section

Heat duty for bottom-coil section $= [(T_{bout} - T_{bin})/(T_{bout} - T_{tin})]$
$\times$ Heat duty

$= [(40 - 20)/(40 - 7)]\ 2643400$

$= 1\ 602\ 000\ kJ$

Required flowrate in bottom-coil section $=$ Heat duty/$[C_P (T_{bout} - T_{bin})]$

$= 1\ 602\ 000/[4.2\ (40 - 20)]$

$= 19\ 100\ kg$

F.2.14 Bleaching Column

In the bleaching column, warm air strips the nitrogen tetroxide out of the 'red' acid to give a product acid completely free of dissolved nitrogen oxides. The red acid enters the top of the column at 45°C and the secondary bleaching air enters the bottom of the column at 80°C. The final product leaves the bottom of the bleaching column at a concentration of 60% wt. nitric acid (40% water). The same type of

calculation as performed on the absorber can be carried out, and yields a product acid at about 55°C, with the bleaching air leaving the column at about 50°C.

F.2.15 Tail-gas Expander

The tail-gas expander takes tail gas at 620°C and approximately 950 kPa and reduces it to atmospheric pressure at a temperature at 290°C. This process enables the recovery of around 80% of the power requirements for the compressor. See Section F.2.1 for more details.

F.2.16 Refrigeration Unit

The refrigeration unit is responsible for cooling both the make-up water and the top-section cooling-coil water (in the absorption column) to 7°C. Make-up water available at 20°C is taken from the deionized-water loop.

$$\text{Refrigeration-unit duty} = F_{muw}\, C_P\, (T_{in} - T_{out}) + F_{cw}\, C_P\, (T_{in} - T_{out})$$
$$= [2338 \times 4.2\ (22 - 7)] + [19\,100 \times 4.2\ (20 - 7)]$$
$$= 1\,190\,154\ kJ$$

F.2.17 Primary Tail-gas Warmer

The primary tail-gas warmer takes the tail gas directly from the absorption column. It provides heat exchange with two distinct streams. Hot secondary air on the way to the bleaching column provides heat initially, entering at 232°C and leaving at 80°C. This raises the tail-gas temperature from 10°C to about 40°C. Hot deionized water on route to the waste-heat boiler provides the next heat exchange, entering at 101 °C and leaving at 96°C. The tail-gas temperature rises from 40°C to about 45°C. (The calculations for the initial stage of primary gas preheat were performed in Section F.2.9 due to their direct relevance in that section).

F.2.18 Boiler Vapour/Liquid Separator

The vapour/liquid separator receives saturated steam at 4000 kPa (250°C) from the waste- heat boiler. Much of this steam leaves for the steam superheater. Some steam is condensed in order to heat the incoming deionized boiler-feed water (at 96°C). The equilibrium temperature of the liquid in this vessel is about 117°C. (These calculations have already been performed in Section F.2.7).

F.2.19 Economizer

The tail-gas exiting the expander at 290°C is used to preheat the high-pressure (HP) boiler feed water and also the low-pressure (LP) deionized water (prior to deaeration for high-pressure boiler-feed water use). The important design criteria in this exchanger is that the tail gas should not leave the exchanger at less than 65°C, otherwise problems with condensation/corrosion in the stack will result. The dew point of the tail-gas is approximately 17°C.

Heat required by HP boiler feed water
$$= F_{HPBFW} \, C_P \, (T_{in} - T_{out})$$
$$= 5470 \times 4.2 \, (102 - 40)$$
$$= 1\,424\,400 \text{ kJ}$$

Heat required by the LP deionized water
$$= F_{LPDW} \, C_P \, (T_{in} - T_{out})$$
$$= 8113 \times 4.2 \, (40 - 20)$$
$$= 681\,500 \text{ kJ}$$

Total heat duty
$$= 1\,424\,400 + 681\,500$$
$$= 2105900 \text{kJ}$$

Exit temperature of tail-gas
$$= T_{in} - [(\text{Total heat duty})/(F_{Tgas} \, C_P)]$$
$$= 290 \quad [2\,105\,900/(28\,678 \times 1.02)]$$
$$= 218°C$$

F.2.20 Steam-turbine Condenser

The calculations for the steam-turbine condenser can only be performed if the steam condensate flowrate is known. This

information is dependent on the inlet and outlet velocity of steam (as specified) from the steam turbine. A detailed design study of this unit would be required to obtain this information.

The cooling water flowrate is estimated to be about 50 000 kg. Steam condensate is returned to the deionized-water loop.

F.2.21 Deionized-water Circuit

The deionized cooling-water circuit has a dual role. First, it provides process make-up water. Second, water, is circulated from within this loop for cooling. The cooler/condenser and lower cooling coils in the absorption column must both use deionized water as the cooling medium because of the prerequisite use of stainless steel in this section of the plant.

The ion-exchange unit for treatment of scheme water to deionized water is outside the process flowsheet shown in Figure 4.1. The flowrate through this unit is 2338 kg, equal to the make-up water requirement. Also required is a fin/fan-type cooler for the circuit itself, so that the temperature of the return deionized-water stream can be lowered from 40°C to 20°C

Fin/fan heat duty $=$ (Total warm water) C_P $(T_{,,} - T_{out})$
$$= (77\ 000 + 19\ 100)\ 4.2\ (40 - 20)$$
$$= 8\ 073\ 000\ \text{kJ}$$

F.2.22 Cooling-water Circuit

The cooling-water circuit provides cooling water for the steam condenser and for the compressor intercooler. Since these units are constructed from mild steel, the cooling water used can be either sea water, river water or scheme water. Probably the second option is best because there are no further equipment materials requirements and this water is free. The total circulation through this cooling-water loop is approximately 120 000 kg (120 cubic metres).

APPENDIX G
Absorption Column Calculations
(Chapter 9)

Nomenclature used in Appendix G

a	Half initial mole fraction of nitrogen monoxide in absorption-gas stream.
A'	Empirical constant in determination of tray equilibrium.
A_a	Sieve-plate active area.
A_{ap}	Area of apron beneath tray downcomer.
A_c	Absorption column internal cross-sectional area.
A_d	Downcomer area.
A_h	Total area occupied by holes in sieve plate.
A_p	Area of perforated zone on sieve plate.
$_A$, B	Two constants whose product give the Bodenstein equilibrium constant relationship.
b	Initial mole fraction of oxygen in absorption-gas stream.
C	Degree of approach to equilibrium for nitric acid production on sieve tray surface.
C_b	Base cost of absorption-column shell.
C_{bcc}	Base cost for absorption-column cooling circuit.
C_{bt}	Base unit-cost per absorption-column tray.
C_{gas}	Average specific heat for total absorption-gas stream.
C_o	Orifice discharge-coefficient for gas passing through plate holes.
$C_{p/l}$	Estimated cost of column accessories, e.g. platforms and ladders.
C_t	Final cost of all column trays.
D_{eff}	Effective diameter of column for calculation of wind loadings.
D_i	Column internal diameter.
D_h	Sieve-tray hole diameter.

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D_o	Column outside diameter.
E	Equilibrium partial pressure of nitrogen dioxide.
f	Design shear stress for pressure testing.
F	Empirical constant in determination of minimum allowable gas-hole velocity.
F_a	Moles of nitric acid formed on sieve tray.
F_D	Design-type factor in tray-cost determination.
F_m	Materials factor in unit-cost determination.
F_p	Pressure factor in unit-cost determination.
F_t	Materials factor in tray-cost determination.
F_{tt}	Tray-type factor in tray-cost determination.
g	Acceleration due to gravity.
G_{NO}	Molar flowrate of nitrogen monoxide before oxidation.
G_{N204}	Molar flowrate of nitrogen tetroxide before oxidation.
G'_{N204}	Molar flowrate of nitrogen tetroxide after oxidation.
G_{gas}	Molar flowrate of total absorption-gas stream.
h_{ap}	Height of apron beneath tray downcomer.
h_b	Height of liquid backup in the downcomer.
h_{ow}	Height of liquid crest flowing over sieve-plate weir.
h_w	Height of sieve-plate weir.
H	Height of liquid on sieve plate.
H_c	Height of absorption column.
I	Moment of inertia.
J	Constant for stress determination.
k_1	Rate constant for oxidation of nitrogen monoxide.
k_2	Rate describing relationship of nitrogen dioxide to nitrogen tetroxide (Bodenstein).
k_3	k_4/k_2.
k_4	Ratio describing relationship of nitrogen monoxide to nitrogen tetroxide (Carberry).
K	Empirical constant in stress calculations.
K_1	Empirical constant in flooding velocity calculations.
K_2	Empirical constant in maximum flooding velocity calculations.
L	Liquid mass flowrate.
L_p	Sieve-plate hole pitch.
L_w	Weir length.
m	Mass of column and contents.
M	Wind bending moment.

n_{HNO3}	Weight % nitric acid in solution on plates.
n_{NO+NOX}	Mass concentration of nitrous gas.
p_{NO}	Partial pressure of nitrogen monoxide in absorption-gas stream.
p_{NO2}	Partial pressure of nitrogen dioxide in absorption-gas stream.
p_{NOX}	Partial pressure of nitrogen peroxide in absorption-gas stream.
p_{N204}	Partial pressure of nitrogen tetroxide in absorption-gas stream.
P	Operating pressure.
ΔP_{dc}	Downcomer pressure loss.
ΔP_{dry}	Pressure drop across dry sieve plate.
ΔP_{total}	Total pressure drop across sieve-tray plate.
s	Fraction of open area on sieve plate.
S	Submergence on sieve plate.
t	Time
t_p	Sieve-plate thickness.
t_r	Average residence time for absorption gases in the void between trays.
t_s	Minimum shell thickness.
T	Temperature.
AT	Temperature change in absorption gases after chemical reaction.
v	Vapour mass flowrate.
V_f	Gas velocity at flooding.
v_{fmax}	Maximum gas velocity at flooding.
V_h	Velocity of absorption gas passing through sieve-tray holes.
V_w	Maximum wind velocity.
V_{min}	Minimum velocity required for passage of absorption gas through sieve-tray holes.
w	Weight fraction of nitric acid solution.
W	Wind-pressure loading.
W_l	Weir length.
W_s	Mass of column.
X	Fraction of nitrogen monoxide oxidized in void space between trays.
ρ	Density.

$\Delta\rho$ Density difference between absorption-column liquid and
        vapour.

μ Ratio of moles of nitrogen peroxide to total moles of gas.

ϕ Inclusion angle on sieve tray.

π Constant (3.142).

Summary

Appendix G details the design method and sample calculations used
for specification of the absorption column. The column was designed
according to an established column geometry, conforming to the
Australian pressure vessel design standard (AS1210). The design
included a rigorous mathematical model adapted for computer
application. The mathematical model is described in Appendix G.2,
and has been written in Advanced BASIC, suitable for direct use
upon the APPLE Macintosh (512K) personal computer. The program
is described in Section G.6 and sample results are given in Table G.4

G.I   Introduction

Design of the absorption column requires determination of the
construction and appropriate operating parameters for the column,
ensuring conversion of given amounts of nitrous gases and acid
condensate.

The absorption column is sized according to two key parameters,
these are to design for optimum mass transfer and optimum unit cost.
A column internal diameter can be estimated according to the liquid
and gas flowrates by utilizing graphs and nomographs such as those
contained in Ref. A3. These recommendations have been refined
using a computer-based mathematical model. The model predicts the
required number of trays for a specified column internal diameter.
These results enable a compromise to be achieved between tower
cost and tower performance.

After selection of the column internal diameter (the fundamental
column specification), the sieve plates must then be designed. This
involves a trial and error approach. A preliminary plate design is
proposed based upon typical tray configurations, then the hydraulic

design is checked for the individual system and appropriate design changes incorporated.

The need for internal cooling within the absorption column makes a tray-type arrangement advantageous compared with a packed column, in this lattertype it is difficult to incorporate a cooling circuit. However, trays are easily adapted to include interstage cooling. Herringbone-type or pancake-type cooling coils are easily accommodated on each of the trays, and are capable of providing sufficient cooling duty.

Specifying the need for a tray-type column, the type of tray must be determined. Sieve trays are considered most appropriate for this application. They offer a simple and inexpensive construction with low pressure drop (if the hydraulic design is adequate). Bubble cap and valve-type trays offer advantages in controlling liquid droplet entrainment, but pose significant difficulties for installation of cooling coils.

G.2 Absorption Column Model

A rigorous model of the absorption column was developed for the nitric acid system. The model considers all of the key factors influencing absorption including temperature, pressure, nitrous gas concentration, and acid concentration. This model has been adapted from a similar system described in Ref. Al. Other useful models have been proposed, e.g. Ref. A2.

The model uses a 'six component' analysis. It is so named because it accounts for N_2, O_2, H_2O, NO, NO_2, and N_2O_4 within the reaction-gas stream. This model represents a significant advance in the design of absorption columns because its important feature is the ability to recognise the nitrogen tetroxide component as a separate entity, rather than including it within the larger non-specific nitrogen peroxide category. Notably, the inert component of the initial air feed is combined with the nitrogen stream (also inert to the reaction processes).

The operation of the column influences a number of key stages, each equally important to the effectiveness of the absorption. These stages are described as follows, with a description of the way in which they are modelled.

Stage _1_ The reaction gas of known composition enters the column at 65°C.

Stage 2 Further oxidation of nitrogen monoxide takes place in the void area prior to the first sieve tray. The degree of oxidation is determined using the Bodenstein equation (from Ref. AI, **p.182**):

$$k_1 \, t \; P^2 = (b - a)^{-2} \{ [X(b - a)/(1 - X)b] - \ln \, [(1 - X)/I - (aX/k_1)] \}$$

where

k_1 = rate constant for the oxidation of NO, i.e. from the equation: $\log \, k_1 = [(635/T) - 1.0261$;

t = residence time in the oxidation space (seconds);

P = operating pressure (bar);

b = initial mole fraction of oxygen in gas;

$2a$ = initial mole fraction of nitrogen monoxide in gas;

X = fraction of nitrogen monoxide further oxidized;

T = gas temperature (K).

Note The model assumes 'plug' flow in the void area between plates.

Stage 3 The reaction gas composition is recalculated based upon the oxidation of nitrogen monoxide to nitrogen dioxide.

Stage 4 The balance between nitrogen dioxide and nitrogen tetroxide is calculated. This involves solution of a quadratic equation incorporating two reaction constants. The quadratic is shown below in terms of the partial pressure of nitrogen dioxide:

$$p_{NO2}^2 \, [(2 - \mu)^2 + A \, B \, \mu] + p_{NO2} \, [A \, (2 - \mu) - A \, B \, P \, \mu \, (2 - \mu)] + A \, P \, \mu \, [2B \, P \, \mu - (2 - \mu)] = 0$$

where

p_{NO2} = partial pressure of nitrogen dioxide (bar);

P = operating pressure (bar);

μ = ratio of moles of nitrogen peroxide to total moles of **gas;**

A, B the reaction constants reported by Bodenstein (Ref. A 1; p. 182) for the equilibrium between nitrogen dioxide and nitrogen tetroxide, as given by the equation:

$$k_2 = p_{NO2}^2/p_{N2O4} = 1 \, 0^{(8.756 - 2838/T)} \, [1 - [(T/21 \, 0.3) + (790/T) - (3.8794)]p_{NOx}]$$

$$= A \, [1 \quad B \, p_{NOx}]$$

Stage 5 The gas composition is again revised. The rise in gas temperature is calculated according to the shift in the dioxide/tetroxide equilibrium and according to the oxidation of monoxide. Both reactions are exothermic.

The temperature rise in the gas mixture is given by:

$$\Delta T = [(6.2 \times 10^3 X G_{NO}) + 2.845 \times 10^4 (G'_{N2O4} - G_{N2O4})]/(G_{gas}C_{gas})$$

where

ΔT = temperature rise in gas (K);
G_{NO} = molar flow of nitrogen monoxide in gas feed (kmol/s);
G'_{N2O4} = molar flow of tetroxide after oxidation (kmol/s);
G_{N2O4} = molar flow of tetroxide before oxidation (kmol/s);
G_{gas} = molar flowrate of the total gas mixture (kmol/s);
C_{gas} = average heat capacity of the gas mixture (J/kmol K).

Therefore, the gas composition and temperature entering the first sieve plate are known.

Stage 6 The number of moles of nitric acid formed as the gas passes through the liquid mixture on the plate is calculated. This calculation includes a term that effectively describes the tray efficiency. The moles of acid formed are given by:

$$F_a = [p_{NO} + p_{NOX} - (k_3 E^3 + 2 E^2)/(k_2 + E)] (G_{NO} + G_{NOX}) C (p_{NO} + p_{NOX})^{-1}$$

where

F_a = moles of acid formed;
p_{NO} = partial pressure of nitrogen monoxide;
p_{NOX} = partial pressure of nitrogen peroxide (dioxide + tetroxide);
k_2 = dioxide/tetroxide equilibrium constant (Bodenstein);
k_3 = k_4/k_2;
k_4 = equilibrium constant describing the relationship between nitrogen monoxide and nitrogen tetroxide as presented empirically by the Carberry equation (Ref. A2, p.8):
$$\log_{10} k_4 = 7.412 - 20.28921 \, w + 32.47322 \, w^2 - 30.87 \,^3 w$$
w = concentration of acid present on the plate;
E = equilibrium partial pressure of nitrogen dioxide as

determined by solving the following equation:

$$3k_3 E^3 + (2 E^2/k_2) + E = 3p_{NO} + p_{NOX}$$

C   = degree of approach to equilibrium (tray efficiency).

This tray efficiency term (C) deserves further examination. It is derived from an empirical equation presented in Ref. AI, p. 183. The correlation accounts for the individual geometry of the sieve plate and other essential operating variables. It takes the form:

$$\frac{C}{A'} = \frac{p^{015} ["(NO + NOX)]^{0.1} X^{0.4} H^{0.15} [n_{HNO3}]^{0.1}}{W^{0.26} D_h^{0.15} s^{0.13} T^{0.87}}$$

where
P   = operating pressure bar;
$n_{(NO + NOX)}$ = mass concentration of nitrous gas (%);
X   = fraction of nitrogen monoxide oxidized in void space
               between trays;
H   = height of liquid on the plate (m);
n_{HNO3} = mass concentration of nitric acid on the plate (%);
W   = linear gas velocity (m/s);
D_h = sieve-tray hole diameter (m);
s   = fraction of open area on the sieve plate;
T   = temperature (K);
A'   = empirical constant, given by:
               A' = 8.73 for R 20.01
       or   A' = 10.86 − 1.65 $R^{970\,000}$ + 37.59 exp (− 28.8 R)
               where R = $(p_{NO} + p_{NOX})/p < 0.01$

This efficiency term (C) is potentially very comprehensive as it accounts for the key factors in plate hydraulic design, e.g. hole diameter, liquid depth, vapour velocity and open area.

Stage 7 The component mole balance is recalculated according to the moles of nitric acid formed. The evaluation technique (Stage 3) for the nitrogen dioxide and nitrogen tetroxide equilibrium balance is again employed. The gaseous concentrations calculated after the formation of nitric acid, are those used as the entry composition values to the next plate.

Stage 8 Knowing the number of moles of acid formed by chemical reaction/absorption, the moles of acid and water required from the

next plate can be calculated, and hence the concentration is determined.

Stage *9* A tray-by-tray energy balance is also possible. Those factors considered are the heat of reaction from nitrogen monoxide oxidation, heat of reaction from any shift in the dioxide/tetroxide equilibrium, heat of solution as weak acid flows from the tray above onto the current plate, and the sensible heat to cool both reaction gases and the acid solution passing through each tray. This heat balance yields the required cooling duty on each sieve tray. A simpler derivation of the equation (Ref. Al, p.184) is employed.

Stage *10* The computation proceeds to the next tray, this constitutes a return to Stage 2 whereby the fraction of nitrogen monoxide oxidized in the void space above the tray is recalculated. The steps proceed as indicated in Stages 2 to 10.

Note
(a) The calculation is terminated when the number of moles of nitric acid required from the 'next tray' is found to be less than or equal to zero.
(b) For the purpose of the model calculations, the tray temperatures are set by the known values of feed and output temperatures. The bottom tray was given the value of 65°C corresponding to the inlet gas temperature. The weak-acid feed tray was set at 50°C (the temperature of this feed condensate) and the top tray was set to 8°C (intermediate between the tail-gas exit temperature and the make-up water feed temperature). Temperatures for all other trays in the column were set by linear interpolation between these points,

G.2.1 Observations Regarding Model Performance

First, a few observations regarding the sensitivity of the column model to the various operating parameters. The formulae presented so far indicate the direct relationship between operating pressure and temperature for the absorption process. Higher pressures and lower temperatures increase the nitrous gas absorption. The amount of nitrous gases dissolved within the acid solution also increases markedly with decreasing temperature.

Second, probably the most critical operating parameter is the partial pressure of the nitrous gases. Absorption is directly proportional to the partial pressure of the nitrous-gas components. Acid production can be carried out using significantly less trays if the nitrous gas component is increased in the absorption column feed. However, this benefit is at the expense of greatly increased emission levels. The feed proportion specified at the start of this section would seem to be very close to the optimum level, with this feed composition the emissions are approximately 960 ppm and the tray requirement is 59 sieve plates.

Regarding the emission levels, it is recommended to negotiate an emissions tolerance with the EPA. Reduction of the emission levels below the required EPA limit of 15 ppm would require another 40 plates. Even inserting a catalytic combustor after the absorption column would probably only reduce emissions to about 100 ppm.

The final two parameters affecting the absorption are the column diameter and the plate design. Correct plate design is imperative for ensuring maximum plate efficiency. The column diameter dramatically affects the nitrogen monoxide oxidation, increasing the diameter provides greater void volume between trays and hence longer residence time between trays. A longer residence time also allows more time for chemical reaction. Consequently, increasing column diameter improves the fraction of the nitrogen monoxide oxidized.

G.2.2 Results from the Modelling Exercise

By varying the geometries in the program, an optimum design can be achieved. The choice of column diameter was a critical decision in the optimization procedure. In order to evaluate the effect of changing the column diameter, various column diameters were used and the number of trays required was determined. The results of these test runs are shown in Table G.I

Sieve trays are widely used in industry with column diameters up to 3.66 m (Ref. A6; p.21.74), this limit was imposed upon the testing procedure. Column diameters of less than 1.5 m would not prove economical under these conditions because of the very large tower height and number of trays required. Although a larger column diameter would substantially reduce the required number of trays,

TABLE G.1

Number of trays VS column diameter

Column diameter (m)	Number of trays
1.5	77
1.7	66
1.8	59
1.9	52
2.0	48
2.5	39
3.0	32
3.5	28

($P = 950$ kPa; $T = 65°C$ (bottom), 50°C (feed point) 8°C (top); tray spacing 0.5 m)

these calculations take no account of weepage. Weepage is the phenomenon whereby liquid starts to percolate through the sieve-tray holes instead of flowing over the weir. This can occur with low vapourvelocities. Weepage can reduce tray efficiencies by up to 90%. The minimum required vapour velocity through the holes in this column is found to be 3.7 m/s (see Appendix G.3). This makes the limiting column diameter only 1.9 m (vapour velocities 3.2-4.0 m/s). As the diameter increases, the vapour velocity falls accordingly and weepage reduces the efficiency. The model also enabled a plate spacing of 0.5 m to be selected.

G.3 Sieve-Plate Hydraulic Design

Sieve trays are used throughout the absorption column, however two distinct hydraulic designs are required. The first sieve plate design is required for trays below the weak-acid feed point. Above the weak-acid feed point, the downcoming liquid flowrate is diminished. The vapour flowrate essentially remains constant throughout the column. Different vapour to liquid ratios above and below the weak-acid feed point require a second hydraulic design to be considered.

All the sieve plates have a hole diameter of 5 mm. This is the metric equivalent of the imperial 3/16 inch. This is suggested in order to

provide the best balance between tray efficiency and cost (Ref. A5, p.14).

The sieve-plate weir length is taken as 0.8 times the column internal diameter (Ref. A5, p.14). The downcomer area is sized using a graph relating downcomer area and weir length (Ref. A3, Figure 11.31), and is found to be 15% of the column area.

The weir height (h,) is obtained from charts relating gas and liquid throughout, tray spacing and weepage limits (Ref. A4, p.298). An optimum submergence (S) is selected such that the plate operates at 20% above the weepage limit. This submergence term refers to the equivalent depth of unaerated liquid flowing over the plate and is the sum of the weir height and the liquid crest height over the weir (any reduction in the liquid level across the plate is neglected). The height of liquid over the weir crest is given by the Francis equation (Ref. A4, p.298):

$$h_w = 0.48 \ (Q_w/L_w)^{0.66}$$

where

h_w = weir height (inches);

Q_w = liquid volumetric flowrate over the weir (gal/min);

L_w = weir length (inches).

G.3.1 Sieve-Plate Design at and below the Weak-Acid Feed Point

The specification is the same for *all* plates at or below the weak-acid feed point. The only operating parameter in this region to change significantly is the temperature. The bottom plate temperature is 65°C and the weak-acid feed tray is closer to 50°C. Even this temperature difference is not expected to affect the tray design such that individual sieve plates need to be sized. The operating parameters at the bottom sieve tray are therefore given as follows:

Operating pressure	950 kPa
Tray temperature	65°C
Vapour flowrate	V = 33 000 kg/h
Liquid flowrate	L = 12 000 kg/h
Vapour density	10.1 kg/m³
Liquid density	1350 kg/m³

An initial estimate of 0.5 m for the tray spacing is selected based upon typical values suggested in Ref. A6 (p.21.75). The accuracy of this estimate can be tested in later calculations after complete speci-fication of the sieve plate.

From Ref. A4 (Figure 12), a submergence of 50 mm (2 in.) is suggested for a plate spacing of 0.5 m and for the designated vapour and liquid flowrates.

Reference A3 details the recommended plate configuration for liquid flowrate versus column internal diameter. This suggests a single-pass crossflow-type sieve plate as shown in Figure 9.1.

The other design criteria are then employed to provide the provisional plate specification given in Table G.2.

Checking for weepage

Employing the empirical correlation presented in Ref. A5 (p.14), the minimum vapour velocity required through the sieve-plate holes is given by:

$$V_{min} (\rho_{vap})^{0.5} = F - 18.8 (1 - D_h)$$

where

V_{min} = minimum vapour velocity through hole (ft/sec);
ρ_{vap} = density of gas stream (lb/ft^3);
F = constant (from Ref. A5; p.15, Figure 1);
D_h = sieve-plate hole diameter (inches).

Therefore:

$$V_{min} = (25 \quad 18.8 (1 - 3/16))/(0.629)^{0.5}$$
$$= 12.3 \text{ ft/sec}$$
$$= 3.7 \text{ m/s}$$

TABLE G.2

Provisional plate design at and below the acid feed

Column internal diameter	D_i	= 1.8 m
Column area	A_c	= 2.5 m^2
Hole diameter	D_h	= 5 mm
Hole area	A_h	= 0.08 × 2.5 = 0.2 m^2
Downcomer area	A_d	= 0.40 m^2
Sieve-plate active area	A_a	= 1.7 m^2
Plate thickness	t_p	= 5 mm
Weir length	W_l	= 0.8 × D, = 1.4 m
Submergence	S	= 50 mm
Liquid crest height over weir	h_{ow}	= 1mm
Weir height	h_w	= 50mm

Actual vapour velocities through the plates at and below the feed plate range from 4.0–4.5 m/s. This is sufficiently above the minimum required value.

Checking for entrainment
Percentage flooding for the sieve plate is obtained from the following calculations.

Flooding velocity (V,) $= V/(3600\ \rho_{vap}\ A,)$
$= 33\ 000/(3600 \times 10.1 \times 1.7)$
$= 0.53$ m/s
Liquid/vapour flow factor $= (L/V)\ (\rho_{vap}/\rho_{liq})^{0.5}$
$= (12\ 000/33\ 000)\ (10.1\ /1\,350)^{0.5}$
$= 0.24$
From Ref. A3 (Figure 11.27), $K_1 = 0.09$. Correcting for surface tension, $K_1 = 0.1$.
Flooding velocity (V_{fmax}) $= K_1\ (\Delta\rho/\rho_{vap})^{0.5}$
$= 0.1\ [(1350 - 10.1)/10.1]^{0.5}$
$= 1.15$ m/s
Percentage flooding $= 100(V_f/V_{fmax})$
$= 100\ (0.53/1.15)$
$= 46\%$
Reference A3 (Figure 11.29) predicts a fractional entrainment of approximately 0.005. From Ref. A5 (Figure 3), the fractional entrainment is 40% lower than the limiting value for this plate separation distance of 0.5 m. This implies entrainment should not present any problems.

Plate pressure drop
The dry-plate pressure drop is given in Ref. A5 (p.16) by the equation:
$$\Delta P_{dry} = 0.187\ V_h^2\ (\rho_{vap}/\rho_{liq})\ (1/C_o)^2$$
where
V_h = vapour velocity through the holes (ft/sec);
ρ_{vap} = density of gas phase (lb/ft³);
ρ_{liq} = density of liquid phase (lb/ft³);
C_o = orificecoefficient (C, = 0.775 for 3/16 inch holediameters).
The dry-plate pressure drop is found to be 13 mm (0.51 in.) of liquid.
The total pressure drop is given by the equation:
$$\Delta P_{total} = \Delta P_{dry} + S + (32.1/\rho_{liq})$$

The total pressure drop across the plate is found to be 73mm (2.86 in.) of liquid, or 0.95 kPa. This is an acceptable value.

Checking do wncomer liquid- backup height
Taking $h_{ap} = h_w - 10$ mm $= 40$mm.
Area under the apron $(A_{ap}) = 1.4 \times 0.040 = 0.056$m².
Downcomer pressure loss $(\Delta P_{dc}) = 166 [33\ 000/(3600 \times 1350 \times 0.056)]^2$
$$= 2.4 \text{ mm.}$$

Backup in the downcomer $(h_b) = 50 + 1 + 73 + 2.4 = 126.4$ mm.
Since 126.4 mm < 0.5 m (plate spacing), the plate spacing is considered to be acceptable.

Checking residence time
Residence time $(t,)$ $= 0.785 \times 0.126 \times 1350/(33\ 000/3600)$
$$= 14.5 \text{ seconds.}$$

Since the residence time is significantly more than the recommended minimum of 1.25 seconds (Ref. A2), this design is acceptable.

Determining number of holes per plate
Allow a 60 mm unperforated strip on the outer edge, and a 100 mm calming zone between the downcomers and the active zone.
From Ref. A3 (p.465, Figure 11.32) for $(L_w/D_i) = 0.8$, then $\phi = 105"$.
Angie subtended at the plate edge $= 180 - 105 = 75"$.
Mean length of unperforated strips $= (1.8 - 0.060)\ \pi\ (75/1\ 80)$
$$= 2.3 \text{ m.}$$
Area of unperforated edge strips $= 0.060 \times 2.3 = 0.137$ m²
Area of calming zones $= 2 (0.060) [1.4 - (2 \times 0.06)]$
$$= 0.02 \text{ m}^2.$$
Total area for perforations $(A,) = A, - \text{(unperforated area)}$
$$= 1.7 - (0.137 + 0.02)$$
$$= 1.58 \text{ m}^2.$$
Therefore, $A_h/A_p = 0.2/1.58 = 0.125$.
From Ref. A3 (p.466; Figure 11.33), $(L_p/D_h) = 2.7$. This value falls within the recommended range of 2.640 (Ref. A3, p.465) and the hole pitch is acceptable.
Number of holes $= 0.2/(\pi \times 0.0025 \times 0.0025)$
$$= 10200$$

Conclusion

These calculations confirm the suitability of the original design specification.

G.3.2 Sieve-Plate Design above the Weak-Acid Feed Point

The specification is the same for *all* plates above the weak-acid feed point. The only operating parameter to change significantly in this region is the temperature. The weak-acid feed plate temperature is 50°C and the top tray is close to 10°C. Even this temperature difference is not expected to alter the tray design to such an extent that individual sieve plates need to be sized. The operating parameters at the top sieve tray are therefore given as follows:

Operating pressure 950 kPa
Tray temperature   10°C
Vapour flowrate $V = 30\ 500$ kg/h
Liquid flowrate $L = 3600$ kg/h
Vapour density   10.1 kg/m³
Liquid density   **1350 kg/m³**

A provisional value of 0.5 m for the plate spacing is selected based upon typical values suggested in Ref. A6 (p.21.75). This value will be tested in later calculations after complete specification of the sieve plate.

From Ref. A4 (Figure 12), a submergence of 40 mm (1.6 in.) is suggested for a plate spacing of 0.5 m and for the designated vapour and liquid flowrates.

Reference A3 (Figure 11.28) details the recommended plate configuration for liquid flowrate versus column internal diameter. A reverse flow-type sieve plate is suggested as shown in Figure 9.3. The pitch of the sieve-tray holes is selected so that the total hole area is reduced to 0.07 times the total column area. The other design criteria employed to provide the provisional plate specification are detailed in Table G.3.

Checking for weepage

Employing an empirical correlation (Ref. A5; p.14), the minimum

TABLE G.3

Provisional plate design above the acid feed

Column internal diameter	D_i	$= 1.8$ m
Column area	A_c	$= 2.5$ m^2
Hole diameter	D_h	$= 5$mm
Hole area	A_h	$= 0.07 \times 2.5 = 0.175$ m^2
Downcomer area	A_d	$= 0.40$ m^2
Baffle length	L_h	$= 1.3$ m
Sieve-plate active area	A_a	$= 2.1$ m^2
Plate thickness	t_p	$= 5$ mm
Weir length	W_l	$= 0.8 D_i = 1.4$ m
Submergence	S	$= 40$mm
Liquid crest height over weir	h_{ow}	$= 1$mm
Weir height	h_w	$= 40$mm

vapour velocity required through the sieve plate holes is given by:

$$V_{min} (\rho_{vap})^{0.5} = F - 18.8 (1 \quad D_h)$$

where

V_{min}	$=$ minimum vapour velocity through hole (ft/sec);
ρ_{vap}	$=$ density of gas stream (lb/ft^3);
F	$=$ constant (from Ref. A5; p.15, Figure 1);
D_h	$=$ sieve-plate hole diameter (in.).

Therefore:

$$V_{min} = [25 - 18.8 (1 - 3.16)](0.629)^{0.5}$$
$$= 12.3 \text{ ft/sec}$$
$$= 3.7 \text{ m/s}$$

Actual vapour velocities through the plate above the feed plate, range from 4.0–4.6 m/s which is sufficiently above the minimum required value.

Checking for entrainment

Flooding velocity $(V_f) = V/(3600\ \rho_{vap}\ A_a)$
$= 30\ 500/(3600 \times 10.1 \times 1.7)$
$= 0.49$ m/s

Liquid/vapour flow factor $= (L/V)\ (\rho_{vap}/\rho_{liq})^{0.5}$
$= (3600/30\ 500)\ (10.1\ /1\ 350)^{0.5}$
$= 0.01$

From Ref. A3 (Figure 11.27), $K_1 = 0.09$. Correcting for surface tension and for fractional hole area: $K_1 = 0.1$.

Flooding velocity $(V_{fmax}) = K_1 (\Delta\rho/\rho_{vap})^{0.5}$
$$= 0.1 \ [(1350 - 10.1)/10.1]^{0.5}$$
$$= 1.15 \text{ m/s}$$
Percentage flooding $= 100 \ (V_f/V_{fmax})$
$$= 100 \ (0.49/1.15)$$
$$= 43\%$$

From Ref. A3 (Figure 11.29), a fractional entrainment of approximately 0.008 is predicted. From Ref. A5 (Figure 3), the fractional entrainment is 30% lower than the limiting value for this plate separation distance of 0.5 m. This implies entrainment should not present any problems.

Plate pressure drop
The dry-plate pressure drop is given in Ref. A5 (p.16) by the equation:
$$\Delta P_{dry} = 0.187 \ V_h^2 \ (\rho_{vap}/\rho_{liq}) \ (1/C_o)^2$$
where
V_h = vapour velocity through holes (ft/sec);
ρ_{vap} = density of gas phase (lb/ft^3);
ρ_{liq} = density of liquid phase (lb/ft^3);
C_o = orifice coefficient (C, = 0.775 for 3/16 in. hole diameter).
The dry-plate pressure drop is found to be 12 mm (0.51 in.) of liquid.
The total pressure drop is given by the equation:
$$\Delta P_{total} = \Delta P_{dry} + S + (32.1 \ /\rho_{liq})$$
The total pressure drop across the plate is found 'to be 62 mm (2.45 in.) of liquid, or 0.81 kPa. This is considered to be acceptable.

Checking do wncomer liquid- backup height
Taking $h_{ap} = h_w - 10 \text{ mm} = 30\text{mm}$.
Area under the apron (A,,) = 1.4 × 0.030 = 0.042 m^2.
Downcomer pressure loss $(\Delta P_{dc}) = 166 \times [30\ 500/(3600 \times 1350$
$$\times \ 0.042)]^2$$
$$= 3.7 \text{ mm}.$$
Back-up in thedowncomer $(h_b) = 40 + 1 + 62 + 3.7 = 106.7$ mm.
Since 106.7 mm < 0.5 m (plate spacing), the plate spacing is deemed acceptable.

Checking residence time
Residence time (t,) = 0.785 × 0.107 × 1350/(30 500/3600)
$$= 13.4 \text{ seconds}.$$

Since the residence time is significantly above the minimum recommendation of 1.25 seconds (Ref. A2), this design is considered acceptable.

Determining number of holes per plate
Allow a 60 mm unperforated strip on the outer edge extending along the length of the downcomer and then skirting the baffle.
From Ref. A3 (p.465, Figure 11.32), for $(L_w/D_i) = 0.8$, then $\phi = 105''$.
Angle subtended at the plate edge = 180 − 105 = 75".
Mean length of unperforated strips = (1.8 − 0.060) π (75/180) = *2.3* m.
Area of unperforated edge strips = 0.060 (2.3/2) = 0.068 m².
Area of calming zones = [0.060 (1.4 − 2 × 0.06)] + (1.35 × 0.06)
$$= 0.1 \text{ m}^2.$$
Total area for perforations (A_p) = A_p − (unperforated area)
$$= 2.1 - (0.068 + 0.1)$$
$$= 1.93 \text{ m}^2.$$
Therefore, $A_h/A_p = 0.18/1.93 = 0.092$.
From Ref. A3 (p.466, Figure 11.33), $(L_p/D_h) = 3.0$. This value falls within the recommended range of 2.6-4.0 (Ref. A3, p.465) and the hole pitch is acceptable.
Number of holes   = *0.18 (a × 0.0025 × 0.0025)*
$$= 9200$$

Conclusion
These calculations confirm the suitability of the original design specification.

G.3.3. Plate Construction

The plates are supported on a ring welded around the vessel wall, and on beams. Lateral 'L' type beams provide support for the 5 mm thick sieve plates. A feature of all plates in the column is a manway. This manway is necessary to provide access to all sections of the column. The manway is secured with removable clamps when not in use.

Another interesting feature of the plates is that the downcomer and weir arrangement are secured in slides so that their height can be adjusted for fine tuning of the column, or if feed flowrates are

permanently changed. Weepage holes are drilled in the tray inlet weirs to provide easy drainage in the event of a maintenance shutdown.

G.4 Shell Structure

A shell must be specified to provide suitable operational safety tolerances (operating pressure), and also to conform to the Australian design standard (AS1 210).

Column Design Specifications

Total column height	32 m
Internal diameter	1.8 m
Operating pressure	950 kPa
Operating temperatures	10–65°C
Number of sieve trays	59
Material of construction	SS304L (nitric acid grade)
Design stress	108 MPa
Vessel fully radiographed (joint factor 1)	

The design pressure is chosen to be 25% above the operating pressure,

$$= 1.25 \times 950$$
$$= 1200 \text{ kPa}$$

Minimum thickness (t_s) of the shell cylinder required for pressure loading:

$$t_s = (P \, D_i)/[(2 f J) - P]$$
$$= (1.2 \times 1.8)/[(2 \times 108 \times 1.1) - 1.2]$$
$$= 0.0088 \text{ m}$$
$$\approx 9 \text{ mm}$$

Allowing an extra 5 mm thickness as a corrosion allowance, the nearest BHP steel plate available (Ref. Al 1) is 16 mm.

Standard ellipsoidal heads are chosen for the ends of the vessel. The minimum thickness required is:

$$t_s = (P \, D_i \, K)/[(2 f J) - (0.2 \, P)]$$
$$= (1.2 \times 1.8 \times 0.5)/[(2 \times 108 \times 1.0) \quad (0.2 \times 1.2)]$$

= 0.0044 m

≈ 5 m m

Therefore, 16 mm plating is chosen for the column heads (for uniformity).

Checking Calculations for Other Stresses

Dead weight stress

Volume   = column shell + column heads

= $[\pi(1.8 + 2 \times 0.016)\, 0.016 \times 32]$ + $[1.1 \times 2\,\pi\,1.81^2 \times 0.01/4]$

= 2.96 m³

≈   3 m³

Density of $SS304L$ = 7.8 tonnes/m³.

Mass of shell   = volume × density

= 3 × 7.8

= 23.1 tonnes

Internals to the column (allow 100% of shell weight) = 23.1 tonnes.

Mass of liquid in the column:

5 cm deep on 59 trays;

active area of each tray = 1.7 m²;

volume of acid = 5 m³;

density of acid = 1350 kg/m³.

Mass of liquid   = volume × density

= (5 × 1350)/1 *000*

= 6.75 tonnes

Total mass   = mass of shell + mass of internals + mass of liquid

= 23.1 + 23.1 + 6.75

≈ 53 tonnes

Examining the stresses

Weight stress   = $m\, g/(\pi\, D_o\, t_s)$

= 53 × 9.8/$[\pi\,(1.8 + (2 \times 0.016))\,0.016]$

= 5635 kPa

=5.6 MPa

Axial stress   = $P\, D_o/(4\, J\, t_s)$

= 0.935 (1.8 + 2 × 0.016)/(4 × 1 .0 × 0.016)

= 26.8 MPa

Hoop stress   = 2 × Axial stress
                    = 53.6 MPa

Wind Loading

Wind velocity (max) V_w = 150 km/h
Effective diameter for wind loading D_{eff} = 1.5 D_i = 2.7 m
Density of air ρ_{air} = 1.2 kg/m^3
Dynamic wind-pressure loading is given by:
$W = D_{eff} (0.5 C_d \rho_{air})/V_w^2$
    = 2800 Pa
Bending moment (M)   = 0.5 W H_c^2
                       = 1.44 MN m

Bending stresses

Column outside diameter (D_o) = D_i + (2 t_s) = 1.832 m.
Moment of inertia (I) = $(\pi/64)$ (D$_o^4$ − D$_i^4$) = 0.04 m^4
Bending stress = (M/I)[(D_i/2) + t_s]
                 = *(1 440* 000/0.4) [(1.8/2) + 0.0161
                 = 33 MPa

Analysis of Stresses

Upwind   Total stress = Bending stress + Axial stress −
                                Radial stress
                              = 33 + 26.8 − 5.6
                              ≈ 54 M P a
Downwind Total stress = Bending stress + Hoop stress −
                                Radial stress
                              = 33 + 53.6 − 5.6
                              ≈ 81 MPa
< -------- > Hoop Stress = 54 MPa
Radial   Radial stress = 0.5 P = 500 kPa
Maximum Stress = 81 MPa
The maximum stress is more than 30% below the design stress of
108 MPa, and therefore the shell design is considered acceptable.

G.5 Estimated Cost of the Absorption Column

The capital cost estimation for this unit is revised because of the
detailed design information now available. The essential features

considered in the costing are:

Column height	32 m
Number of sieve trays	59
Column shell thickness	16mm
Material of construction	SS304L
Mass of steel	46.2 tonnes
Design pressure	1200 kPa

This unit is costed according to the recommendations of Ref. Al 3. This method utilises correlations based upon the cost of a simple carbon-steel structure. Factors are then applied to account for the cost of other materials, the inclusion of trays, for operating pressure and for incidentals such as ladders and railings. The correlations are in SI units:

Base cost (C_b) = exp $[6.950 + 0.1808$ In $(W_s) + 0.02468$ In $(W_s)^2]$

= US$125 000

Material of construction factor for stainless steel (F_s) = 1.7.

Shell cost in stainless steel = 1.7 × 125 000

= uss213 000

For platforms and ladders, the correlation estimates:

$C_{p/1}$ = 834.86 $(D_i)^{0.63316} (H_c)^{0.80161}$

= US$19 500

The unit tray-cost is given by:

C_{bt} = 278.38 exp $(0.5705 \ D_i)$

= uss777

The material of construction factor for stainless steel is given by:

F_t = 1.189 + 0.1894 D_i

= 1.53

The type of tray is accounted for by the factor (F_s) = 0.85.

Total tray cost (C_t) = 59 × 777 × 1.53 × 0.85

= US$60 000

Finally, the cost of the cooling circuit is included. This cost is estimated on the basis of the heat-transfer area (total area = 185 m^2) using the correlation:

C_{bcc} = exp $[8.202 + 0.01506$ In $(A_c) + 0.06811$ In$(A_c)^2]$

= US$26 000

Applying the design-type factor $(F_D = 0.8)$, the pressure factor $(F_P = 1.5)$, and the materials factor $(F_m = 2.4)$, then:

Cooling circuit cost = US$75 000

Summing the shell cost, the tray cost, the cost of accessories and the cost of the cooling circuit, the total cost is US$367 500.

This cost is based upon 1979 figures. The M & S costing index (Ref. AI 4) is applied to update this figure.

1979 M & S Index = 593.3

1985 M & S Index = 786.6

The 1986 cost in US dollars = 367 500 × 786.6/593.3

                                   = US$487 000

The exchange rate at 10 November 1986 was US$1 = As0.6421 (Ref. AI 5).

The final Australian cost of the absorber = 487 000/0.6421

                                   = As760 000

G.6 Program Description and Results

The mathematical model developed to describe the nitric acid absorption process is discussed in detail in Section G.2. It uses a tray-by-tray approach that incorporates reaction-mass balance corrections and heat balance calculations. Tray efficiency calculations are also included in the model, the efficiency being a function of the tray geometry and gas velocity. Rate equations and other data specific to the nitric acid/nitrous gas system are applied.

The computer program written to describe and implement the mathematical model was run on an APPLE Macintosh personal computer. A listing of the program is not included due to limitations of space. However, sample results obtained from the program for the particular design problem considered in this project are given in Table G.4.

Correspondence regarding the computer program should be addressed to D W Johnston, Shell Refinery, Corio, Geelong, Victoria, 3220.

TABLE G.4

Results from computer program for absorption column design

Plate number	Acid concentration (weight %)	Cooling duty (kW)	Plate number	Acid concentration (weight %)	Cooling duty (kW)
1	60.00	-0.721	29	12.19	6.062
2	58.12	− 0.576	30	11.16	6.072
3	56.97	− 0.375	31	10.22	6.081
4	55.69	-0.168	32	9.36	6.089
5	54.27	0.036	33	8.57	6.097
6	52.75	0.235	34	7.85	6.103
7	51.16	0.422	35	7.18	6.109
8	49.55	0.591	36	6.55	6.114
9	47.98	0.728	37	5.98	6.119
10	46.46	0.860	38	5.44	6.123
11	45.04	0.975	39	4.94	6.127
12	43.73	1.075	40	4.48	6.131
13	42.53	0.968	41	4.04	6.134
14	40.55	5.628	42	3.64	6.137
15	37.88	5.686	43	3.26	6.139
16	35.26	5.737	44	.2.90	6.142
17	32.76	5.782	45	2.57	6.144
18	30.39	5.820	46	2.26	6.146
19	28.18	5.854	47	1.96	6.148
20	26. 11	5.884	48	1.68	6.149
21	24.20	5.910	49	1.42	6.151
22	22.43	5.934	50	1.18	6.152
23	20.79	5.955	51	0.94	6.153
24	19.27	5.988	52	0.72	6.154
25	17.52	6.007	53	0.52	6.155
26	15.96	6.024	54	0.32	6.156
27	14.57	6.038	55	0.14	6.156
28	13.22	6.05 1			

Reaction-gas composition entering the absorber:

Nitrogen gas + inerts	27 482 kg	981.5 kmol
Oxygen gas	1501 kg	46.9 kmol
Nitrogen monoxide	117 kg	3.9 kmol
Nitrogen dioxide	2099 kg	45.6 kmol
Nitrogen tetroxide	1302 kg	14.2 kmol
Total	32 501 kg	1092.1 kmol

Plate I: 'Red' product acid (60% wt.) is withdrawn from this plate at 11 784 kg/hour. Dissolved contents include 1.623 kg of nitrogen monoxide and 115.474 kg of nitrogen dioxide (ignoring dissolved nitrous gas impurities).

Plate 13: Feed plate for 42% wt. weak acid at 5531 kg/hour.

Tail-gas composition leaving the absorber:

Nitrogen gas + inerts	*21482 kg*	*98* 1 SO kmol
Oxygen gas	*257* kg	*8.04* kmol
Nitrogen monoxide	**10 kg**	*0.33* kmol
Nitrogen dioxide	*25* kg	0.55 kmol
Nitrogen tetroxide	6kg	0.06 kmol
Total	*27 780 kg*	990.48 kmol

Emission levels   927 ppm   *16.34* **g**/**m³**

Make-up water required: 1532 kg/hour

APPENDIX H
Steam Superheater Calculations (Chapter 10)

Nomenclature Used in Appendix H

A Heat transfer area.

A_l Total leakage area.

A_s Crossflow area in shell.

A_{sb} Area between shell and baffles.

A_t Area of one tube.

A_{tb} Area between tubes and baffles.

A_w Area of window zone.

B_c Baffle-cut fraction.

B_b Bundle-cut fraction.

c_s Baffle to shell clearance.

c_t Tube to baffle clearance.

C Empirical constant for determination of heat-transfer coefficients.

C_c Corrosion allowance for calculation of shell ortubethickness.

C_p Average heat capacity of fluid.

D_b Tube-bundle diameter.

D_e Equivalent diameter for shell cross-section.

D_{si} Shell inside diameter.

D_{so} Shell outside diameter.

D_{ti} Tube inside diameter.

D_{to} Tube outside diameter.

E_J Joint efficiency.

F_i Tube-side fouling coefficient.

F_b Bypass correction factor.

F_l Leakage correction factor.

F_o Shell-side fouling coefficient.

F_n Correction factor for tube rows.

F_t Log-mean temperature-difference (LMTD) correction factor.
F_w Correction factor for window leakage.
G_{steam} Mass flowrate of steam through shell.
h_i Tube-side heat-transfer coefficient.
h_o Shell-side heat-transfer coefficient.
h_s Revised shell-side heat-transfer coefficient (Bell method).
H_b Height from baffle chord to top of tube bundle.
H_c Baffle-cut height.
j_f Friction factor for calculation of shell and tube side pressure drop.
j_h Heat-transfer factor.
k_f Fluid thermal conductivity.
k_s Thermal conductivity of steel.
K_1 Empirical constant for determination of tube-bundle diameter.
l_b Baffle length.
L_t Tube length.
M Molecular weight of gas.
n Empirical constant for determination of tube-bundle diameter.
N_b Number of baffles.
N_c Number of tubes in crossflow zone.
N_{cv} Number of constrictions crossed in shell fluid-passage.
N_{Nu} Nusselt number.
N_p Number of tube-side passes.
N_{Pr} Prandtl number.
N_{Re} Reynolds number.
N_s Number of sealing strips.
N_t Number of tubes.
N_w Number of tubes in the window zone.
N_{WY} Number of restrictions in crossflow for window zone.
p_t' Vertical tube pitch.
P Maximum pressure differential.
P' Average gas pressure.
P_t Tube pitch.
ΔP_c Pressure drop in crossflow area of shell flow.
ΔP_e Pressure drop over end of exchanger.
ΔP_i Ideal shell pressure-drop.
ΔP_s Shell-side pressure drop.

ΔP_t Tube-side pressure drop.

ΔP_w Pressure drop in window zone of shell flow.

R Empirical constant for finding the LMTD correction factor (F_t).

R' Ideal gas constant.

R_a' Empirical constant for determination of window-leakage correction factor.

R_w Ratio of number of tubes in window zone to total number.

S Empirical constant for finding the LMTD correction factor (F_t).

S' Maximum allowable working stress.

S_b Baffle spacing.

t_s Shell thickness.

t_t Tube thickness.

t_1 Inlet temperature of shell-side fluid.

t_2 Outlet temperature of shell-side fluid.

T Average fluid-temperature.

T_1 Inlet temperature of tube-side fluid.

T_2 Outlet temperature of tube-side fluid.

ΔT_{mean} Corrected mean-temperature difference.

ΔT_{LMTD} Log-mean temperature difference.

u_s Shell-side fluid linear velocity.

u_t Tube-side fluid linear velocity.

u_w Fluid velocity through window zone.

u_z Geometric mean velocity.

U_o Overall heat-transfer coefficient.

ρ Fluid density.

μ Fluid viscosity.

β_1 Leak factor coefficient.

π Constant (3.142).

H.I   Introduction

A split ring-type floating-heat heat exchanger is selected for use as the steam superheater because of the need to provide for thermal expansion of the tube bundle. The floating head also enables easy withdrawal of the tube bundle for cleaning purposes. This may prove very advantageous, not because the streams suffer from fouling, but

because platinum carryover from the reactor may be deposited on the walls of the tubes. Platinum recovery is made easier by providing simple access.

The reaction-gas stream is allocated to the tube side for several reasons:

(a) It is the most corrosive of the two streams.
(b) This stream will have the largest tendency for fouling.
(c) It is the higher temperature stream and therefore heat losses are minimized by passage through the tubes rather than the shell.
(d) It has the highest flowrate, and typically if the lowest flowrate stream is allocated to the shell side a better result is achieved.

The higher pressure fluid is usually allocated to the tubes rather than the shell. However, because of the advantages outlined above the tube side is selected as the best choice.

H.2 Thermal and Physical Data

The reaction-gas stream is assumed to possess the same thermal and physical properties as air. This is an acceptable approximation because the only significant change made from standard air composition is that a fraction of the oxygen is lost in favour of nitrogen monoxide. These two minor components also have very similar molecular weights.

Data:
Average heat capacity of reaction gas   1.19 kJ/(kg/K)
Average heat capacity of steam   2.62 kJ/(kg K)
Heat load   *586* kW
Other thermal properties (for air) used in the calculations were taken from Ref. E5 (Section 3). Steam properties were obtained from Ref. E6.

H.3 Preliminary Design Calculations

A provisional design is obtained based upon the known physical parameters and typical exchanger configurations. Standard pipe/

tubing sizes are used wherever possible, and the design conforms to the British design code: BS 3274. The design parameters are given in Table H.I.

The logarithmic-mean temperature difference is given by:

$$\Delta T_{LMTD} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln[(T_1 - t_2)/(T_2 - t_1)]}$$

$$= \frac{(645 - 380) - (595 - 250)}{\ln[(645 - 380)/(595 - 250)]}$$

$$= 303°C$$

The proposed design uses one shell pass with two tube passes, therefore:

$$R = (T_1 - T_2)/(t_2 - t_1)$$
$$= (645 - 595)/(380 - 250)$$
$$= 0.385$$

$$S = (t_2 - t_1)/(T_1 - t_2)$$
$$= (380 - 250)/(645 - 250)$$
$$= 0.329$$

From Ref. E2 (p.531, Figure. 12.19), the temperature correction factor is $F_t = 0.99$.
Therefore, the mean temperature difference is:

$$\Delta T_{mean} = F_t \Delta T_{LMTD}$$
$$= 0.99 \times 303$$
$$= 300°C$$

A provisional heat-transfer coefficient is obtained from Ref. E2 (p.513, Table 12.1), assumed to be typical of a transfer coefficient for two gaseous streams. This reference suggests a value of U_o = 25 W/m^2 K), however this value is multiplied by a factor of four because of the very large temperature difference and good gas velocities encountered in this application, Therefore, the value of U_o = 100 W/m^2 K) is used in the design calculations.

TABLE H. 1

Steam-superheater   duty   parameters

Reaction gas	Flowrate in/out	31 706 kg/h
	Temperature in (T,)	645°C
	Temperature out (T_2)	595°C
	Pressure in	1035 kPa
Saturated steam	Flowrate in/out	5775 kg/h
	Temperature in (t_1)	250°C
	Temperature out (t_2)	380°C
	Pressure specification	4000 kPa
Total heat duty on the unit	2108 MJ/h (586 kW)	

The provisional area is given by:

A = Heat/(U_o $\Delta T,,,,$)

  = 586 × 1 000/(100 × 300)

  = 19.5 m²

A preliminary design selection is made to use 38 mm (1.5 in.) o.d. tubes. This is a common tube size that is readily available and, since the gas flowrates are high, this should enable reasonable gas velocities to be maintained. The other factor to be considered is that larger tube sizes enable easier cleaning. If platinum needs to be recovered from these tubes, then this is certainly a worthwhile consideration. Initially tube lengths of 1.83 m (6 ft) are assumed because the exchanger area required is not large and this length is commonly available.

Note These are only provisional selections.

The tube thickness is selected according to the design constraints imposed by the design codes. Design standard ASME VIII (an American code, see Ref. E4) suggests that the tube thickness is sized according to the following formula:

$t_t = [(P\ r_e)/(S'\ E_J - 0.6\ P)] + C_c$

where

t_t = tube thickness (in.);

P = maximum pressure differential, 430 psi;

r_e = external radius of the tubes, 0.5 in.;

S' = maximum allowable working stress, 9700 psi (Ref. E4);

E_J = joint efficiency (dimensionless), 0.8 (Ref. E4);

C_c = corrosion allowance, 1 /16 in.

Therefore, t_t = 0.091 in. = 2.7 mm.

After consulting Ref. E2 (p.520, Table 12.3) concerning standard tube thicknesses, a tube thickness of 3.2 mm is selected.

Internal tube diameter = 32 mm.

Allowing for the tube sheet thickness, the effective tube length (L,) is 1.8 m.

Area of one tube (A,)   = L_t a D_{to}
                              = 1.8 π 0.031
                              = 0.175 m²

Number of tubes (N_t)   = A/A,
                              = 19.5/0.175
                              = 112

The shell-side fluid can be regarded as being relatively clean, thus a tube pitch of 1.25 times the tube outside diameter is recommended (Ref. EI; p.228). A triangular pitch is chosen. The tube bundle diameter (D_b) is given by the following equation (Ref. E2; p.523):

$$D_b = D_{to}(N_t/K_1)^{1/n}$$

where

D_b = bundle diameter (mm);

D_{to} = tube outside diameter (mm);

N_t = number of tubes;

K_1 = constant appropriate to exchanger geometry (0.249);

n = constant appropriate to exchanger geometry (2.207).

Therefore, D_b = 38.1 $(112/0.249)^{1/2\ 207}$

               = 607 mm.

From Ref. E2 (p.522, Figure 12.10), the diametrical clearance required between the shell and tube bundle is 62 mm.

Shell internal diameter (D,,) = 609 + 62 = 671 mm.

The wall thickness required for the shell is obtained from the equation given in Ref. E4:

$$t_s = [(P\ r_i)/(S'\ E_J - 0.6P)] + C_c$$

where

t_s = shell thickness (in.);

P = maximum pressure differential, 700 psi;

r_i = internal radius of the shell, 13.2 in.;

S' = maximum allowable working stress, 12 000 psi (Ref. E4);

E_j = joint efficiency (dimensionless), 0.8 (Ref. E4);

C_c = corrosion allowance, 1 /8 in.

The design pressure used in this calculation is taken as 1.2 times the normal pressure differential, i.e. 1.2 × 4000 kPa = 4800 kPa (700 psi).

Therefore, t_s = 1 .13 in. = 29 mm.

This suggests a shell outside diameter (D,,) of 727 mm.

However, if the exchanger shell is to be constructed using standard pipe (see Ref. E5; Table 6.6), the design should use a pipe of nominal size 30 and schedule number 80. This pipe specification has a thickness (t,) of 31 mm (1.219 in.), a shell external diameter (D,,) of 762 mm (30 in.), and a shell internal diameter (D_{si}) of 700 mm.

H.4 Tube-side Heat-transfer Coefficient

(The Kern Method)

Mean reaction-gas temperature = (645 + 595)/2 = 620°C.

Tube cross-sectional area = $(\pi/4)$ 32² = 804 mm².

Tubes per pass = 112/2 = 56.

Total flow area = 56 × 804 × 10^{-6} = 0.045 m².

Reaction-gas mass velocity   = (31 706/3600)/0.045
                                      = 196 kg/(s m²).

Reaction-gas average density   = (M P')/(R' T)
(ρ_{rg})   = (28.8 × 1035)/(8.314 × 625)
                                      = 5.7 kg/m³

Reaction-gas linear velocity ($V,,$) = 196/5.7 = 34.3 m/s.

Reynolds Number (N_{Re}) = V_{rg} ρ_{rg} D_e/μ_{rg}
                              = 34.3 × 5.7 × 0.02/(4 × 10^{-5})
                              = 97 800

This ensures that the flow regime is truly turbulent, and the following relationship can be used:

$N_{Nu} = C\ N_{Re}^{0.8}\ N_{Pr}^{0\ 33}\ (\mu/\mu_{wall})^{0.14}$

where

N_{Nu} = Nusselt number = h_i D_e/k_f;

C   = constant = 0.021 for gases;

N_{pr} = Prandtl number = C_p μ/k_f;

k_f = thermal conductivity of the reaction gas = 0.062 W/(m K);

h_i = tube-side heat-transfer coefficient.

Reference E5 contains a table of Prandtl numbers for air at various temperatures and pressures (Table 3-316). In this case, the Prandtl number is 0.707. (Note The viscosity term is insignificant and is neglected.)

N_{Nu} = 0.021 (97 800)$^{0.8}$ (0.707)$^{0.3}$ = 184.

The tube-side heat-transfer coefficient is given by:

$h_i = N_{Nu} k_f/D_e$ = 184 × 0.062/0.02 = 570 W/(m² K).

An alternative method uses the heat-transfer factor (j_h), and the internal heat-transfer coefficient is given by the equation (Ref. E2;p.538):

$h_i = (k_f/D_i) [j_h N_{Re} N_{Pr}^{0.33} (\mu/\mu_{wall})^{0.14}]$

The value of j_h is obtained from Ref. E2 (Figure. 12.33), its value is 0.0033 (for N_{Re} of approximately 98 000).

Therefore, h_i = 890 W/(m² K).

As a conservative approximation, the first value obtained is used for the following calculations. The second value does however confirm that the initial value was of the correct order of magnitude.

The tube-side heat-transfer coefficient is 570 W/(m² K).

H.5   Shell-side   Heat-transfer   Coefficient

(The Kern Method)

Specify six internal cross baffles.

Choosing a baffle spacing $(S_b) = D_{si}/5 =$ 700/5 = 140 mm.

Tube pitch (P,) = 1.25 × 38.1 = 48 mm.

Crossflow area (A,) $= (P_t - D_{to})/(P_t D_{si} S_b)$
$= (48 - 38.1)/(48 × 700 × 140 × 10^{-6})$
$= 0.02$ m²

Shell-side mass velocity = (5775/3600)/0.02 = 79.6 kg/(s m²).

The shell-side equivalent diameter is calculated for this triangular tube pitch using a formula in Ref. E2 (p.548):

Equivalent diameter (D,) $= (1.1/D_{to}) (P_t^2 - 0.917 D_{to}^2)$
$= (1.1/38.1) [48^2 - (0.917 × 38.1^2)]$
$= 28$ mm

Mean shell-side temperature = (380 + 250)/2 = 315°C

Steam density $= (M P')/(R' T)$

$$= (18 \times 4000)/(8.314 \times 315)$$
$$= 27.5 \text{ kg/m}^3$$

Steam viscosity $= 3.7 \times 10^{-5}$ Pa s.

Steam average heat capacity $= 2.62$ kJ/(kg K).

Steam thermal conductivity $(k_f) = 0.0475$ W/(m K).

Shell-side Reynolds number $(N_{Re}) = G_{steam} D_e/\mu_{steam}$
$$= 79.6 \times 0.028/0.000037$$
$$= 59\ 500$$

Shell-side Prandtl number $(N_{Pr}) = C_p\,\mu/k_f = 1.40$.

Selecting a 25% baffle cut (Ref. E2; Figure 11.29): $j_h = 0.0037$

This value is used in the previous formula, therefore:

$$h_o = (k_f/D_e)\, [j_h\, N_{Re}\, N_{Pr}{}^{0.33}\, (\mu/\mu_{wall})^{0.14}]$$

Therefore, $h_o = 545$ W/(m^2 K).

The shell-side heat-transfer coefficient is 545 W/(m^2 K)

H.6   Overall   Heat-transfer   Coefficient

The thermal conductivity (k,) of the steel is given as 45 W/(m K), (Ref. E8).

Two fouling factors are obtained from Ref. EI (p.845, Table 12) and converted into metric equivalents. A coefficient representative of air streams is used for the reaction-gas stream ($F_i = 0.002$). The second coefficient is a typical value for treated boiler-feed water or steam (F, $= 0.001$).

The overall heat-transfer coefficient is therefore given by:

$$1/U_o = (1/h_i) + (1/h_o) + [D_{to}\, \ln(D_{to}/D_{ti})/(2\, k_s)] + F_i + F_o$$
$$= (1/570) + (1/545) + [0.0381\ 1\ n(38.1/32)/(2 \times 45)] +$$
$$0.002 + 0.001$$
$$= 0.006663$$

The overall heat-transfer coefficient (U,) is 150 W/(m^{2}K).

It is apparent that the first design is based on a heat-transfer coefficient only two thirds of that finally calculated. This would appear to be a reasonable design basis on which to continue, although another iteration may be advisable.

H.7 Tube-side Pressure Drop (The Kern Method)

The tube-side pressure drop is calculated using the formula in Ref. E2 (p.542).

$$\Delta P_t = N_p \left[8 j_f (L_t/D_{ti}) + 2.5 \right] (\rho \, u_t \, u_t/2)$$

where

ΔP, = pressure drop across the tubes;
N_p = number of tube-side passes;
j_f = tube-side friction factor;
L_t = length of tubes;
D_{ti} = internal diameter of the tubes;
ρ = density of the gas;
u_t = velocity of gas through the tubes.

For the Reynolds number of 97 800, the value of j_f can be obtained from Ref. E2 (Figure 12.24) as 0.0027. Therefore:

$$\Delta P_t = 2 \left[8 \times 0.0027 \times (1.83/0.032) + 2.5 \right] \times 5.7 \times 34.3 \times 34.3/2 = 26 \text{ kPa}.$$

This is a satisfactory pressure drop across the tubes for this gas stream.

H.8 Shell-side Pressure Drop (The Kern Method)

The shell-side pressure drop is calculated using the formula proposed by Kern:

$$\Delta P_s = 8 j_f (D_{si}/D_e) (L_t/l_b) (\rho \, u_s \, u_s/2)$$

where

ΔP, = pressure drop across the shell;
j_f = shell-side friction factor;
D_{si} = internal diameter of shell;
D_e = shell equivalent diameter;
L_t = length of tubes;
l_b = baffle length;
ρ = density of the steam;
u_s = velocity of steam through the shell.

For a Reynolds number of 59 500, the shell-side friction factor is obtained from Ref. E2 (Figure 12.39) as 0.36. Therefore:

$$\Delta P, = 8 \times 0.36 \times (700/28) (1.83/0.305) \times 27.5 \times 2.9 \times 2.9/2 = 50 \text{ kPa}.$$

This shell-side pressure drop appears ideal considering the application.

This design based upon the Kern method appears to be quite adequate, on the basis of the results obtained.

Provisional Design

Number of tubes 112
Shell i.d. 700 mm
Bundle diameter 607 mm
Tube o.d. 38.1 mm
Tube pitch 48 mm
Baffle pitch 140 mm

H.9 Shell-side Heat-transfer Coefficient (The Bell Method)

Some doubts are expressed in Ref. E2 regarding the accuracy of the Kern design method. Therefore, the provisional design is checked using the comprehensive approach suggested by Bell (Ref. E2; p.553). The Bell method seeks to correct the shell-side heat-transfer coefficient by applying several correction factors. These factors account for the shortcomings in the Kern method, these are the effects of leakage, bypassing and flow in the window zone.

The revised shell-side heat-transfer coefficient is given by:

$h_s = h_o F_n F_w F_b F,$

Ideal Heat-transfer Coefficient

Ideal tube-bank coefficient (h,)

$$\text{Crossflow area } (A,) = (P_t - D_{to})/(P_t D_{si} S_b)$$
$$= (48 - 38.1)/(48 \times 700 \times 140 \times 10^{-6})$$
$$= 0.02 \text{ m}^2$$

Shell-side mass velocity = $(5775/3600)/0.02 = 79.6$ kg/(s m²).

The shell-side equivalent diameter is calculated for this triangular tube pitch (Ref. E2):

$$\text{Equivalent diameter } (D,) = (1.1/D_{to}) (P,' - 0.917 D_{to}^2)$$

$$= (1.1/38.1) \, [48^2 - (0.917 \times 38.1^2)]$$
$$= 28 \text{mm}$$

The mean shell-side temperature = $(380 + 250)/2 = 315°C$.

Steam density $= (M \, P')/(R' \, T)$

$$= (18 \times 4000)/(8.314 \times 315)$$
$$= 27.5 \text{ kg/m}^3$$

Steam viscosity = 3.7×10^{-5} Pa s.

Steam average heat capacity = 2.62 kJ/(kg K).

Steam thermal conductivity $(k_f) = 0.0475$ W/(m K).

Shell-side Reynolds number (N_{Re}) $= G_{steam} \, D_e/\mu_{steam}$
$$= 79.6 \times 0.028 \, /0.000037$$
$$= 59 \, 500$$

Shell-side Prandtl number $(N_{Pr}) = C_p \, \mu/k_f = 1.40$.

Selecting a 25% baffle cut (Ref. E2; Figure 11.29): $j_h = 0.0037$.

Using the formula stated previously:

$$h_o = (k_f/D_e) \, [j_h \, N_{Re} \, N_{Pr}{}^{0.33} \, (\mu/\mu_{wall})^{0.14}]$$

This yields a value of $h_o = 545$ W/(m² K).

(i) Calculating the tube-row correction factor (F_n).

Tube vertical pitch = $0.87 \, p_t' = 0.87 \times 48 = 42$ mm.

Baffle-cut height $(H_c) = 0.25 \, D_{si} = 0.25 \times 700 = 175$ mm.

Height between baffle tips $= D_{si} - 2 \, H_c = 700 - (2 \times 175)$
$$= 350 \text{ mm.}$$

Number of constrictions crossed $(N_c) = 350/42 = 8.3$

From Ref. E2 (Figure 12.32), the tube row correction factor $(F_n) = 0.98$.

(ii) Calculating the window correction factor (F_w).

Height of baffle chord (H_b) $= [D_b/2 \quad D_{si}] \, (0.5 - B_b)$
$$= [607/2 - 700] \, (0.5 - (175/700))$$
$$= 129 \text{mm.}$$

Bundle cut = $129/607 = 0.21$.

From Ref. E2 (p.564, Figure 12.41), $R_a' = 0.14$.

Tubes in one window area $(N_w) = 112 \times 0.14 = 16$.

Tubes in cross flow area $(N_c) = 112 - (2 \times 16) = 80$.

Therefore, $R_w = (2 \times 16)/112 = 0.29$.

From Ref. E2 (Figure 12.33), $F_w = 1.07$.

(iii) Calculate the bypass correction factor (F_b).

$A_b = (D_s - D_b) \, P_b = (700 - 607) \times 140 \times 10^{-6} = 0.01302$ m².

$A_b/A_s = 0.01302/0.02 = 0.651$.

From Ref. E2 (p.556), the bypass correction factor (F_b) is given by the formula:

$F_b = \exp [- 1.35 (A_b/A_s)]$

$\quad = \exp(-1.35 \times 0.651)$

$\quad = 0.415$

This suggests that sealing strips are required; assume one every three vertical rows (i.e. $N_s/N_{cv} = 0.33$). The bypass factor (F_b) is then predicted by the formula (from Ref. E2; p.556):

$F_b = \exp [- 1.35 (A_b/A_s) (1 - 2N_s/N_{cv})]$

$\quad = \exp [- 1.35 \times 0.651 (1 - 0.66)]$

$\quad = 0.75$

This result is more acceptable.

(iv) Calculate the leakage correction factor (F,).

Use clearances suggested in BS 3274.

Tube to baffle clearance (c,): $1/32$ in. = 0.8 mm.

Baffle to shell clearance (c,): $3/16$ in. = 4.8 mm.

Area between tube and baffle $\quad = c_t \pi D_{to} (N_t - N_w)$

(A_{tb}) $\qquad\qquad\qquad = (0.0008/2) \pi 0.0381 (112 - 16)$

$\qquad\qquad\qquad\qquad = 0.0046$ m²

From Ref. E2 (p.564, Figure 12.41), a 25% cut suggests $\phi_b = 2.1$ radians.

Area between the shell and baffle is predicted by the formula (Ref. E2; p.564):

$A_{sb} = c_s D_{si}/[2 (2\pi - \phi_b)]$

$\quad = 0.0048 \times 0.700/[2 (2\pi - 2.1)]$

$\quad = 0.007$ m²

$A, = A_{sb} + A_{tb} = 0.007 + 0.0046 = 0.0116$ m².

Therefore, $A_l/A_s = 0.0116/0.02 = 0.58$.

From Ref. E2 (p.558; Figure 12.35), the leak factor coefficient (β_1) is equal to 0.37. From the equation for the leak factor (Ref. E2; p.556):

$F_l = 1 - \beta_1 [(A_{tb} + 2 A_{sb})/A_1]$

$\quad = 1 - 0.37 [(0.0046 + 2 \times 0.007)/0.0116]$

$\quad = 0.45$

Therefore, the shell-side heat-transfer coefficient can be revised according to the formula given previously:

$h_s = h_o F_n F_w F_b F,$

= 545 × 0.98 × 1.07 × 0.75 × 0.45

= 193 W/(m² K)

This value is obviously much lower than that calculated by the Kern method. This means that the previously calculated value for the overall heat-transfer coefficient must also be revised.

H.10 New Overall Heat-transfer Coefficient (The Bell Method)

The new shell coefficient is used in the equation for the overall heat-transfer coefficient:

$1/U_o = (1/h_i) + (1/h_o) + [D_{to} \ln(D_{to}/D_{ti})/(2 k_s)] + F_i + F_o$

$= (1/570) + (1/193) + [0.0381 \ln (38.1/32)](2 \times 45)] + 0.002 + 0.001$

$= 0.009804$

The overall heat-transfer coefficient (U,) is 102 W/(m² K).

This new calculated value is almost one third of the original calculated value, and thus highlights the need to account for the leakage and bypass factors. Fortunately this revised overall heat-transfer coefficient is only marginally larger than the original specification value, and the design can be regarded as satisfactory.

H.ll New Shell-side Pressure Drop (The Bell Method)

The shell-side pressure drop must be recalculated using the Bell method, i.e. accounting for the various losses previously mentioned.

The shell-side pressure drop is calculated using the formula proposed by Bell (Ref. E2; p.558), and given in Chapter 10 as Equation 10.8. Therefore,

$\Delta P_i = 8 j_f N_{cv} (\rho u_s u_s/2)$

where

ΔP_i = ideal pressure drop across the tube bank;

j_f = shell-side friction factor;

N_{cv} = number of constrictions crossed (8.3);

L_t = length of tubes;

I_b = baffle length;
ρ = density of steam;
u_s = velocity of steam through the shell.

For a Reynolds number of 59 500, the shell-side friction factor can be
obtained from Ref. E2 (Figure 12.30) as 0.36.

Therefore, ΔP, $= 8 \times 0.36 \times 8.3 \times 27.5 \times 2.9 \times 2.9/2$
$$= 2764 \text{ Pa}$$

(i) The pressure drop (AP,) in the cross-flow zone is calculated.
The bypass correction factor is given by:

F_b $= \exp\left[-4.0\,(A_b/A_s)\,(1 - 2N_s/N_{cv})\right]$
    $= \exp\left[-4.0 \times 0.651\,(1 - 0.66)\right]$
    $= 0.65$

From Ref. E2 (P.561; Figure 12.38), $\beta_1 = 0.52$, therefore:

F_1 $= 1 - \beta_1\left[(A_{tb} + 2A_{sb})/A_1\right]$
    $= 1 - 0.52\left[(0.0046 + 2 \times 0.007)/0.0116\right]$
    $= 0.17$

Therefore, ΔP, $= 2764 \times 0.65 \times 0.17 = 305$ Pa.

(ii) The window-zone pressure drop (AP,) is calculated.
From Ref. E2 (p.564; Figure 12.41), for a 25% baffle cut, $R_a' = 0.19$.
The area in the window zone is equal to the difference between the
shell internal cross-sectional area and the area of those tubes
confronted in this zone:

A_w $= (\pi/4)\,(700^2 \times 0.19) - (4\,\pi/38.1^2)$
    $= 0.055 \text{ m}^2$

The steam velocity in the window zone is therefore given by:

u_w $= (5775/3600)\,(I/27.5)\,(I/0.055)$
    $= 1.06 \text{ m/s}$
u_z $= (u, u,)^{0.5}$
    $= (1.06 \times 2.9)^{0.5}$
    $= 1.75 \text{ m/s}$

The number of tubes in the window zone $(N,,) = H_b/p_t' = 129/42$
$$= 3.$$

Therefore, ΔP, $= 0.17\,(2 + 0.6 \times 3)\,27.5 \times 1.75 \times 1.75/2$
$$= 29 \text{ Pa}$$

(iii) Finally the end zone is considered
ΔP, $= 2764\,(3 + 16)/16 \times 0.66$
    $= 2166 \text{ Pa}$

Details: 2 passes on tube side, carbon-steel shell and stainless-steel
        tubes, floating head.

Use the correlations presented in Ref. E4 to estimate the purchase
cost based upon heat-exchange area.

Base cost $(C_r) = \exp [8.202 + 0.01506 \ln(A) + 0.06811 (\ln(A))^2]$

= US\$7000 (Base cost for carbon steel)

Apply the various design indices.

Design-type factor $(F_D) = \exp [-0.9003 + 0.09606 \ln(A)]$
                    $= 0.54$.

Design-pressure factor $(F,) = 1.2002 + 0.0714 \ln(A) = 1.4$.

Materials factor $(F_M) = 2.7$.

Therefore, the US exchanger cost is approximately:

Exchanger cost = $C_B F_D F_P F_M$

                = $7000 \times 0.54 \times 1.4 \times 2.7$

                = US\$14 500

These correlations are based upon 1982 United States figures and
must be adjusted for both inflation and currency differences. The M &
S equipment index as specified in Ref. E1 1 (p.7) has been used:

Final exchanger cost = 14 500 $(786.0/745.6) (1/0.61)$

                     = As25 000

This appears tn be a reasonable figure based upon other available

APPENDIX I
Pump Calculations (Chapter 11)

Nomenclature used in Appendix I

A   Total pipe area.

C_B Pump base cost.

D   Pipe diameter.

F   Dynamic friction loss per metre.

F_p Pipe friction factor.

g   Acceleration due to gravity.

G   Mass flowrate of acid through the line.

ΔH Differential head required by the pump.

N   Actual rotational speed for the pump.

N_s Specific pump speed.

N_{Re} Reynolds number.

Q   Liquid volumetric flowrate.

ΔP Pressure drop.

R   Pipe-flow resistance factor.

T   Average fluid temperature.

v Normal liquid velocity envisaged through the pipe line.

v_{max} Design maximum for liquid velocity through the pipe line.

ρ Fluid density.

μ Fluid viscosity.

π Constant (3.142).

1.1 Introduction

This appendix contains the calculations required to specify the pump for delivery of 60% wt. 'red' nitric acid from the absorption column to

the bleaching column. The pumping diagram for this arrangement is shown in Chapter 11 (Figure 11 .1). The design data required in order to specify this pump are given as follows:

Pump Specification Data

'Red' nitric acid

Flowrate	11 800 kg/h
Temperature	65°C
Density at 65°C (p)	1250 kg/m^3
Viscosity at 65°C (μ)	1.9 $\times 10^{-3}$ Pa s
Partial pressure of nitric acid	16 kPa
Partial pressure of water	80 kPa

Estimating the pump diameter required:
  Mass flowrate (G) = 11 800/3600 = 3.28 kg/s
  Volumetric flowrate (Q) = 3.28/1 250 = 2.63 $\times 1 0^{-3}$ m³/s

1.2 Piping Specification

Atypicalvelocityforfluidflow is2 m/s (Ref. P1; p.163). This enables determination of the pipe area:

Area of pipe (A)   = Volumetric flowrate/Velocity
                        = 2.63 $\times 10^{-3}/2$ m²
                        = 1.31 $\times 10^{-3}$ m²
Diameter of pipe (D) = $[4A / \pi]^{0.5}$
                        = $[4 \times 1.3 \times 10^{-3}/\pi]^{0.5}$
                        = 41 mm

This value is compared with the result achieved by applying the 'economic pipe diameter' formula for stainless steel from Ref. P1 (p.161):

Optimum diameter   = 226 $G^{0.5} \rho^{-0.35}$
                        = 226 $(3.28)^{0.5}(1 250)^{-0.35}$
                        = 33 mm

Accept the larger value as a conservative estimate. This suggests that a standard pipe of nominal pipe size 1.5, schedule number 80S, is suitable (Ref. P2; Table 6.6). This piping is 48 mm (1.9 in.) o.d. and 38 mm (1.5 in.) i.d.

Pipe cross-sectional area (A) $= (\pi/4)\, D^2$
$$= (\pi/4)\, (38 \times 10^{-3})^2 \text{ m}^2$$
$$= 1.13 \times 10^{-3} \text{ m}^2$$

1.3 Normal and Design Fluid Velocities

The normal fluid velocity is given by:
Normal fluid velocity (u)   = Volumetric flowrate/Area
$$= (2.63 \times 10^{-3})/(1.13 \times 10^{-3})$$
$$= 2.3 \text{ m/s}$$
Maximum design fluid velocity is assumed to be given by the correlation suggested in Ref. P1 (p.164):
Maximum design fluid velocity $(u_{max}) = (1.2)^2\, u$
$$= (1.2)^2 \times 2.3 \text{ m/s}$$
$$= 3.3 \text{ m/s}$$

I.4 Friction Loss per Metre of Line

Calculate the pressure drop in the line due to fluid flow (i.e. friction loss). Using Genereaux's formula (Ref. P1; p.160):
$$\Delta P_{line} = 4.07 \times 10^{10}\, G^{1.84}\, \mu^{0.16}\, \rho^{-1}\, \sim^{-4.84}$$
$$= 4.07 \times 10^{10}\, (3.28)^{1.84}\, (0.0019)^{0.16}\, (1250)^{-1}\, (38)^{-4.84}$$
kPa/m
$$= 2.22 \text{ kPa/m}$$
The friction loss can also be estimated using the formula and friction factor chart given in Ref. P3 (p.143):
$$N_{Re} = (4\, G)/(\pi \mu D)$$
$$= (4 \times 3.28)/(\pi \times 0.0019 \times 0.038)$$
$$= 58\ 000$$
The absolute roughness for commercial steel pipe is obtained from Ref. P3 (p.377, Fig. 1 O-9) as 0.045 mm.
Relative roughness = 0.045/38
$$= 1.18 \times 10^{-3}$$
For a Reynolds number of 58 000, the friction factor chart (Ref. P3; p.376, Figure 1 O-8) gives: $R/(\rho u^2) = 0.0025$.
Friction loss per metre (F) $= 4\, [R/(\rho u^2)]\, (\rho u^2/D)$
$$= 4 \times 0.0025 \times 1250 \times 2.3^2/0.038$$
kPa/m
$$= 1.7 \text{ kPa/m}$$

Accept the higher value and design for a maximum flowrate at 20%
above the normal (average) flow, therefore:

Design friction loss per metre $= (1.2)^2 \, \Delta P_{line}$
                                            $= (1.2)^2 \times 2.22 \ kPa/m$
                                            $= 3.2 \ kPa/m$

1.5 Line Equivalent Length

The pressure loss through the bends and check valves can be
included in the line pressure-loss calculations as an 'equivalent
length of pipe'. The equivalent-length recommendations are taken
from Ref. P1 (p.164). All the bends are assumed to be 90" elbows of
standard radius, and the isolation valves as plug-type valves.

Elbow equivalent length = 30 D
                               $= 30 \times 38 \times 10^{-3} \, m$
                               $= 1.14 \, m$

Plug-valve equivalent length $= 18D$
                               $= 18 \times 38 \times 10^{-3} \, m$
                               $= 0.684 \, m$

Entry losses (at maximum design velocity) are calculated from the
equation:

Entry loss $= \rho \, u_{max}^2/2$
                 $= 1250 \, (3.4)^2/2 \ kPa$
                 $= 7.2 \ kPa$

The control-valve pressure drop is suggested in Ref. P1 (p.165) as:

Maximum $\Delta P = 1.2^2 \times$ Normal ΔP
                     $= 1.2^2 \times 140 kPa$
                     $= 200 \ kPa$

Distributor pressure loss is also suggested in Ref. P1 (p.165):

Maximum $\Delta P = 1.2^2 \times$ Normal ΔP
                     $= 1.2^2 \times 15 \ kPa$
                     $= 22 \ kPa$

Referring to Figure 17. 1

Line to pump suction
From absorption column:

length = 1.0 + 3.0 = 4.0 m
elbows = 1 × 1.14 = 1.2 m
no valves

$$\underline{5.2 \ m}$$

Line from pump discharge has two options
Option 1: To bleaching column.
length = 4.0 + 5.0 + 10.0 + 0.5 = 19.5 m
elbows = 2 × 1.14   = 2.3 m
valves = 1 × 0.684   = 0.7 m

$$\underline{22.5 \ m}$$

Option 2: Recycle to absorption column.
length = 4.0 + 5.0 + 8.0 + 6.0   = 23.0 m
elbows = 2 × 1.14   = 2.3 m
valves = 1 × 0.684   = 0.7 m

$$\underline{26.0 \ m}$$

1.6 Other Pressure Considerations

Fluid head in the absorption column   0.5 m
Absorption column operating pressure   950 kPa
Bleaching column operating pressure   950 kPa

1.7 Pump Selection

The results of the pump selection calculations are summarised in Table 11.1.
Important design parameters are:
Flowrate = (2.63×10^{-3}) 3600 = 9.5 m³/h
Differential head (AH maximum) = 34 m

The best choice of pump type can be selected from Ref. P1 (p.155, Figure 5.6), this is a *single-stage, single-suction, centrifugalpump.* This is consistent with the general philosophy that a radial flow machine is best suited to higher heads and lower flows (Ref. P3; p.580).

1.8 Pump Speed

Centrifugal pumps are characterised by their specific speed (NJ. Specific speed is a dimensionless variable. Different types of pumps have different efficiency envelopes according to their specific speed. The task is to select a pump speed such that the calculated specific speed places the operating zone into the high performance envelope. The efficiency-envelope curve in Ref. P3 (p.581) is used for this determination. From Ref. P3, radial pumps show improved performance within a specific speed range of $0.02 < N_s < 0.5$, where $N_s = (N\ Q^{0.5})/(\Delta H\ g)^{0.75}$ Solving this equation within the limits of the suggested specific speed suggests a pump speed (N) within the range: 2000 rpm $< N <$ 45 000 rpm.

Following the recommendations of Ref. P4, single-stage centrifugal pumps should be operated at specific speeds greater than 0.07. A conservative estimate suggests an optimum pump speed (N) of 3000 rpm. This figure corresponds to a specific pump speed of about 0.09.

1.9 Pump Power

The pump power output iscalculated according to the power formula in Ref. P2 (p.6.5):

Power $= (AH\ Q\rho)/(3.670 \times 1\ 0^5)$

$= (34 \times 9.5 \times 1350)/(3.670 \times 10^5)$ kW

$= 1.2$ kW

This power output of 1.2 kW (1.6 hp) is small. This is expected due to the small volumetric flowrate, and the relatively small differential head.

1.10 Manufacturer's Recommendations

Having performed the calculations and obtained an approximate specification for this pump, an industrial pump supplier (AJAX Pumps, Booragoon, WA) was approached. After consulting their literature (sections reproduced at the end of this appendix), pump E32-20 was considered suitable for this application. AJAX recommend a pump power formula of the form: Power = K AH Q s.g./ Efficiency, where K is a constant depending upon the units used in the formula (s.g. is the specific gravity of the fluid being pumped). The pump power calculated using this formula is 3.5 kW, which is significantly higher than the original estimate of 1.2 kW. Other information regarding pump E32-20 is contained in the manufacturer's literature (relevant sections are reproduced at the end of this appendix).

1.11 Suction Pressure Considerations —

Cavitation

After consulting Ref. P3 (p.582; Figure 14.2), it is confirmed that the high suction pressure available in this application ensures that cavitation will not occur in the pump. For the calculated specific pump speed, there is a safety factor of over 500%. Cavitation should not be a problem provided that the column is only partially repressurised before turning on the pump (a usual start-up procedure on the absorption column).

1.12 Final Pump Specification

The final pump specification based upon the recommendations in Ref. P5 is presented in Table 11.2.

1.13 Cost Determination

The base cost for this centrifugal pump is estimated using a graph of pump costs versus head and capacity in Ref. P2 (p.6.7). A base cost

of US$900 is calculated. This base cost is revised to account for the casing pressure required and for the materials of construction, values of the modification factors (F', F") are obtained from Ref. P2 (p.6–7). Therefore, the pump cost (USS) is given by:

Pump cost $\quad = C_B$ F' F"
$\quad\quad\quad = 900 \times 1.4 \times 1.6$
$\quad\quad\quad = $ US$2020

These calculations are based upon 1979 United States figures and must be adjusted for both inflation and currency differences. The M & S equipment index (Ref. P6; p.7) has been used, therefore:

Final pump cost $\quad = 2020$ (786.0/745.6) (1/0.61)
$\quad\quad\quad = $ As3500

This cost was verified by reference to pump suppliers and found to be reasonably accurate.

The following data are reproduced from the manufacturer's technical catalogue with the permission of AJAX Pumps Pty Ltd.

TABLE I.1

Technical data for AJAX 'E' range pumps

Pump model	Nominal flange size		Maximum nominal speed, direct coupled (rpm)	Maximum vee-belt drive speed (SG = I) (rpm)	Impeller seal-ring diametral clearance (mm)		Impeller diameter (mm)		Bearing type
	suction (mm)	discharge (mm)			max	min	max	min	
E32-13	50	32	3600	2900	0.4	0.3	139	100	6305
E32-16	50	32	3600	2900	0.4	0.3	174	125	6305
E32-20	50	32	3600	2420	0.4	0.3	214	165	6305
E32-26	50	32	3000	2650	0.4	0.3	264	205	6305
E40-13	65	40	3600	2900	0.4	0.3	139	100	6305
E40-26	65	40	3000	2480	0.4	0.3	264	205	6305
E50-13	65	50	3600	2900	0.4	0.3	139	120	6305
E50-26	65	50	3000	2050	0.4	0.3	264	205	6305
E80-16	100	80	3000	2850	0.5	0.4	174	145	6305
E80-32	100	80	1800	1450	0.5	0.4	329	255	6307
E125-26	150	125	1800	1420	0.5	0.4	264	205	6307
E125-40	150	125	1500	1060	0.5	0.4	409	320	6309
E150-32	200	150	1800	1150	0.5	0.4	329	255	6309
E150-40	200	150	1500	920	0.5	0.4	409	320	6309

FIGURE I.1 General drawing of AJAX 'E' range pumps. (From manufacturer's technical catalogue, reproduced with permission.)

The following text appears within the drawing:

- φ HOLES   ROTATION

DISCHARGE

PCD

PIPE

SUCTION FLANGE

SUCTION

- φ HOLES

PCD

PIPE

- φ HOLES

DISCHARGE FLANGE
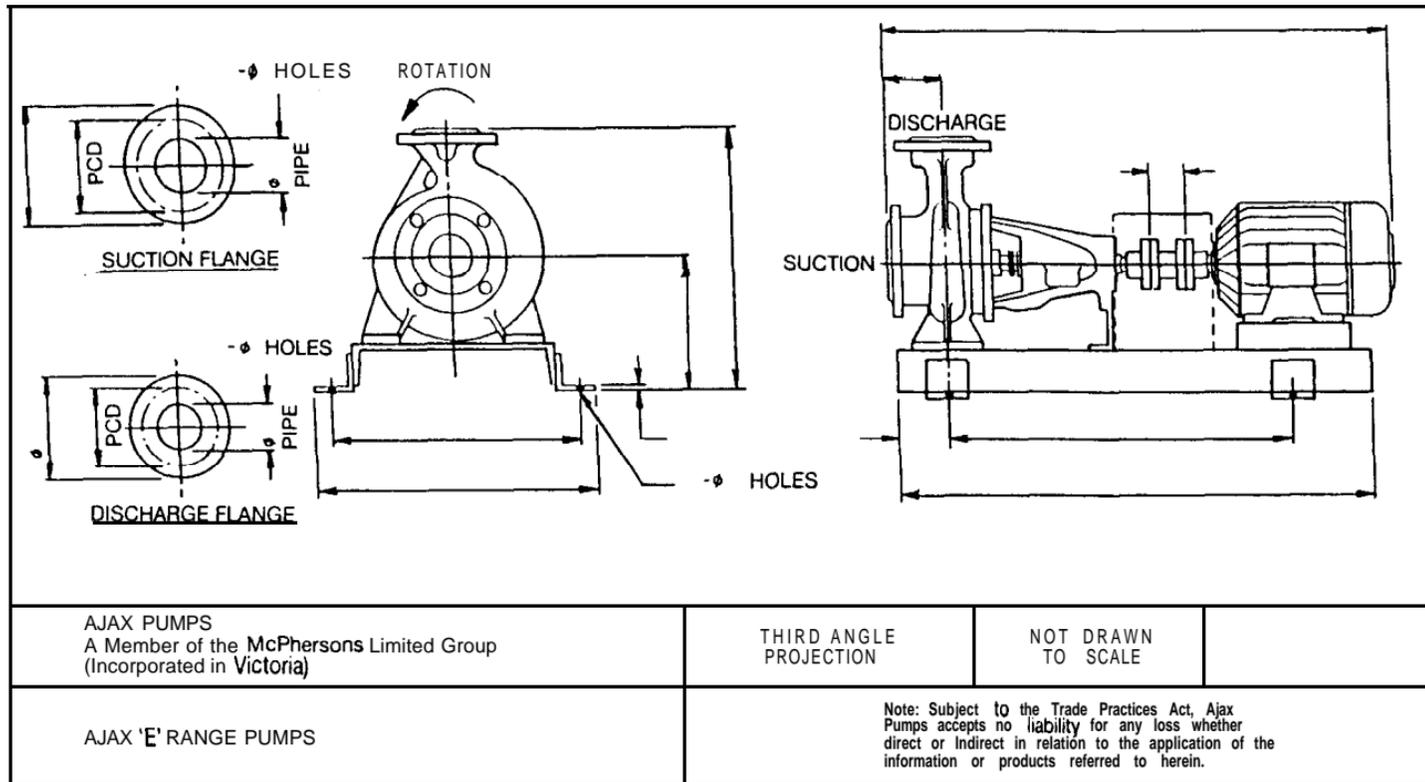
AJAX PUMPS A Member of the McPhersons Limited Group (Incorporated in Victoria)	THIRD ANGLE PROJECTION	NOT DRAWN TO SCALE	
AJAX 'E' RANGE PUMPS	**Note: Subject to the Trade Practices Act, Ajax Pumps accepts no liability for any loss whether direct or Indirect in relation to the application of the information or products referred to herein.**		

FIGURE 1.2   Pump performance curves for AJAX 'E' range pump E32–20.
(Reproduced   with   permission.)
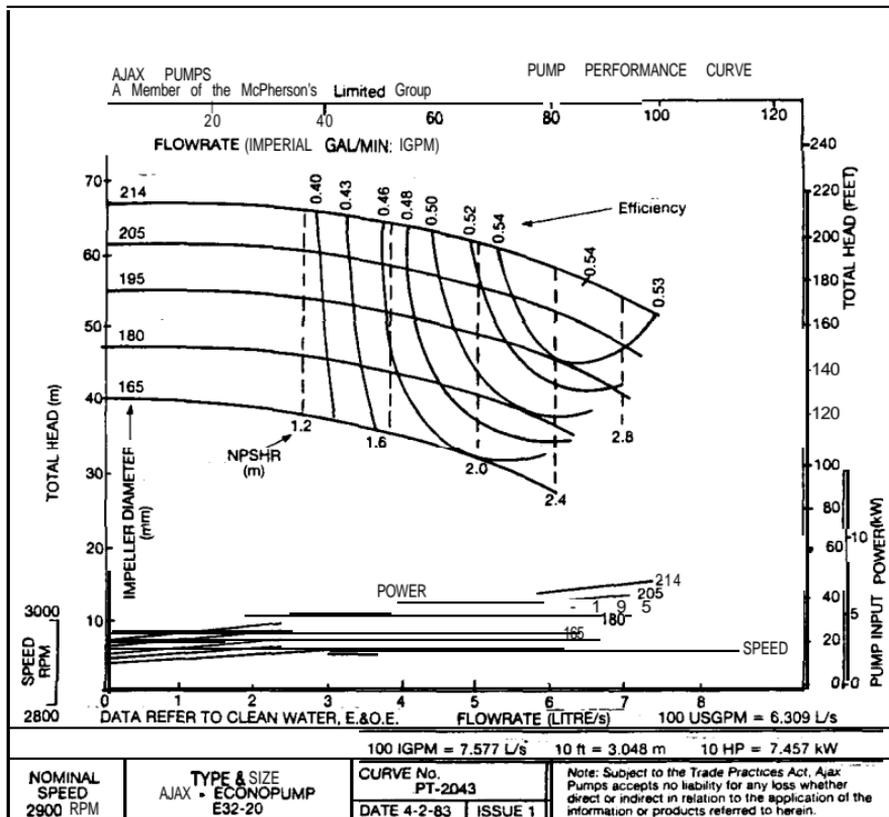
FIGURE I.3 Pump performance curves for Ajax 'E' range pump E32–20. (Reproduced with permission.)
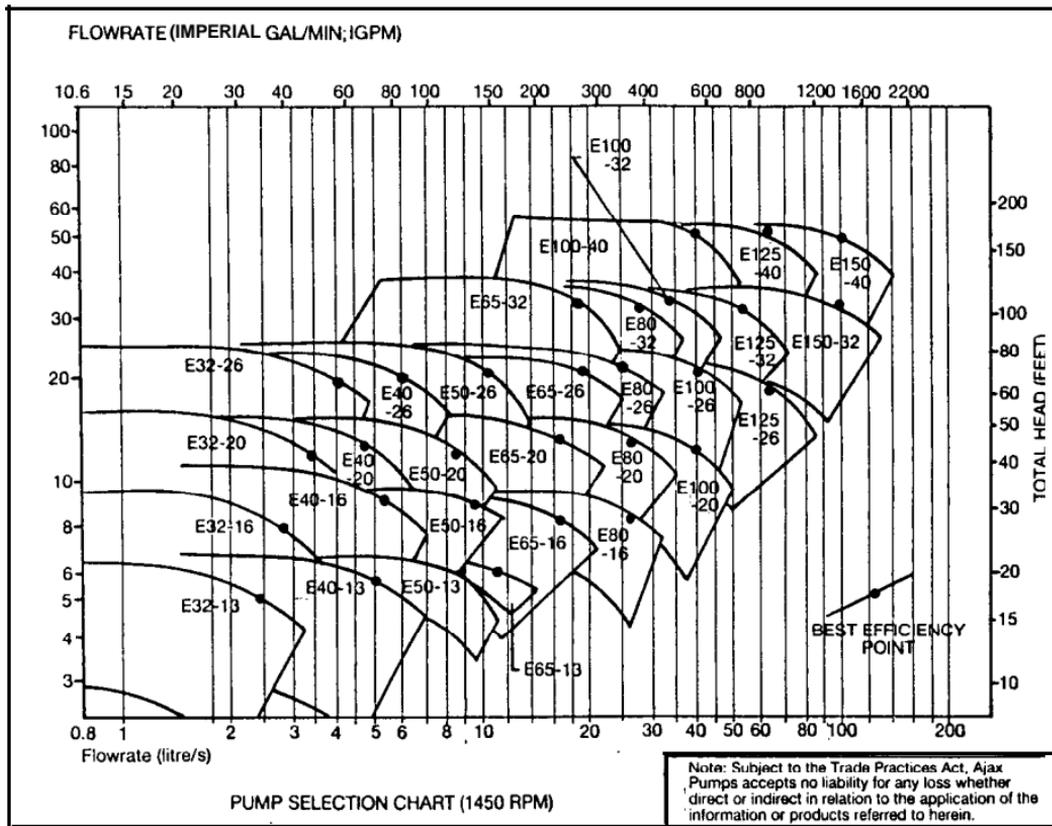
FIGURE 1.4 Pump selection chart for AJAX 'E' range pumps. (Reproduced with permission.)
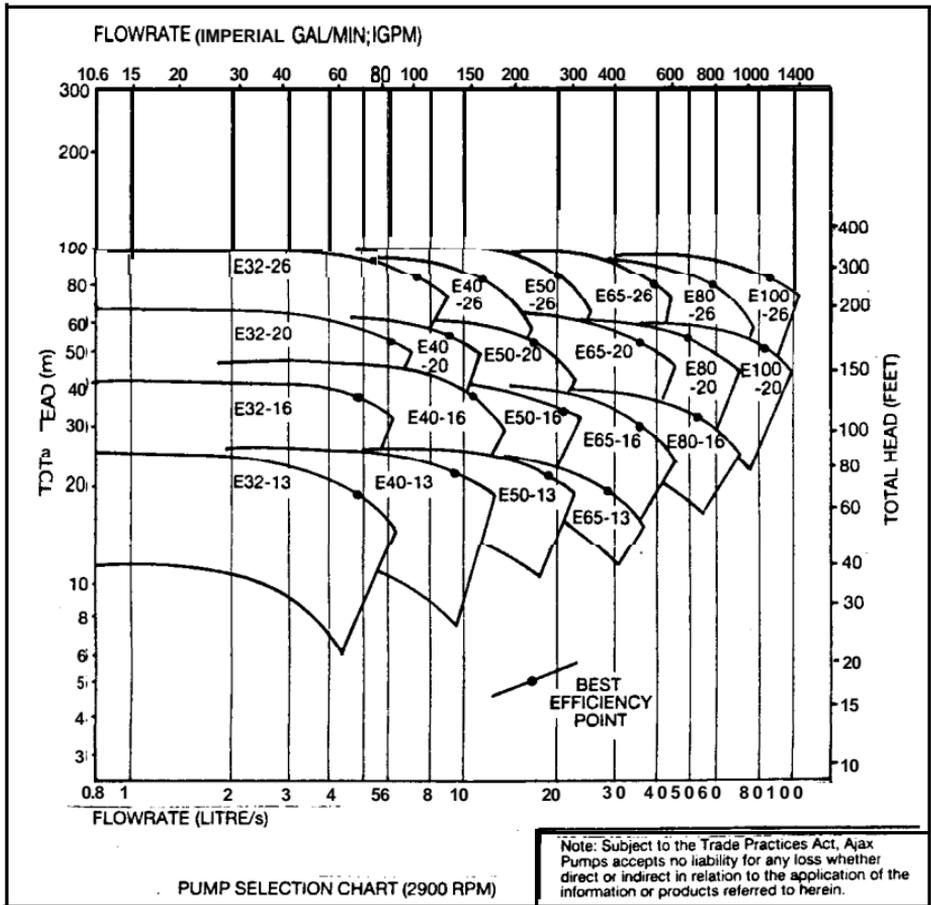
FIGURE I.5   Pump selection chart for AJAX 'E' range pumps.
(Reproduced   with   permission.)

APPENDIX J
Tank Calculations (Chapter 12)

J.I Introduction

This appendix contains the calculations related to the specification of the nitric acid product-storage tank.

The tank should have the capacity to store one week of full production, this allows continued production in the event of unscheduled shutdowns in the adjacent ammonium nitrate plant. This minimum capacity of 1500 m^3 should also prove adequate to handle any external nitric acid sales.

The tank must be constructed of stainless steel type 304L ('nitric acid grade'), the specification of this material is given in Appendix D.

The design data for this unit are specified in Table 12.1. The design tank capacity is 2000 m^3, corresponding to the standard tank size as supplied by the Denver Company (USA). The extra capacity allowance ensures that the tank is not filled beyond 80% of the total height, thus avoiding any problems of tank overflow or 'roll-over'. The design temperature represents the maximum for process-acid feed. The working pressure represents atmospheric pressure plus acid-vapour pressure at the design temperature.

5.2 Tank Dimensions

The tank dimensions required for this capacity are predetermined by selection of a standard tank configuration. The nearest tank sizing of this capacity has an inside diameter of 15.2 m (50 ft.) with a tank height of 10.7 m (35 ft.) This gives a tank capacity of 1950 m^3.

The minimum wall thickness required to support the weight of the contained acid and the roof structure is determined using the formula

given in the Australian design code AS 1210 (Ref. T1; p.76) for the specification of pressure vessels.

Thick-walled cylindrical and spherical shells (internal pressure), minimum thickness based upon circumferential stress (longitudinal joints)

$t = (D_i/2)(Z^{0.5} - 1) = (D_o/2)[(Z^{0.5} - 1)/Z^{0.5}]$

where

t = wall thickness (mm);
D_i = tank inside diameter (mm);
D_o = tank outside diameter (mm);
Z = $(FE + P)/(FE - P)$;
F = design tensile strength of material (MPa);
E = joint efficiency;
P = internal pressure (MPa).

The internal pressure is equivalent to the head of liquid inside the tank. This maximum head of 10.7 m represents a pressure of 142 kPa. The recommended wall thickness is:

$Z = (FE + P)/(FE - P)$
$= [(108 \times 0.8) + 0.142]/[(108 \times 0.8) - 0.1421$
$= 1.003$

$t = (15\ 200/2)(1.003^{0.5} - 1) = 11$ mm

Adding a corrosion allowance of 5 mm, the final recommendation is for 16 mm plating for the tank shell based on circumferential stress.

Minimum thickness based upon longitudinal stress (circumferential joints)

Formula as given above for circumferential stress, except that:

$Z = [P/(FE)] + 1$.

With the internal pressure at 140 kPa, the calculation is:

$Z = [P/(FE)] + 1$
$= [0.140/(108 \times 0.8)] + 1$
$= 1.0016$

$t = (15\ 200/2)(1.0016^{0.5} - 1)$
$= 6$ mm

This indicates that circumferential stresses are more important, and they determine the minimum shell thickness for this tank. The BHP steel plate available and closest to this specification (Ref. T2) is 16 mm thickness.

Checking Calculations for Other Stresses

Dead- weight stress
Volume of metal = Tank shell + Tank roof
$= [\pi\{15.2 + (2 \times 0.016)\}\ 0.016 \times 10.71 + 1.3\ [(\pi/4)\ 15.2^2 \times 0.0161 = 12\ m^3$
Density of SS304L = 7.8 tonnes/m^3
Mass of tank = Volume x Density
$= 12 \times 7.8$
$= 93.6$ tonnes

Considering the stresses:
Weight stress = m g $(\pi D_o t,)$
$= (93.6 \times 9.8)/(\pi\{15.2 + (2 \times 0.016))\ 0.016)$
$= 1200$ kPa
$= 1.2$ MPa
Axial stress $= P\ D_o/(4\ J\ t_s)$
$= 0.140\ \{l\ 5.2 + (2 \times 0.016)\}/(4 \times 1.0 \times 0.016)$
$= 33$ MPa
Hoop stress $= 2 \times$ Axial Stress
$= 66$ MPa
Wind stresses are disregarded because of the high ratio of tank diameter to tank height.

Analysis of Stresses

Up wind   Total stress   = Axial stress − Radial stress
$= 33 \quad 1.2$
$= 31.8$ MPa
Downwind Total stress   = Hoop stress   Radial stress
$= 66 - 1.2$
$= 64.8$ MPa

Radial Radial stress = 0.5 × P = 70 **kPa**
Maximum stress = 70 MPa

The maximum stress is approximately 35% below the design stress of 108 **MPa**, and therefore the shell thickness used in this design is considered acceptable.

inlet/Outlet Line Diameters

The inlet and outlet line diameters are sized for a recommended liquid flowrate of 2 m/s (Ref. T3; p.163).

(a) /n/et line The inlet line sizing is determined by allowing 20% above the normal product flowrate of 11 700 kg/h. This corresponds to a volumetric flowrate of 2.9×10^{-3} m³/s.

Area of pipe = Volumetric flowrate/Velocity
$$= 2.9 \times 10^{-3}/2$$
$$= 1.44 \times 10^{-3} \text{ m}^2$$

Diameter of pipe= $[\text{Area} \times 4/\pi]^{0.5}$
$$[1.44 \times 10^{-3} \times 4/\pi]^{0.5}$$
$$= 40\text{mm}$$

The nearest commercial pipe size (Ref. T4; Table 6.6) is a nominal pipe size of 1.5, schedule number **40S** (with inside diameter of 41 mm and a wall thickness of 4 mm).

(b) Out/et line The outlet flowrate is calculated based upon the need to fill a standard 30 tonne tanker in 15 minutes. This corresponds to a volumetric flowrate of 2.5×10^{-2} m³/s

Area of pipe = Volumetric flowrate/Velocity
$$= 2.5 \times 10^{-2}/2$$
$$= 1.23 \times 10^{-2} \text{ m}^2$$

Diameter of pipe= $[\text{Area} \times 4/\pi]^{0.5}$
$$= [1.44 \times 10^{-3} \times 4/\pi]^{0.5}$$
$$= 40\text{mm}$$

The nearest commercial pipe size (Ref. T4; Table 6.6) is a nominal pipe size of 6, schedule number 120 (with an inside diameter of 140 mm and a wall thickness of 14 mm).

J.3 Mechanical and Safety Features

A standard metal staircase and railing should skirt the outer edge of the tank providing access to the tank roof. A manhole in the tank roof

provides access for any internal repairs. Discrete inlet and outlet lines are required to feed into the base of the column.

A pressure relief valve is attached to the roof. This valve is opened automatically when pumping product to, or withdrawing product from, the tank. The valve shuts when pumping stops so that vapour losses from the tank are contained. A bursting disc on the roof also provides emergency pressure relief for sudden pressure rises. The final tank specification is summarized in Table 12.1.

5.4 Tank Cost

The cost of this tank is estimated from Ref. T4 (Figure 6.142). The tank base cost is estimated according to the size, material of construction and plate thickness used.

The base cost from Ref. T4 is US$225 000. These calculations are based upon 1979 United States figures and must be adjusted for both inflation and currency differences. The M & S equipment index (Ref. T5; p.7) is used, therefore:

Final tank cost = 225 000 $(786.0/745.6)(1/0.61)$

= As390 000

APPENDIX K
Design Projects Information

K.l Institution of Chemical Engineers (I. Chem. E.) Design Projects

The following list contains a summary of the design projects set by the I. Chem. E. since 1959, more details are included in the actual papers, e.g. process description, design data, references, etc. It should be possible to design these plants using information that is freely available in the technical literature.

Year	Chemical	Production (per year)	Process/Specifications
1959	Acetic anhydride (95% in acetic acid)	20 000 longtons	Thermal cracking of acetone.
1960	Monomeric styrene (95% purity)	17 500 tons	Catalytic dehydrogenation of ethyl benzene.
1961	Methyl chloride ($<$ 50ppm water, only impurity)	6000 longtons	Hydrochlorination of methanol.
1962	Butadiene ($>$ 98%w/w)	20 000 longtons	Catalytic dehydrogenation of n-butenes; feedstock of liquid mixed hydrocarbon stream containing 80.5 mol % n-butenes, 11.5 mol % n-butane, and I mol % of higher hydrocarbons.
1963	Crystalline ammonium nitrate ($<$ 0.1% moisture)	26 250 longtons	Reaction of nitric acid (47.5% w/w) and liquid ammonia.
1964	Commercial formaldehyde (formalin): 37% formaldehyde 10% methanol, 53% w/w water.	8750 longtons	Vapour-phase autothermal catalytic oxidation of methanol.
1965	Acetaldehyde (99.5%)	8750 tonnes metric	Liquid-phase hydration of acetylene.

Year	Chemical	Production (per year)	Process/Specifications
1966	Acetone (99.5 wt%)	26 250 tonnes	Vapour-phase dehydroeenation of isopropanol (85.9wt% plus 12.8% water).
1968	Aniline (99.9% w/w min)	20 000 longtons	Hydrogenation of nitrobenzene (copper on silica gel catalyst).
1969	Amine penicillin salt	10 000 kg	From a fermentation broth containing 5000 units/ml whole broth.
1970	Chlorine	10 000 longtons	Catalytic oxidation of HCl gas.
1971	Hydrogen (95% purity)	20 million standard cubic feet per day (0.555 × 10⁶ standard m³/day)	Partial oxidation of a heavy oil feedstock.
1972	Urea	100 000 tonnes (metric)	Reaction of ammonia and carbon dioxide at elevated temperature and pressure, using total-recycle process and CO_2-feed stripping.
1973	Styrene butadiene rubber (SBR) latex	25 000 tonnes	Continuous isothermal reaction (5°C) in a series of reactors ($33m^3$ capacity each).
1974	Methyl ethyl ketone	10 000 tonnes	Catalytic oxidation of secondary butyl alcohol.
1975	Acrylonitrile	100 000 tonnes	Fixed-bed catalytic reactor for ammoxidation process for propylene and ammonia reaction.
1976	Monochlorobenzene: Dichlorobenzene:	20 000 tonnes 2000 tonnes	Direct chlorination of benzene.
1977	2-Ethylhexanol	40 000 tonnes	Reaction of propylene and synthesis gas.
1978	Methanol	100 000 tonnes	Catalytic reaction of synthesis gas (CO and H_2).
1979	Gas oil	100 000 tonnes	Catalytic hydrogenation process for sulphur removal.
1980	Nitric acid (62% HNO, by weight)	100 000 tonnes	Oxidation of ammonia.
1981	Vinyl acetate	50 000 tonnes	Vapour-phase reaction of acetylene and acetic acid.
1982	Trichloroethylene: Tetrachloroethylene: Hydrogen chloride:	40 000 tonnes 7000 tonnes by product	Chlorination and cracking stages.
1983	Sulphuric acid (98%)	400 000 tonnes	Sulphur-burning process, followed by catalytic oxidation of SO_2 (vanadium pentoxide catalyst).
1984	Substitute natural gas	Feed: 600 tonnes per day of coal	Gasification reactor for processing of bituminous coal.
1985	Ethylene	30 000 tonnes	Vapour-phase catalytic dehydration of ethanol.
1986	Methanol	2000 tonnes	From natural gas by steam reforming and low-pressure synthesis.

K.2 Instructions for the I. Chem. E. Design Project, 1980

The following information is reproduced by permission of The Institution of Chemical Engineers (UK).

Before starting work read carefully the enclosed copy of *The Regulations for the Design Project* in conjunction with the following details for the Design Project for 1980.

In particular, candidates should note that all the questions should be answered in the section headed 'Scope of Design Work Required'.

The answers to the Design Project should be returned to The Institution of Chemical Engineers, 165-17 1 Railway Terrace, Rugby, CV21 3HQ, by 17.00 hours on December lst, 1980. In the case of overseas candidates, evidence of posting to the Institution on November 30th will satisfy this requirement. The wrappings must be marked on the OUTSIDE with the Candidate's name and words: 'DESIGN PROJECT'.

The Design Project will be treated as a test of the ability of the candidate to tackle a practical problem in the same way as might be expected if he were required to report as a chemical engineer on a new manufacturing proposal. The answers to the Design Project should be derived by the application of fundamental principles to available published data, they should on no account include confidential details of plant or processes which may have been entrusted to the candidate. Particular credit will be given to concise answers.

References must be given to all sources of published information actually consulted by the candidate.

The answers should be submitted on either A4 or foolscap paper, but preferably on A4. Squared paper and drawing paper of convenient size may be used for graphs and drawings respectively. The text may be handwritten or, preferably, typewritten; in the latter case it is permissible for another person to type the final copies of the answers. Original drawings should be submitted. Copies, in any form, will not be accepted.

Each sheet and drawing must be signed by the candidate and this signature will be taken to indicate that the sheet or drawing is the candidate's unaided work, except typing. In addition, the declaration forms enclosed must be filled in, signed, witnessed and returned with the answers. The manuscript, drawings and any other documents should be fastened in the folder supplied, in accordance with the instructions appearing thereon.

Answers to the Design Project itself must be written in the English language and should not exceed 20 000 words excluding calculations. The use of SZ units is compulsory.

Candidates may freely utilise modern computational aids. However, when these aids are employed, the candidate should clearly indicate the extent of his own contribution, and the extent of the assistance obtained from other sources. For computer programs which have been prepared by the candidate himself, a specimen print-out should be appended to the report. Programs from other sources should only be used by the candidate provided adequate documentation of the program is freely available in recognised technical publications. The candidate must demonstrate clearly that he fully understands the derivation of the program, and the significance and limitation of the predictions.

The answers submitted become the property of the Institution and will not be returned in any circumstances.

1980 Design Project (I. Chem. E)

Design a plant to produce 100 000 tonnes/year of nitric acid assuming an operating period of 8000 hours per year on stream.

Process Description

The process consists essentially of the oxidation of ammonia followed by the absorption of nitrogen oxides in water.

A gaseous mixture of ammonia and primary air at approximately atmospheric pressure is preheated before passing to a catalytic reactor where the ammonia is oxidised to nitric oxide. The gases leaving the reactor are cooled in a low-pressure heat-exchange system (which also serves to generate steam and preheat the ammonia/air feed stream) before passing to a water-cooled condenser. The condensate is mixed with a process water stream before entering the top of an absorption column (see below). Secondary air is added to the residual gases and the combined gaseous stream enters a compressor.

Using waste gases and cooling water, the compressed gases are cooled in a high-pressure heat-exchange system, where most of the nitric oxide is oxidised to nitrogen dioxide and most of the water vapour condenses

to produce weak aqueous nitric acid.

   The residual gas and condensate streams pass, separately, to a tray
column where the remaining oxides of nitrogen are absorbed in water.
The waste gases leaving the top of the absorption column are used to
cool the compressor discharge stream before undergoing expansion in a
turbine and final discharge as stack gases. The liquid nitric acid product
leaves from the bottom of the column.

Feed Specification

 (i) Pure gaseous ammonia is available from storage at atmospheric
     pressure and 303 K.
(ii) Atmospheric air, suitably dried, is available at 303 K.

Product Specification

 (i) Nitric acid: 62% HNO_3 (by weight) solution in water.

Operating   Parameters

 (i) *Excess air* The 'total air' used by the process is 11% in excess of that
     based on the overall equation: $NH_3 + 2O_2 = HNO_3 + H_2O$
     'total air' = primary air + secondary air.
(ii) *Ammonia oxidation reactor* Primary air/NH, ratio = 10: 1 (molar).
     Temperature = 1120 K; pressure = atmospheric.
     Catalyst: platinum-rhodium gauze.
     Conversion: 98% to nitric oxide, 2% to nitrogen.
(iii) *Low-pressure condensation* Condensate and residual gas streams
     leave at 303 K.
(iv) *Compression* Discharge pressure = 8.0 bar.
     It may be assumed that no oxidation of nitric oxide occurs during
     compression.
 (v) *High-pressure cooling* Oxidation of nitric oxide occurs according to
     the reaction: $2NO + O_2 \rightarrow 2NO_2$, and the nitrogen dioxide is in
     equilibrium with its dimer, N_2O_4. The condensate may be assumed
     to be a 50% wt aqueous solution of nitric acid consistent with the

overall equation: $4NO_2$ (g) + $2H_2O$ (1) + O_2 (g) → $4HNO_3$ (1).
Both condensate and residual gas streams leave at 303 K.

(vi) Absorption Isothermal operation at 303K. The total amount of oxides of nitrogen in the gas stream is to be reduced to less than 1500 parts per million by weight.

Utilities

(i) Saturated steam at 18 bar and 2 bar.
(ii) Cooling water at 293 K and 4.5 bar (at ground level).
 Maximum allowable discharge temperature = 320 K.
(iii) Process water, boiler feed water, instrument air, inert gas, electricity and liquid ammonia refrigerant are all available at the conditions required by the plant.

Scope of the Design Work Required

Answer all the following sections. Candidates will be expected to show full calculations in support of Sections 1 and 2.

1. *Process Design*
(a) Prepare materials balance and energy balance flow diagrams for the entire process showing a tabulated summary of the process stream flowrates (kg) and compositions (wt%) on *a basis of one hour.* Also indicate all heat exchanger duties (kW), stream pressures (bar) and temperatures (K).

(b) Prepare a process flow diagram for the plant showing all major items of equipment, in approximately correct elevation relative to each other, together with a suitable control and instrumentation scheme. Indicate all utilities requirements and the nominal size of all major pipelines. Detailed design of the compressor and turbine are not required.

2. *Chemical Engineering Design*
Prepare a detailed chemical engineering design of the absorption column assuming that sieve trays are used.

3. Mechanical Design

Prepare mechanical design sketches of the absorption tower suitable for submission to a draughtsman, paying particular attention to the tray layout and any associated cooling equipment.

4. Loss Prevention

A full-scale Design Project would include an operability study followed by appropriate hazard analysis. For the purposes of this examination, candidates are only required to make recommendations to minimize environmental pollution during operation of the plant.

Data

  (i)   *Velocity constant for nitric oxide oxidation* (k_1)

$$2 \, N \, 0 \; + \; O_2 \xrightarrow{k_1} 2NO_2$$
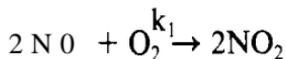
There is a discrepancy between References 1 and 2 in the value given for k, and the following expression should be used in this design project:

$$\log_{,,} \; k, \; = \; \frac{641}{T} \; - \; 0.725$$

where T is in K and k, is in $atm^{-2}s^{-1}$

  (ii) *Heat of solution of nitric acid in water at 298 K* (ΔH_s°)

$\dfrac{mol \; H_2O}{mol \; HNO_3}$	1	2	3	4	5	6	8	10	15	20	30	40	50	100
AH", (kJ/mol HNO₃)	13.1	20.1	24.3	27.0	28.8	29.9	31.2	31.9	32.5	32.7	32.8	32.8	32.8	32.8

References

1. Chilton, T.H., *Chem. Eng. Prog. Monograph Series, 55, No. 3* (1960).
2. Solomon, C.H. and Hodges, A.W., *Brit. Chem. Eng., 8,* 551 (1963).

3. Bump, T.R. and Sibbitt, W.L., *Ind. Eng. Chem., 47,* 1665 (1955).
4. Nonhebel, G., *Gas Purification Processes for Air Pollution Control,* Chapter 5, Part B: Absorption of Nitrous Gases, Newnes-Butterworths (1972).
5. Sherwood, T.K., Pigford, R.L. and Wilke, C.R., *Mass Transfer,* Chapter 8, McGraw-Hill (1975).

APPENDIX L
Information Sources

Books

The following books are arranged alphabetically by author, they all have their good points but all could be improved in several aspects-as could even the best design! These books are mainly concerned with a range of aspects of chemical engineering plant and process design, rather than particular operations such as distillation, reactor design, etc. Books describing specific chemical engineering topics are included in Chapter 8.

Aerstin, F., and Street, *G., Applied Chemical Process Design,* Plenum Press, New York (1978).

Austin, D.G., and Jeffreys, G.V., *The Manufacture of Methyl Ethyl Ketone from 2-Butanol (A Worked Solution to a Problem in Chemical Engineering Design),* The Institution of Chemical Engineers (UK) and George Godwin Ltd. London (1979).

Austin, G.T., *Shreve's Chemical Process Industries,* 5th Edn, McGraw-Hill Book Co., New York (1984).

Baasel, W.D., *Preliminary Chemical Engineering Plant Design,* Elsevier Publishing Co., New York (1976).

Backhurst, J.R., and Harker, J.H., *Process Plant Design.,* Elsevier Publishing Co., New York (1973).

Cook, T.M., and Cullen, D.J., *Chemical Plant and irs Operation (Including Safety and Health Aspects),* 2nd Edn, Pergammon Press, Oxford (1980).

Coulson, J.M.; Richardson, J.F., and Sinnott, R.K., *Chemical Engineering, Volume 6: An Introduction to Chemical Engineering Design,* Pergammon Press, Oxford (1983).

Douglas, J.M., *Conceptual Design of Chemical Processes,* McGraw-Hill Book Co.. New York (1988).

Edgar, T.F.. and Himmelblau. D.M., *Optimization of Chemical Processes,* McGraw-Hill -Book Co., New York (1987).

Felder. R.M., and Rousseau, R.W., *Elementary Principles of Chemical Processes,* 2nd Edn, John Wiley and Sons, Inc., New York (1986).

Hussain, A., *Chemical Process Simulation,* Halstead Press, New York (1986).

Kirk-Othmer Encyclopedia of Chemical Technology, 3rd Edn, 25 volumes, John Wiley and Sons, Inc., New York (1978–84).

Landau, R., and Cohan, AS., *The Chemical P/ant,* Van Nostrand Reinhold Publishing Co., New York (1966).

Luyben, W.L., and Wenzel, L.A., *Chemical Process Analysis,* Prentice-Hall, Inc., New Jersey (1988).

Mecklenburgh, J.C., *Process Plant Layout,* Halstead Press, New York (1985).

Perry, R.H., and Green, D.W., (Eds), *Perry's Chemical Engineers' Handbook,* 6th Edn, McGraw-Hill Book Co., New York (1984).

Peters, M.S., and Timmerhaus, K.D., *Plant Design and Economics for Chemical Engineers,* 3rd Edn, McGraw-Hill Book Co., New York (1980).

Raman, R., *Chemical Process Computations,* Elsevier Publishing Co., New York (1985).

Rudd, D.F., *Process Synthesis,* Prentice-Hall, Inc., New Jersey (1973).

Ulrich, G.D., *A Guide IO Chemical Engineering Process Design and Economics* John Wiley and Sons, Inc., New York (1984).

Van den Berg, P.J., and De Jong, W.A., (Eds), *Introduction to Chemical Process Technology,* Reidel-Holland, New York (1980).

Vilbrandt, F.C., and Dryden, C.E., *Chemical Engineering Plant Design,* McGraw-Hill Book Co., New York (1959).

Wells, G.L., and Rose, L.M., *The Art of Chemical Process Design,* Elsevier Publishing Co., New York (1986).

Both the American Institute of Chemical Engineers (AIChE), New York, and the Institution of Chemical Engineers (IChemE), UK, publish a wide range of symposium series, books, design guides, pocket guides, user guides, standards, directories, procedures for equipment testing, etc. It would be useful to obtain a catalogue of the publications from each institution.

Chemical Engineering magazine (published by McGraw-Hill, New York) publishes reprints of particular articles on selected topics, e.g. distillation, absorption, design, etc. In addition to these reprints of small groups of related published papers, bound volumes containing usually 100 or more relevant papers are also available, titles include Physical Properties, Process Technology and Flowsheets (Volumes I and II), Capital Cost Estimation, Process Heat Exchange, Modern Cost Engineering, etc.

Journals

The following journals are useful to the process engineer and are arranged in order of (decreasing) usefulness to the student undertaking a chemical engineering design.

Chemical Engineering ▪ published biweekly, the 'feature articles' provide excellent updates and overviews of particular topics. Single article reprints and bound volumes of selected papers are also available.

Hydrocarbon Processing ▪ excellent flowsheets, thermodynamic data series, and major articles.

Chemical Engineering Progress ▪ published by the AIChE (New York), articles concerning engineering and technical subjects. Also produces a symposium series of volumes on selected topics.

The Chemical Engineer ▪ published monthly by the IChemE (UK).

Chemical Week and *Chemical and Engineering News* ▪ provide facts and figures of the chemical industry.

Industrial and Engineering Chemistry Process Design and Development

Industrial and Engineering Chemistry Product Research and Development (These two journals ceased publication in 1986 and were replaced by *Industrial and Engineering Chemistry Research).*

Chemical Technology

Chemistry and Industry

Oil and Gas Journal

Process Engineering

Journal of Chemical and Engineering Data

Journal of Physical and Chemical Reference Data

The following journals feature papers concerned with research studies:

AIChE Journal

Canadian Journal of Chemical Engineering

Chemical Engineering Research and Design (formerly the *Transactions of the IChemE}.*

Chemical Engineering Science

The Chemical Engineering Journal

Chemical Engineering Communications

Industrial and Engineering Chemistry Fundamentals (ceased publication in 1986)

International Chemical Engineering

International Journal of Heat and Mass Transfer

Journal of Heat Transfer (Transactions of the ASME)

Journals dealing with specific subjects are also available, e.g. *Cost Engineering, Plastics World,* etc.

Other Sources

Chemical Abstracts and ***Engineering Index*** provide a useful source of data from the technical literature (chemistry and engineering) published worldwide. Other indexes associated with metals, mining, etc., are also available. Information can also be obtained from symposium series, conference proceedings and company literature.

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Topics in chemical engi
Volume 6

CHEMICAL ENGINEERING DESIGN PROJECT
A Case Study Approach

Martyn S. Ray and David W. Johnston

The work is a concise text for final year chemical engineering under-graduates, providing fundamental and practical instruction for the successful completion of a design project. Utilizing an original and award-winning student project as a model, the text guides the reader through all major aspects of a chemical engineering design, providing practical advice and background information illustrated by a detailed case study example. The work is unique in examining all stages of a design project, providing comprehensive information on both the necessary initial technical and economic feasibility study and the detailed design stage with its associated topics. The book is an invaluable source of information and example for all final year chemical engineering graduates engaged in a design project.

About the authors
Educated at the University of Surrey, UK, Martyn S Ray holds a B.Sc. and Ph.D. in Chemical Engineering. He has industrial experience gained as a chemical engineer with BOC International (UK) Ltd. and has held posts as Lecturer in Chemical Engineering at Huddersfield Polytechnic (UK), the University of the West Indies and the Polytechnic of Wales. He is currently a Lecturer in Chemical Engineering at Curtin University of Technology Western Australia. A Member of the Institution of Chemical Engineers, he has published several papers and three previous textbooks. David W Johnston graduated from Curtin University of Technology with in Chemical Engineering, winning the 1987 National Student De Award for his design project. He is currently employed as a che engineer with Shell Refining Australia Pty, Ltd.

Other titles in the series
Heat and Mass Transfer in Packed Beds, N Wakao & S Kaguei
Three-Phase Catalytic Reactors, P A Ramachandran & R V Chaudari
Drying: Principles, Applications and Design, by Cz Strumillo & T Kudra
The Analysis of Chemically Reacting Systems: A Stochastic Approach, L K Doraiswamy & B D Kulkarni
Control of Liquid–Liquid Extraction Columns, by K Najim