DEVELOPMENT OF RAPID CAPITAL COST ESTIMATION TECHNIQUES FOR THE CHEMICAL PROCESSING INDUSTRIES

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SUMMARY

Development of Rapid Capital Cost Estimation Techniques

in the Chemical Process Industries

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This thesis describes new developments in the 'step counting' concept of rapid pre-design capital cost estimation of chemical process plants. A series of simple, easy to use, mathematical cost models of varying sophistication and accuracy has been derived for estimating the battery limits investment cost of chemical plants processing gaseous/vapour phase materials.

The research was developed from a literature survey of existing methods of capital cost estimation. Particular attention was given to pre-design estimating techniques which are reviewed and their shortcomings identified. From this analysis, new theories for advancing step counting methods are proposed, with careful attention being paid to the definition of the parameters in the cost correlations. These theories were applied to over one hundred sets of published capital cost data for petrochemical and organic chemical processes handling only gas phase materials. After careful standardisation a wide range of cost models of varying complexity were derived using multi-linear regression modelling techniques. The results obtained showed very good agreement with the source data, but when compared with other estimation procedures and tested on recent cost data, gave consistent but unnacceptably low results. This is due to a variety of factors of which unsatisfactory cost indices is believed to be the most significant. The models derived have been accordingly modified to give acceptable results.

The range of models form a flexible and practical capital cost estimating package capable of meeting the various estimating situations and needs encountered by both process and cost engineers in the early stages of process design and development.

Key Words

Capital Cost
Estimation
Cost Indexes
Processes
Economics

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1 INTRODUCTION

The aim of all commercial organizations is to make profit. Hence, any organization considering investing money in the development and construction of a chemical process plant will carry out a detailed economic analysis regarding its profitability before deciding whether or not the plant should be built.

This analysis will consist of comparing the forecasted incomes from product sales against the expenditures required for production, namely the capital and operating costs, in order to evaluate its economic viability and potential. This research study is concerned with only one part of this complex exercise; that of forecasting or estimating the capital cost of the process.

Because the chemical industry is an extremely competitive one it requires sophisticated and accurate methods of cost estimating. This requirement applies to both the manufacturing organizations, who invest money in process plant with the aim of recouping the investment, and more, through product sales over a period of years, and to the contracting organizations who have to build plants within the cost they have tendered to make a profit. It is essential to both that their estimates are correct. The contracting organizations are very competent at estimating costs. They have the necessary skills, experience and up to date cost information to produce accurate figures. Their estimates are specific, detailed costings based on the process and mechanical designs derived from the customers specifications and are made to very tight error limits.

The manufacturing or client organizations however, although notable exceptions do exist, very often do not have the skills and information that are available to the contracting firm and as a result their cost estimating procedures are usually less sophisticated and consequently less accurate. Furthermore, their estimating requirements are more varied since they not only have to concern themselves with finalised and specific construction estimates of their proposed projects but also they have to ensure that the plant they commit themselves to build is economically viable and the most profitable one. Hence, they need cost estimates capable of screening alternative process routes in order to select their optimum production unit. It is desirable for them to complete this screening study as early as possible in process developments, certainly before detailed process design is required, in order that valuable time, money and engineering manhours are not wasted.

It is clear then that cost estimates are required for a process right from its initial conception, through its development stages of chemical and detailed mechanical engineering design, to its final construction and start up, in order that it may be constantly monitored and evaluated.

Since it is known that estimating accuracy is very much dependent upon the amount and quality of the information available to produce the estimate the most significant stage in process development from a cost estimating point of view is the chemical engineering design stage. At this point concrete, detailed information about the process hardware becomes available to the estimator and as such he can begin to use accepted 'in-house' estimating methods and historical cost information together with outside equipment vendor assistance, to produce fairly accurate and reliable estimates in which he has a high degree of confidence. From this point onwards, as the more detailed engineering proceeds and construction information such as the project schedule becomes available, estimating accuracy and reliability improve rapidly.

Before chemical engineering design however, the estimator has to proceed with a high degree of uncertainty and is faced with the problem of producing estimates to which a high significance is attached, from a limited amount of fundamental process information. This information includes the conceptual flowsheet, the required process capacity, probable materials of construction, operating conditions and material phases together with the location and planned construction date of the plant. Pre-design estimating has considerable importance in process development studies by providing a major screening criterion to decide whether or not to proceed with more experiments or design. It has been the subject of much research over the past fifteen years but, it still seems to be an area in need of more study and in which much improvement in existing techniques is required.

This research project then will concentrate its study on pre-design cost estimating with the aim of producing a range of rapid cost estimating models of varying sophistication and accuracy which are capable of being applied to the various situations encountered during the early stages of process development.

In order for this research to be of practical use the estimating models produced should be easy to use and understand, involve little calculation and thus minimum cost, and be accurate to within the desired limits for their intended application so that they may be of use in a wide range of situations.

2 LITERATURE REVIEW OF EXISTING CAPITAL COST ESTIMATING METHODS

2.1 Introduction

Although it is intended to develop step counting pre-design estimating methods it is considered important and necessary to gain a complete knowledge and understanding of existing estimating techniques so that they may be utilized in the research if and when applicable. A survey was made of as wide a range as possible of estimating methods and those widely recognised as significant and useful are identified. Emphasis is given to existing pre-design estimating methods which are examined in detail. In order that the methods may be described and discussed in a logical manner they are grouped and classified so that they are presented in a coherent format as described below.

2.2 Estimate Classifications

A wide range of estimating methods exist. They are generally classified according to their accuracies (1) or by the type of estimating theory employed in their derivation. (2,3). Classification by accuracy is considered most suitable as it is intended to produce a wide range of cost models of varying sophistication and accuracy which can be applied throughout process development to accommodate the variety of situations which exist. How, and where, each model will be used will depend on its accuracy, together with any information constraints it may have. Hence, estimate accuracy is of major importance. As such it is thought that by classifying and analysing a method from an accuracy viewpoint it is more likely to be determined why it is (or is not) successful and identify the factors which cause it to be so, relative to other methods.

It was decided to adopt the classification system proposed by the American Association of Cost Engineers (AACE) in 1958 (1) to describe and discuss the major capital cost estimating methods is present day use. It is considered to be a logical, definitive, and comprehensive system and is as follows:-

1. Order of Magnitude Estimates (least accurate)
2. Study Estimates
3. Preliminary Estimates
4. Definitive Estimates
5. Detailed Estimates (most accurate)

Each of the above groups serve different purposes and have different accuracies. Each is studied and the specific estimating methods within the group analysed in order to understand the techniques available and to identify the respective advantages and disadvantages associated with their use. From this critical review of the literature it is hoped to identify and overcome previous problem situations.

2.2.1. Order of Magnitude Estimates

Order of magnitude estimates have variable accuracies but in general they have probable errors of ± 50%. They are often referred to as ratio methods because they are mostly derived from historical ratios noted between cost and plant size. They produce a very quick estimate of capital cost and are obtained without flowsheet or detailed equipment information by applying ratios and escalation to published data for previous similar installations. That they do require historical cost data is a severe limitation on their use. However, these methods require minimal calculation effort and if used correctly can estimate costs with sufficient accuracy for preliminary process economics studies and subsequent decision making.

The well known order of magnitude methods in use today are reviewed below:

2.2.1.1. Turnover Ratio Method

Turnover ratio is defined as the ratio of gross annual rates (calculated as the annual production rate x average products selling price) to the fixed capital investment. The annual predicted product sales value divided by the relevant turnover ratio therefore is used to obtain a rough estimate of the capital cost. The ratios are usually derived from the users' own files or from published sources, of which a number exist. Aries and Newton (4) give turnover ratios for a number of different chemicals;

Kidoo (5) lists ratios for specific chemical processes; Lynn (6) reports ratio values for different industry types, and Wessel (7) supplies ratios for specific companies.

Turnover ratios of less than one are usually found for large volume, stable market intermediates manufactured from basic raw materials. In chemical plants where raw materials are a major portion of manufacturing cost, high labour costs exist, or high risks are encountered, turnover ratios of greater than one generally apply.

The major advantage of the method is that it is inflation proof. However, Schweyer (8) noted a limitation which causes it to give unpredictable cost estimates in that it requires an accurate forecast of sales volume and price. These are sensitive areas and subject particularly to fluctuation in the economic environment, such as those in the past eight years.

2.2.1.2. Unit Price Methods (or Investment Cost per Unit of Capacity) [9,10]

This is a simple rule of thumb method in which the annual plant capacity is multiplied by a unit cost to give the total capital cost. The unit price is typically expressed as installed capital cost per ton of annual production, based on historical data.

The most common error associated with the method is assuming a particular unit price is essentially constant over a range of plant capacities. This is not the case (see section 2.2.1.3) as is illustrated by the curve in

Figure 2.2.1.2.1. which shows the typical and significant variation in installed capital cost decreasing as plant capacity increases. It is possible however to determine the unit investment costs which apply for average conditions. The cost estimate for a given process can then be easily obtained by multiplying the appropriate investment cost per unit of capacity by the annual production capacity of the proposed plant.

Many sources exist in the literature giving the fixed capital investment required for various processes per unit of annual production capacity (11-13). The method is not limited to process 'battery limits' estimates and is often used for estimating offsite costs, specific unit operations or unit processes.

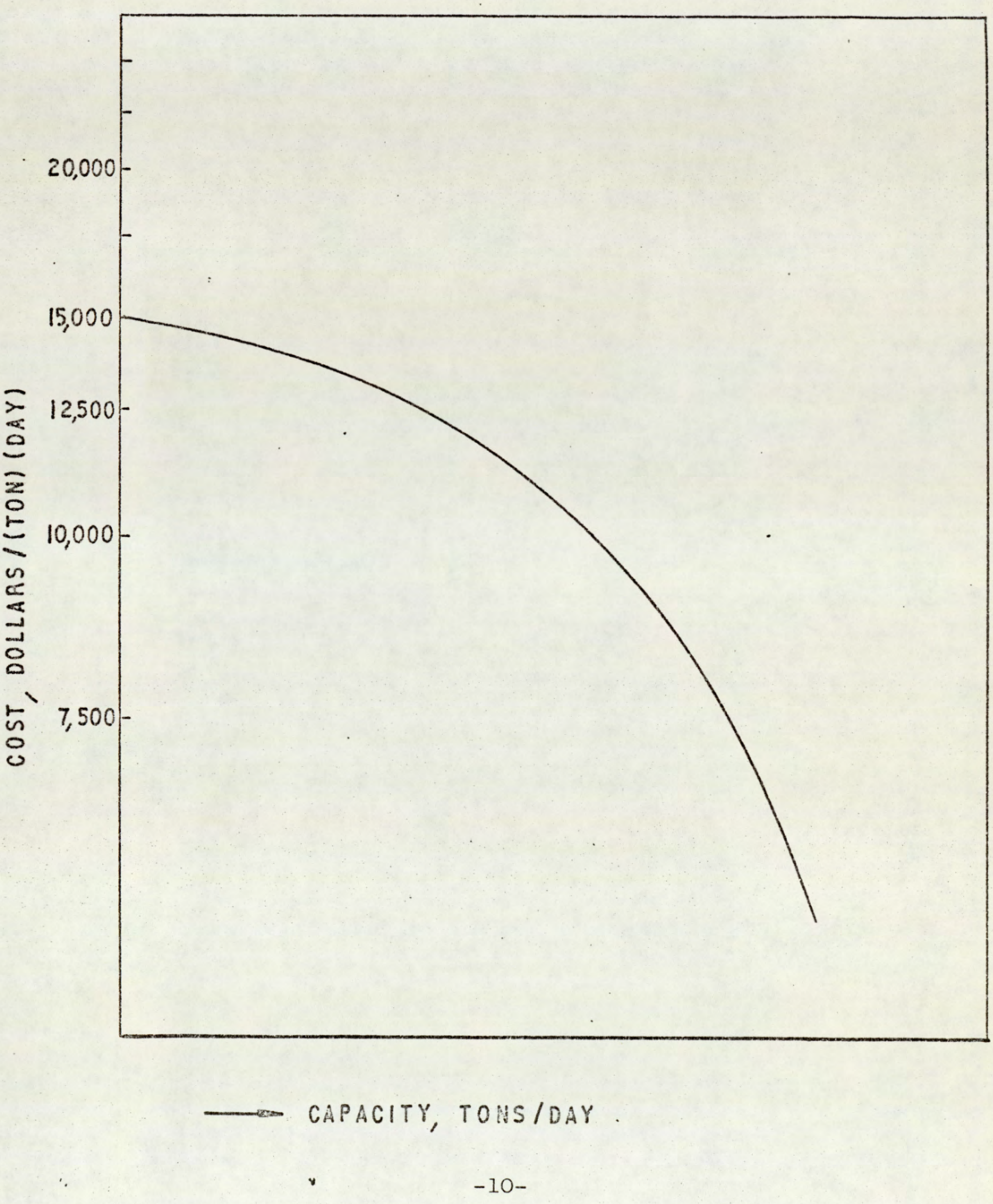
It is generally considered to be the least accurate of the order of magnitude methods. However, the concept of unit costs has been adopted in more sophisticated methods in which a unit cost is applied to unit operations and even to individual equipment items such as piping and instrumentation (see section 2.2.4.).

2.2.1.3. Exponential Method of Capacity Adjustment [9,10]

This is a well established method long known to industry. An account of its history and development is given by Estrup (14). The technique relates the fixed capital investment of a new process plant to the fixed capital investment of a similar previously constructed

FIGURE 2.2.1.2.1

UNIT COST v PLANT CAPACITY.



plant(s) by an exponential power ratio. The capacity adjustment is as follows:-

$$C_N = C_o . R^b \times I \qquad (1)$$

where C_N = estimated fixed capital investment of the proposed plant

C_o = historical fixed capital investment of a similar known plant

R = capacity ratio (proposed historical)

b = exponential factor

I = cost index.

A closer approximation for this relationship which involves the direct and indirect plant costs has already been proposed

$$C_N = f(D(R)^b + I) \qquad (2)$$

where f = lumped cost index factor relative to the original installation cost. This factor is the product of a geographical labour cost index, the corresponding area labour productivity index and a material and equipment cost index

D = Direct Cost

I = Total Indirect cost for the previously installed utility of a similar unit on an equipment site.

Earlier applications of equation (1) indicated that the exponential factor ranged between 0.6 and 0.7 for many types of process (10) and for this reason the method was often referred to as the two third's power law. The two third's phenomena was first noticed by Williams (15) and was later examined objectively by Bruni (16) who justified the law by noting that the fixed capital investment of a continuous production unit is proportional to surface area whereas output is proportional to volume, and that the ratio of the surface areas of two cylinders (equating to common equipment items), whose height is related to the diameter, is equal to the ratio of their respective volumes raised to the power two thirds.

The application of the two thirds rule of thumb for most cost-capacity curves is an over-simplification of a variable cost concept since the actual values of the cost exponent vary from 0.3 to 1.0 according to the process under consideration (10). The variation of cost with capacity of various sections within a process plant is also important (20). It should be noted that there can be a significant variation in the 'sectional' exponents within a plant and that the overall ratio for a particular plant may be comprised of widely diverging sectional exponents. This applies especially to complex, multi-product processes. Various exponents have been collated for different types of chemical plant and they are available in the literature (17,18,19). Unfortunately, there are some discrepancies among the exponent factors published which can be attributed mainly to variations in the definition of the scope and the size limitations associated with the respective exponents.

The limitations associated with the use of the exponential method of capacity adjustment are:-

(i) Historical data is required for its application. The data used has to be well defined.

(ii) The scale factors used in the method are only applicable over a certain capacity range and can vary for each process as a function of process size. For high capacities extrapolation may lead to estimation of plant of which some of the constituent plant items may dimensionally exceed the maximum that can be constructed and may involve stresses greater than those normally permissible for the materials employed. This would necessitate duplication of equipment and increasing scale up effects. Alternatively, for small capacities costs can remain fixed or vary little within a certain capacity range due to economics involved with purchasing and installing 'standard' size equipment and the fact that contractors probably have to quote relatively high fixed costs for smaller plants to cover overheads and incidentals which are almost the same in the case of larger plants. This acts to decrease the size exponent and improve the economy of scale.

In general, the method should not be used beyond a roughly five fold range of capacity (20) and for better results the cost of a similar plant within the desired capacity range needs to be known.

The technique is possibly more applicable to individual items of chemical plant rather than complete installed plants. Nevertheless, the method is quick and easy to apply provided sufficient data is available and hence is very popular with industry for this class of estimate. The theory and conclusions recognized for this method were later widely adopted as an integral part of future step counting methods.

2.2.1.4. Curve Pricing Method [9,10,13]

This technique consists of comparing the cost of the design under consideration with previous historical cost data that have been plotted for that design relating installed capital cost to plant capacity. This is a useful technique that allows rapid approximation of plant cost provided the curve is well defined and its limitations are recognized.

Some of the more common errors associated with the use of the curve pricing technique are:-

(i) it is assumed that the cost obtained from the curve is applicable to all plants manufacturing the same chemical whereas it may only refer to a specific process design.

(ii) the basis for the plotted curve is not always adequately defined and so leads to errors in its use. For correct use of the method the curves need to specify what cost they are defining, the cost time base, the process design and plant location so that any necessary adjustments may be made.

(iii) As with the exponential method of capacity adjustment, care needs to be taken when extrapolating the curve.

Even with these limitations however, it should be recognized that the curve pricing method if used correctly allows rapid approximation of plant costs, the accuracy of which is usually sufficient for preliminary economic studies.

2.2.2. Study Estimates

Study estimates of capital cost are directed at pre-design cost estimating situations and are usually prepared from limited flowsheet and process information. As such the techniques and methods within this classification are of particular interest.

Study estimating methods are used mostly for screening or comparing process design alternatives to enable the most economic design to be selected and for making initial assessments of project profitabilities. The error limits generally associated with study estimates are of the order of ±30% but it would be incorrect to apply this as a 'rule of thumb' for defining and identifying study estimating techniques. The methods in this category have been developed to fill the various estimating needs and situations found in the pre-design stage of process development and as such are derived from varying amounts and quality of information. As a result variable accuracies are expected and achieved within the study estimate class.

The following techniques are used for preparing study estimates and include the 'step counting' methods which are developed later.

2.2.2.1. Lang Factor Method

One of the earliest researchers into rapid capital cost estimating was Lang (21,22) who related total plant investment to delivered equipment costs using derived overall factors based on historical cost data analysis, such that

(DEC) Delivered Equipment Cost x Factor = Fixed Capital Investment.

Lang's factors varied with the type of process plant being considered and were concluded to be a function of material phases. The original factors proposed by Lang were:-

	3.10 for solid phase processes		Total Estimated Plant Cost (Fixed Capital Investment)
DEC x	3.63 for solid/fluid phase processes	=	
	4.74 for fluid phase processes		

Peters and Timmerhaus (10) give a table showing the derivation of factors for estimating the fixed investment for major additions to an existing plant and for estimating 'grass roots' investments. The factor values given are slightly different from those of Lang. Nicholls (23) showed that estimates made by Lang's method may vary from -30% to +20% of actual costs (see section 2.3).

The Lang factor method was an important addition to the field of capital cost estimation because the principle of 'factorial' estimating came to be widely supported and well established and is used for cost estimation at all stages of project development. However, when used for rapid cost estimation a number of well known limitations are known to exist (20):-

(i) Considerable information is needed to obtain delivered equipment cost, such as a fully detailed flow-sheet and equipment specifications. Such information is not usually available in the early stages of process development without much preliminary work to obtain individual costs of items to build up the bulk process equipment cost. This can be overcome to a certain extent using the unit cost approach as advocated by later workers (47,48)

(ii) Problems arise over the definition of delivered equipment cost.

(iii) The effect of plant size on Lang factor values needs investigation. Little information is available in the literature but the problem can be overcome by using individual factors related to equipment unit cost (24). However, this requires even more information and problems arise in defining equipment size.

(iv) The use of delivered equipment cost as a base for calculating installed project costs when materials of construction may be changed for plant items of identical size, (25) can give rise to error. This requires equipment costs in other materials to be reduced to a mild steel equivalent basis, then multiplying by the appropriate factor(s), finally adding the incremental cost of the special materials.

(v) The validity of the factors derived by Lang has been questioned by later workers. (10), (26). With time the factors may be expected to change with time (they were produced in 1948) due to technology advancements and changes in standards. There is little information available on the extent of the change.

(vi) The factors take no account of market effects on delivered equipment costs. If competition amongst vendors is fierce then delivered equipment costs can be significantly lower than 'normal' and application of Lang factors to these costs can result in low estimates. The opposite applies if competition is light and equipment vendors have full order books.

It can be seen then that the Lang factor approach at its simplest is fraught with problems and uncertainty.

However, realizing that the factorial concept proposed by Lang for capital cost estimation was an important and valid one, later workers in this field set out to overcome some of the problems associated with his method (see section 2.2.2.2.).

2.2.2.2. Modified Lang Factor Methods

Later workers proposed that greater accuracy of capital investments could be achieved by using not one factor as in Langs method, but a number of factors or detailed factoring. Many developments of the Lang factor concept have been published such as:

(i) The use of different factors for different types of equipment. Hand (27) compiled a series of factors for specific equipment types related primarily to the petroleum industry (see Table 2.2.2.2.1.). The ratios apply to mixtures of materials of construction that consist of mainly carbon steel (with relatively small amounts of alloy materials), but are not applicable to all alloy systems. His explanation in this area was somewhat vague. Clerk (28) published a more general approach which allows for differences in materials of construction. Both point out that items built on site need separate factor status.

(ii) The use of separate factors for erection of equipment foundations, piping, utilities etc. or even to break up each item of cost into material and labour

TABLE 2.2.2.2.1

HAND'S FACTORS RELATING TOTAL BATTERY LIMITS COSTS TO EQUIPMENT COSTS (27)

Fractionating Columns	4
Pressure Vessels..............	4
Feed Exchangers...............	3½
Fixed Heater..................	2
Pumps.........................	4
Compressors...................	2½
Instruments...................	4
Miscellaneous Equipment.......	2½

factors. With this approach each factor has a range of values and experience is needed to judge for each case whether to use a high, average or low figure. Bach (26), Gilmore (29) and Hackney (30) for example have put forward various ranges of values for these factors.

(iii) Hirsch and Glazier (31) developed an equation for the estimation of new capital investment which combined the separate cost factors

$$C = f_I \, E \, (1 + f_F + f_P + f_m) + E_i + A$$

where the various parameters are defined accordingly:-

C = fixed capital investment

E = purchased equipment cost on f.o.b. basis

f_I = indirect cost factor representing engineering, contractors overheads and fee, supervision and contingencies.

f_F = cost factor for field labour.

f_P = cost factor for piping materials.

f_m = cost factor for miscellaneous material costs such as insulation, instruments, foundations, structural steel, electrics, painting, building and the cost of freight.

E_i = installed equipment cost.

A = incremental cost of alloy materials.

The installation cost factors are defined by the following equations:-

$$\log f_F = 0.635 - 0.154 \log 0.001E - 0.992 \frac{e}{E} + 0.506 \frac{fv}{E}$$

$$\log fp = -0.226 - 0.014 \log 0.001E - 0.156 \frac{e}{E} + 0.556 \frac{P}{E}$$

$$\log f_m = 0.334 + 0.033 \log 0.001E + 1.194 \frac{t}{E}$$

where e = total heat exchanger cost.

f_v = total cost of field fabricated vessels

P = total pump + driver costs

t = total cost of tower shells.

The estimating equation was derived from an analysis of cost data kept by the author's company. The various equations have been combined into an easy to use monograph by Walas (32).

Hence, it can be seen that Lang's original concept of factorial estimating has been subsequently well developed since its introduction. Other techniques which may be described loosely as modified Lang factor methods have been developed, namely those of Wadell (33) and Miller (34), who developed a technique which combined factoring and step counting theory. For reasons explained in section 2.2. and 2.2.3. these two methods have been classified as preliminary class estimating techniques and as such are discussed under that heading.

To summarise the section of modified Lang factors then the following points are noted. The methods described are well established and used widely in the process industries to produce varying classes of cost estimate depending on the sophistication of the method employed. However, they require significant amounts of detailed information in order to use them which in many cases makes them unsuitable in situations of minimum process information and so limits their use in the early stages of process development. Also, considerable experience is needed in selecting and adjusting detailed component factors for a given situation which limits their use to experienced engineers. It is largely because of these limiting constraints that step counting methods come to be derived and achieve prominence in the field of rapid pre-design capital cost estimation. Such methods have the advantage of calculating total installed costs directly without the use of detailed factoring, and from little process information. They are ideally suited to the requirements and constraints which exist in the early stages of process development. Step counting methods are reviewed next and their development described.

2.2.2.3. Hill's Method

Hill's method was derived to estimate the capital cost of low pressure fluid petrochemical processes (35). The method does not require design calculations or detailed cost data and is the first published record of the step counting concept for rapid, pre-design cost estimation. Hill states that his method is quick and easy to use, and gave results accurate to within ± 25% in 90% of over 20 low pressure processes estimated, with a maximum deviation of 40%.

In the method major items of process equipment are classified as being equivalent to one or two 'units' depending on their complexity, type and construction. Simple major equipment items such as carbon steel towers and low pressure reactors are classified as one unit. More complex and expensive items, such as stainless steel towers and furnaces, are classed as two units. Table (2.2.2.3.1.) shows Hill's classification of various pieces of major equipment. One unit is added for each liquid feed material and liquid product that requires storage. Two units are added for each solid or gaseous product. For equipment with pressures over 100 psi, the number is multiplied by working pressure (psi)/100. For example, a 300 psi carbon steel reactor would be rated as three units. Hence, the total number of units in the petrochemical process was determined.

Hill calculated that for a petrochemical plant producing 10 million lb/year of total product, each unit had an installed cost of US$ 30,000. This figure was based on a Marshall and Steven's equipment cost index of 185 in 1954-55. The total installed equipment cost for a 10 million lb/year plant can thus be calculated as the number of units, N, multiplied by US$ 30,000. Hill suggested that the installed equipment cost for plant capacities other than his base case of 10 million lb/year should be calculated using a capacity adjustment exponent of 0.6.

TABLE 2.2.2.3.1

HILL'S MAJOR PIECES OF EQUIPMENT CLASSIFICATION (35)

One Unit	Two Units
Carbon Steel Tower	Stainless Steel Column
Reactor	Furnace
Evaporator	Centrifuge
Blower	Compressor
Precipitator	Refrigeration Unit
Liquid Feed Storage	Solid Product Storage
Liquid Product Storage	Gaseous Product Storage

The total fixed capital investment is then estimated from the installed equipment cost by adding percentages/ factors as proposed by Chilton (36) to include piping, instrumentation, buildings, auxilliaries and other costs. These percentages are shown in Figure (2.2.2.3.2.). The total fixed capital investment is finally updated to the present day by use of the current Marshall and Stevens equipment cost index, such that

$$\text{US \$ } C = IEC \times F \times CCI$$

where C = estimated fixed capital investment

IEC = installed equipment cost

F = appropriate factors to account for piping, buildings, electrics, instrumentation etc.

CCI = construction cost index (Marshall and Stevens)

Hill's method proved to be an important addition to the area of pre-design cost estimating. The technique provided a short cut method for obtaining the installed equipment cost. Total unit installed cost was then derived using appropriate factors for the non-equipment items. Although Hill recognized that the cost of units could be averaged, he still preserved two levels - simple and complex. It is not difficult to see Hill's 'units' being developed into other, more sophisticated forms. Disregarding the two levels, if the factors were applied to each 'unit' before summing, this would effectively give the total capital cost of each unit and the method would become a version of the step counting/functional unit

TABLE 2.2.2.3.2

PERCENTAGE FACTORS FOR USE WITH HILL'S METHOD (35)

Item	Percentage of Installed Equipment Cost
Processing Piping	10 to 40
Low - solids plant	10
Average - solids/fluids plant	25
High - fluids plant	40
Instrumentation.	5 to 15
Low - little or no automatic controls	5
Average - some automatic controls	10
High - centralised complex controls	15
Manufacturing Buildings.	0 to 80
Minimum - installation in existing buildings	0
Low - outdoor construction	15
Average - mixed outdoor and indoor	40
High - indoor construction	80
Auxiliaries	0 to 75
Minimum - existing facilities adequate	0
Low - minor additions needed	5
Average - major additions needed	25
High - complete new facilities	75
Outside Lines.	0 to 20
Low - close integration	0
Average - separate processing units	10
High - scattered processing units	20

Sum of Installed Equipment Cost and the items listed above

= Total Physical Cost.

TABLE 2.2.2.3.2 (continued)

Item	Percentage of Total Physical Cost
Engineering & Construction.	30 to 40
Low - straight forward engineering	30
High - complex engineering	40
Contingencies	10 to 40
Low - firm process	10
Average - subject to change	20
High - speculative process	40
Size Factor	0 to 25
Low - large commercial unit	o
Average - small commercial unit	10
High - experimental unit	25

Sum of Total Physical Cost and above items

= Total Plant Cost

approach. Alternatively, by a change in definition and emphasis each 'unit' could become a main plant item which leads to the module or unit estimating approach.

Hill's technique might therefore be seen as the point of divergence and consequent development of two techniques of capital cost estimation and his proposal thus assumes a high significance.

Apart from providing a new, workable concept which has subsequently been refined and developed by later workers, Hill's method has justified his claims for it and it has proved to be a practical, working method. It is quick and easy to use and a recent analysis found his accuracy claim of ± 25% to be justifiable (20), (48).

Specific advantages noted for the method are:

- It requires very little process information; only a process flowsheet, process capacity, operating pressures and major equipment items.
- Estimation of the total number of units is simple, provided there is no deviation from the examples provided.
- The method does indirectly incorporate some allowance for material of construction effects.
- Percentage factors are used for conversion of installed equipment cost to total fixed capital investment.
- At the same time however, limitations of the method have been identified, namely:-

- The method makes no allowance for temperature effects on cost. As the method is based on low pressure processes it seems probable that it was also based on 'moderate' temperatures (say 0-500°C). Hence, any high temperature process may be under estimated.

- Hill's method is based on the annual output capacity of the main process stream. It is thought that this is not a suitable means of representing process 'size' which is best measured by throughput of the process plant units.

- Hill did not publish or explain his cost data base, either the dates or geographical origin.

- The method is now over twenty years old (published in 1956) and considerable change in construction methods has taken place in that time which may make his standard 'unit' cost of $30,000 incorrect (after allowing for inflation), under prevailing conditions, due to learning effects and changing standards.

- Hill states that the method is applicable to processes operating at pressures of under 100 psia. However, he makes no mention of units operating under vacuum conditions. Page (48) in fact emphasised that the method was clearly not applicable to high pressure processes.

2.2.2.4. Zevnik and Buchanan's Method

Zevnik and Buchanan developed an estimating technique for fluid type chemical plants (37) based on the supposition that the fixed capital investment is a function of two variables; process complexity and process capacity. The results they obtained were inconsistent, although the general accuracy levels they claim (±25%) are well within the acceptable limits for the intended application of the method.

Their method is reasonably quick and easy to apply, involving only a comparatively small amount of work, and requires relatively little process information. Their correlation requires only six items of input information:-

- Capacity of the proposed process plant (Q)
- Number of process steps or 'functional units' in the process plant (N). They define a functional unit as 'all the equipment necessary to carry out a single significant process function'.
- Process plant complexity (CF), which allows for cost variations due to materials of construction, operating temperature and operating pressure.
- Construction cost index (CCI).

Their method is essentially a graphical technique, the estimating procedure being as follows:-

1) Assessment of process complexity (CF) which is defined as

$$CF = 2 \times 10^{(F_T + F_p + F_a)}$$

where F_T = temperature factor

F_p = pressure factor

F_a = alloy factor

Figures (2.2.2.4.1) and (2.2.2.4.2) show the graphs published by Zevnik and Buchanan from which the temperature and pressure factors are determined. The graph assessing the temperature factor, F_T, may be numerically represented as (20);

$$F_T = 1.8 \times 10^{-4} \ (T-300), \text{ where } T = \text{maximum process temperature } (^oK)$$

for above ambient temperatures and,

$$F_T = 0.57 - (1.9 \times 10^{-2} T)$$

for when sub-ambient temperatures apply.
The pressure factor, F_P, may be mathematically represented as

$$F_P = 0.1 \log_{10} P, \text{ where } P = \text{maximum process pressure (atm. abs).}$$

for above atmospheric pressures and,

$$F_P = 0.1 \log_{10} (1/P_{MIN})$$

for below atmospheric pressure.
Table (2.2.2.4.1) gives the alloy factor for various materials of construction.

2) The fixed capital investment per functional unit (CPF) is obtained from a graph relating cost, plant capacity and the process complexity factor. The graph is illustrated in Figure (2.2.2.4.3).

For capacities above 10 million lb/annum the cost v capacity curves are based on the traditional 'six-tenths' rule. (38)

$$\$ \ CPF = 6000 \ (Q)^{0.6} . \ (CF)$$

FIGURE 2.2.2.4.1

ZEVNIK AND BUCHANAN'S TEMPERATURE FACTOR CURVE[37]

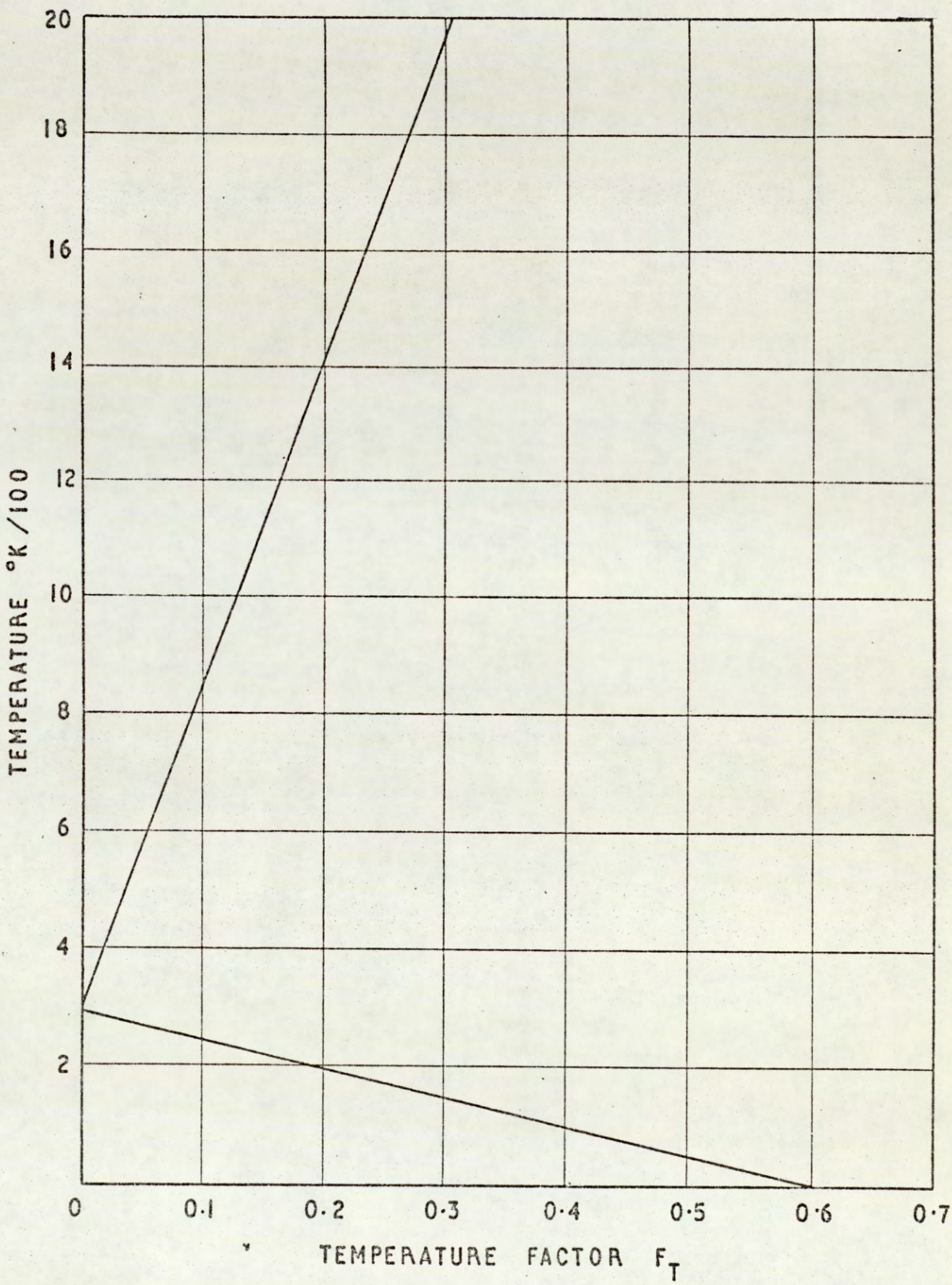


FIGURE 2.2.2.4.2

ZEVNIK AND BUCHANAN'S PRESSURE FACTOR CURVE (37)

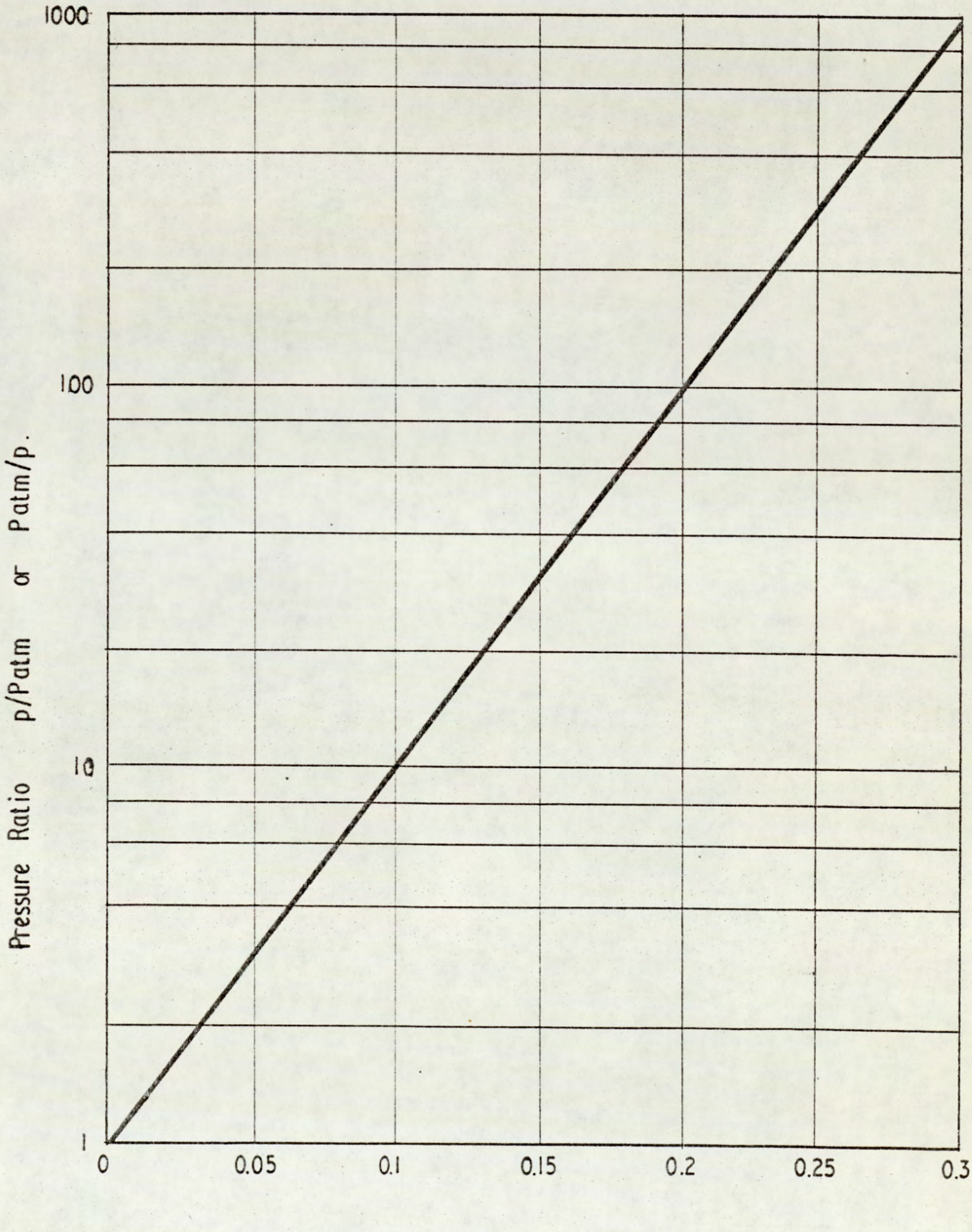


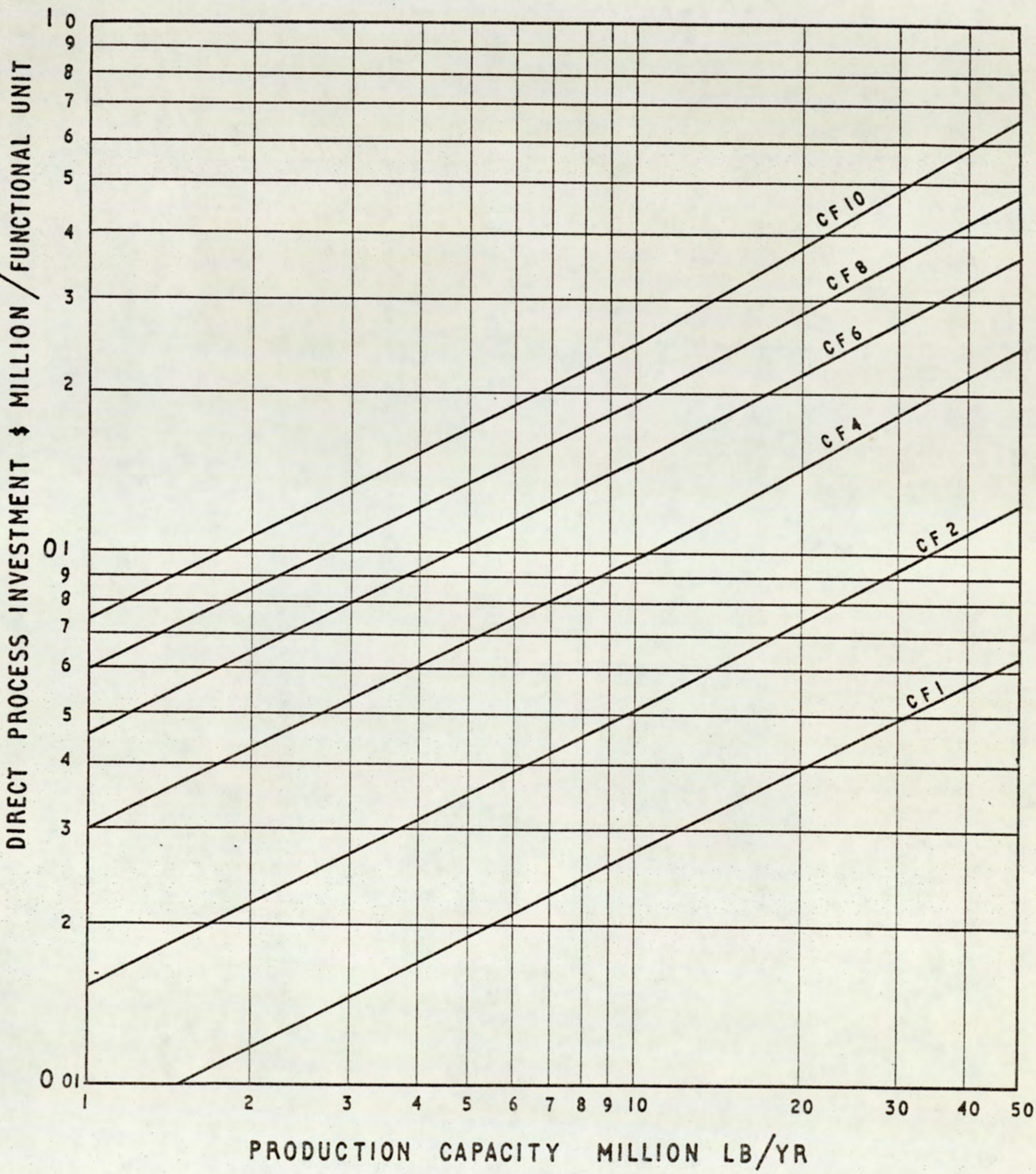
TABLE 2.2.2.4.1

ZEVNIK AND BUCHANAN'S ALLOY FACTORS (37)

Alloy Factor, F_a	Construction Material
0	Cast iron, carbon steel, wood
0.1	Aluminium, copper, brass, stainless steel (400 series)
0.2	Monel, Nickel, Inconel, stainless steel (300 series)
0.3	Hastelloy, etc.
0.4	Precious metals

FIGURE 2.2.2.4.3

ZEVNIK AND BUCHANAN'S COST/CAPACITY CORRELATION[37]



For capacities less than 10 million lb/annum it was considered that an exponent of 0.5 was more suitable.

$$\$\ CPF = 7600\ (Q)^{0.5}.\ (CF)$$

For easier comparison with other correlations the capacities if given in long tons/annum change the constants to 154 and 360 respectively.

3) The fixed capital investment for a process plant is estimated from the following equation

$$I = (N) \times (CPF) \times (1.33) \times (CCI/300)$$

where I = direct process investment (i.e. fixed capital investment)

N = number of functional units

CPF = cost per functional unit

CCI = construction cost index.

The multiplying factor of 1.33 is an arbitrary allowance proposed by the authors for the cost of allocated utilities and general facilities.

4) In establishing their correlation Zevnik and Buchanan used as a base value an Engineering News Record Construction cost index of 300 (base 100 in 1939). The respective Engineering News Record Index (CCI) divided by 300 is utilized to update the fixed capital investment to the present day.

A number of limitations have been noted in applying Zevnik and Buchanan's method. Without wishing to detract from the value of their work, it would be helpful to identify the shortcomings of their proposal:

- Their definition of a functional unit is vague and leads to errors in estimating N. A clearer understanding of the term is important. Page (48) clearly found the concept difficult as he managed to identify seven different types of functional unit before rejecting the approach in favour of his own alternative.
- Their 'capacity' is based on the main process stream output. Later workers in the field have more or less unanimously agreed that the size and hence cost is more likely to depend on throughput rather than output. A process of low conversion and high degree of recycle would be expected to cost more than a process of comparable output with little or no recycle.
- It is thought that volumetric rather than mass measurement units would be more effective in measuring plant capacity (see Chapter 3, section 3).
- The use of maximum process temperature and pressure may be misleading. For example, a process where only one unit is operating at the extreme temperature and/or pressure is unlikely to be as sensitive to these conditions as a process where most of the units are operating at or near the extreme.
- Their correlation is restricted to fluid type processes (gas and gas/liquid phase systems). However, it is important to note that the dominant phase of a process is clearly important after considering the points above. (see Chapter 3, section 1).

- . The construction cost index used in their correlation is more suited to civil construction projects than to chemical plant construction (see Chapter 4, section 4.1.2.2).
- . No contingencies for accounting for learning effects are mentioned.
- . The 1.33 utilities factor is proposed for all plants. This is too crude an approximation.
- . They do not state the origin of their cost data and reported difficulty in obtaining real and reliable costs. Derivation of realistic cost estimation correlations is strongly dependent on reliable data.
- . They do not define the bases for their three 'F' factors. No account was given of the assumptions they made or the methods and standards they used in their calculations to arrive at the F_T, F_P and F_m values.

Page (48) questioned the derivation of the complexity factor. When analysing their cost per functional unit v capacity graph with complexity factors as parameters (Figure 2.2.2.4.2) he highlighted a notable discrepancy. The authors plotted lines on the graph for factor values 1,2,4,6,8 and 10. Values which fell between these integers had to be interpolated. In the base, when the plant was operating at atmospheric pressure and normal temperature and constructed from cast iron the F_T, F_P, F_m values were all zero (i.e. their minimum value). Hence, the minimum value for Zevnik and Buchanan's complexity factor was 2 (i.e. 2×10^0). How then had they managed to plot a line on a graph for a factor value of 1? The fact that all the lines on the graph were parallel leads one to suspect perhaps that several of them had been obtained by speculation rather than calculation. From their equation for complexity factor a value of one was impossible but, the authors still managed to construct a line on their investment graph corresponding to it.

This tended to cast doubts on the validity of the graphs as a whole.

Recent attempts to use Zevnik and Buchanan's method have proved relatively unsuccessful (20), (48) with accuracy rarely exceeding ^{x} 1.7 (48).

2.2.2.5. <u>Gore's Method</u>

Gore's work (39) resulted from an examination of Zevnik and Buchanan's earlier work on the functional unit concept of cost estimating. His study was restricted to processes where the feedstock and product was gas. He was concerned mostly with the definition of functional unit, and also use of throughput rather than capacity on a volumetric rather than mass basis. His capital cost correlation was derived from a regression analysis of 65 sets of published cost data for 11 different processes. Accuracy of $\pm$ 25% is associated with his technique (2).

Gore's correlation for gas-phase based processes is as follows:-

$$C = 4680\ (N).\ (Q)^{0.62}\ (T)^{X}\ (P)^{0.395}.\ Fm.\ \frac{ENR}{400}$$

where C = battery limits capital cost (US $)

N = number of functional units

Q = average throughput (million lb moles per year), which was obtained by multiplying the process capacity by a 'recycle factor' which was derived empirically.

T = temperature factor = $\left(\frac{T_{max}\ {}^{o}K-300}{300}\right)$

X = 1.07 approximately, but is a function of Q such that $X = (Q^{0.206})/2.52$

P = pressure factor = P_{max}(atm)

Fm = materials of construction factor, which was not evaluated as there did not appear to be any significant difference in the plants considered.

ENR = ENR cost construction index, base 100 in 1913.

The estimating procedure is as follows:-

1) Calculate N, the number of functional units in the process. Gore defined a functional unit as (38) -

a) "that equipment which is necessary to achieve a chemical or physical transformation of the major process stream, and is consistent with the equation:-

Output = f(input)".

and/or, alternatively

b) "a significant piece of plant which carries out an operation in the main process stream. It is usually represented by an individual block in an initial flow diagram. Examples include stills, absorbers and other significant unit operations, but exclude pumps, heat exchangers, reboilers and items which are subsidiary to a unit operation or unit process.

Storage tanks and hoppers are excluded. Pieces of equipment that carry out mechanical separation only count as a functional unit if they constitute substantial systems in their own right, and cannot reasonably be built into a unit operation. Such equipment would include crushers, centrifuges and rotary vacuum filters but not cyclones or simple gravity settlers without mechanical gear. Mechanical items for feeding and discharging complete systems count as a functional unit, as do heating or cooling main process streams where the heat load is excessive or substantial or the media employed is unusual. Heat transfer equipment which is dependent on local economics is excluded, but waste heat boilers, quench towers and chequer systems are included. Multi-stage operation (as in a multi-effect evaporator) is counted as a single functional unit".

2) Calculate Q, the average throughput:-

Gore calculated throughput by multiplying the process capacity by a 'recycle factor' which was empirically derived for the processes he studied. Alternatively throughput could be calculated by averaging the following two quantities

Number of lb moles of feed reactants and recycle.

Number of lb moles of product and recycle.

If a detailed mass balance is not available then the first expression may be satisfactorily used.

3) Calculate temperature factor and pressure factor as previously shown.

4) Calculate material of construction factor:-

Whilst Gore recognized that material of construction was a significant factor influencing capital cost he did not develop any factors for use in his correlation, and assumed similar materials of construction for all his data. i.e. $F_m = 1.0$.

5) Modifications to the constant:-

Although Gore made no mention of varying constants in the initial thesis later modifications were suggested (48) to give more realistic results that the 4680 published. It was noted that the constant changed as a function of throughput (see Table 2.2.2.5.1).

6) Escalation of the cost to present day using the ENR construction cost index.

Gore himself stated that the following limitations existed when using his correlation:-

The correlation is only valid for well established gas-phase processes. A learning allowance may have to be made for novel processes.

If N, the number of functional unit is less than four then the estimate should be treated with great caution.

TABLE 2.2.2.5.1

GORE'S CORRELATION CONSTANTS AS A FUNCTION OF PLANT SIZE

Q (million lb mols/annum)	k
1-5	20,000
5-10	15,000
10-50	12,500
50-100	10,000
100-500	7,500
500-1000	5,000

The correlation is valid for the following constraints:

$10 < Q > 50$ million lb mols/annum

$450 < T_{max} > 1250°C$

$1.5 < P_{max} > 300$ atm.

Other disadvantages about the method have been noted about the method however, namely,

. His definition of a functional unit is vague and difficult to interpret and so leads to possible errors.

. Calculation of throughput poses a problem. Either detailed mass balance data is required or the recycle factor needs to be calculated for the process. Gore was vague about this part of his work and no definition of recycle factors or procedures to calculate it were given.

. Gore only uses 11 processes in deriving his correlation. Published cost data was used. Gore was conscious of the need to use consistent data (he screened his cost data to separate reliable and unreliable information) and it is likely that the data source is good enough in quality and quantity to give reliable results.

. Gore's use of the ENR cost index to update costs is not suitable for process plant cost escalation, and is more suited to civil construction applications.

However, it should be noted that Gore's study did provide some useful contributions in the area of rapid cost estimating. Although it is difficult to use his method, his study did provide some interesting information, notably that the throughput term apparently agrees with the accepted 'six tenth's' or 'two thirds' power law, and that the temperature and pressure effects are significant as might be expected for gaseous phase processes, and seem to affect the capital cost more than is suggested by Zevnik and Buchanan's proposal.

Gore's method is considered to provide a more accurate result than Zevnik and Buchanan, but requires a little more information and considerably more care in its use. It's major limiting factor is that it needs the total feed to the process to be known. It is doubtful that this parameter may always be available in the early stages of process development, and so Gore's objective of producing a correlation from 'minimum process information' may be defeated.

2.2.2.6. Bridgwater's Method(s)

Bridgwater has been concerned with the development of the functional unit concept of cost estimating for several years and has produced a series of cost correlations.

He has made a significant contribution to this area of cost estimating since his methods are the only ones available for estimating solid and/or solid/liquid phase processes and so filled an obvious need in this area of study.

His initial study was developed as an offshoot of a larger study, directed at hydrometallurgical extraction processes (2). Employing the same principles as Gore, though with some modifications, a cost correlation was developed via a regression analysis of 16 processes (24 sets of cost data). Accuracies of $\times\!\!\div$ 1.2 are claimed by the author.

The correlation is as follows:-

$$C = 50.26\ (N)\ \left(\frac{Q}{S^{0.5}}\right)^{0.85}\ \left(T.\frac{n}{N}\right)^{-0.17}\ \left(P.\frac{n'}{N}\right)^{0.14}.\ F_m.CCI.$$

where C = grass roots capital cost (UK £)

N = number of functional units

Q = plant capacity, long tons/annum

S = "conversion" factor

T = maximum process temperature, ^{o}C

n = number of functional units with temperature $T_{max}/2$

P = maximum process temperature, atm.

n' = number of functional units with pressure $P_{max}/2$

F_m = material of construction factor

CCI = construction cost index (ENR base 100 in 1913)

The estimating procedure is as follows:-

1) Determination of N, the number of functional units (as per Gore's definition)
2) Calculation of throughput - as a function of plant capacity (Q) and reactor conversion (S) such that -

$$\text{Throughput} = \frac{Q}{S^{0.5}}$$

where Q = plant capacity, expressed in long tons/annum.

NB. weight based measurement concluded more suitable than volumetric, which would cause difficulty when considering complex materials of unknown composition or molecular weight.

S = conversion factor, defined as

$$\frac{\text{weight of desired reactor product}}{\text{weight of total reactor input (or output)}}$$

3) Calculation of the temperature effect - via a weighting approach, since he considered that the use of T_{max} was inaccurate.

4) Calculation of the pressure effect - using the same approach as above.

5) Calculation of the material of construction effect. Since all his original processes were ones containing aqueous acidic materials he was not able to examine the effect of materials of construction initially. He later stated however that non-acid processes would cost about 20% less.

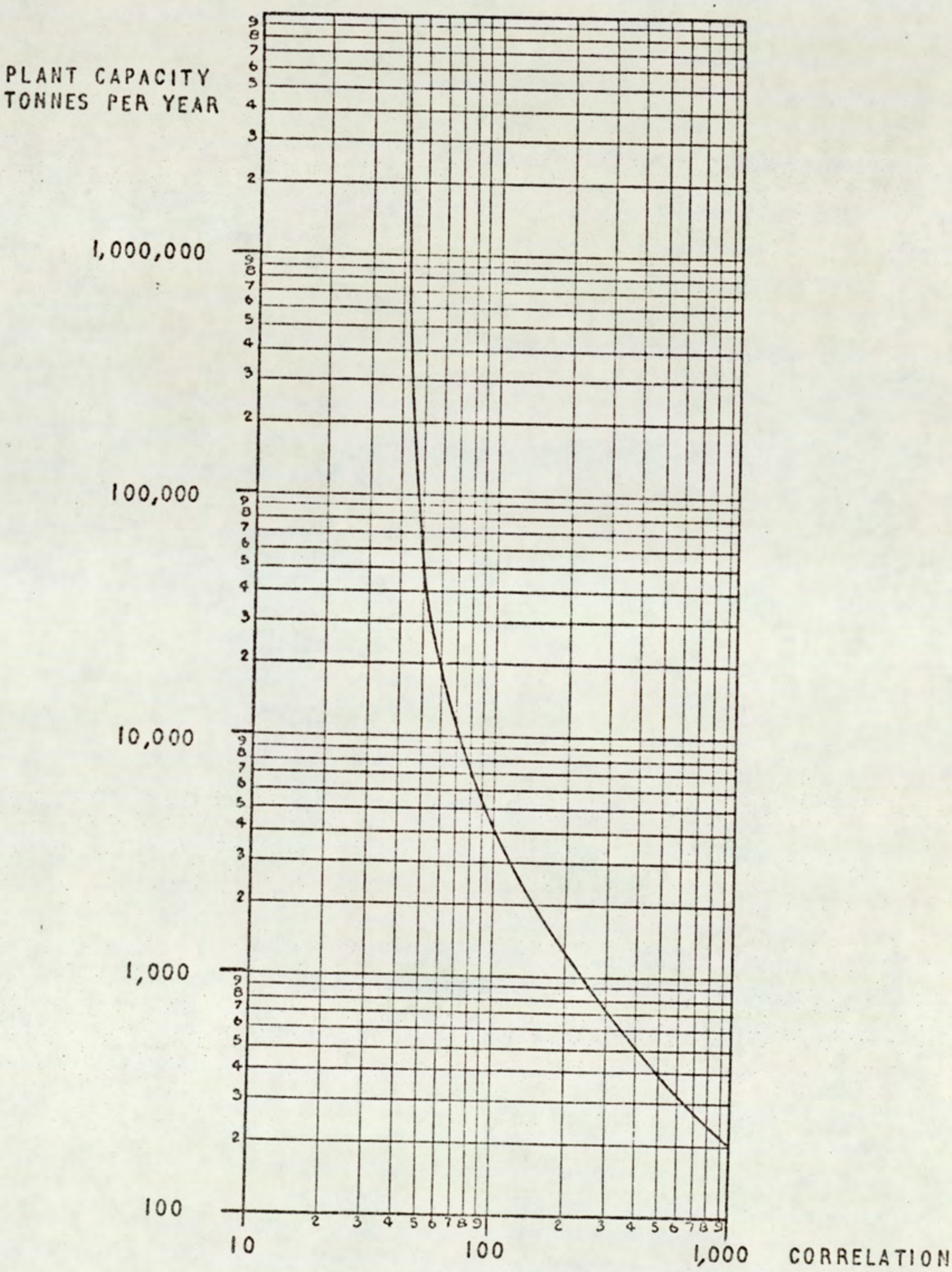
6) Modification of the constant value - which was concluded to vary as a function of plant capacity. The following modifications were proposed:-

- for small processes, Q   1000 ltpa, K = 400.
- for medium sized processes, $1000 < Q > 10{,}000$ ltpa, K = 140.
- for large scale processes the initial constant of 50.26 was maintained.

Bridgwater later published a graph showing the variation of the constant with size (see Figure 2.2.2.6.1)

GURE 2.2.2.6.1.

BRIDGWATER'S CORRELATION CONSTANT AS A FUNCTION OF PLANT SIZE.[2]



Bridgwater placed the following constraints on his correlation:-

It should only be used for solid and/or solid/liquid processes.

It is valid only for processes operating at above ambient temperature and pressure and with a capacity of $>$ 1000 long tons/annum.

He states that the method gives greenfield site costs, but this was difficult to determine due to problems with the data.

Apart from these constraints however, others are noted, namely:-

a) Like Gore, problems arise in determining the functional unit definition.

b) Difficulties can arise when using the conversion factor S to calculate process throughput since mass balance data is needed to define S (though only for the reactor).

c) Although the weighting approach used to define temperature and pressure effects is a progressive step, the representation of the temperature variable should be treated with caution. The negative value of -0.17 seems suspect and counter to intuition, although the author postulates that it may be due to temperature increasing reaction rate and so reducing cost. However, theoretically, it can give illogical cost estimates.

For example, if it is required to increase a process of N units by one unit (say a high temperature incinerator) the cost of the process will increase. However with Bridgwaters correlation this would not be the case. Since n (and possibly T_{max}) will increase then, $(T_{max} \frac{n}{N})^{-0.17}$ will decrease, thereby reducing the estimated cost. Hence caution is required in this area.

d) Like Gore, Bridgwater initially uses the unsuitable ENR construction cost index to escalate costs. However, later methods recognise this limitation and this index was replaced in future correlations.

Details of Bridgwaters later correlations can be found in the literature (20). They cover a wide variety of applications such as waste disposal and solids refuse processes. However, they will not be covered here since the principles contained in them have been covered in the previous description.

2.2.2.7 Stallworthy's Method

Stallworthy published a paper in 1970 (41) in which he outlined the most sophisticated development of the step counting, or functional unit, approach published to date.

His method was derived from an analysis of his company's (Courtaulds Ltd.) cost data which had taken him several years to collect (48). The plants he considered were nearly all fluid petrochemical processes with limited amounts of gas sections. Individual items of equipment were cost analyzed rather than whole plants. For each piece of equipment approximately 100 cost samples were taken. The maximum operating temperature and pressure considered were 1500°C and 200 atmospheres respectively. Stallworthy states that he was only interested in developing a quick costing method for use within his own company and which was intended to have a limited range of application to his company's plants. He had no intention of the method being used universally (42).

Stallworthy's method was really a modification of Zevnik and Buchanan's step counting method. However, instead of considering a process as a simple sequence of units with an average capacity or throughput, he developed a correlation based on a study of each stream - mainstream, recycle streams and side streams. This overcomes the assumption that the flows are constant throughout the process, but means that considerably more information is required (practically a detailed mass balance). The correlation is still based on plant capacity however, and the mass relationships of each stream to the main stream and hence capacity. Stallworthy proposed the following equation for the estimation of process plant capital investment.

$$\pounds C = \frac{0.00075}{A} \sum_{i=1}^{S} (N \times F_m \times F_p \times F_T \times R)_i$$

where C = estimated battery limits capital cost of plant

S = number of main, recycle and side streams

R = ratio of stream to main stream on a weight basis

N = number of significant process steps in the main stream or side stream

F_m = factor for the specific materials of construction

F_p = factor for design pressure

F_T = factor for design temperature

A = size factor for the capacity of the plant required.

Page (48) attributes a graphical method for obtaining temperature and pressure factors to Stallworthy which resulted from a private communication. These graphs appear in an identical form in Wilson's paper (47) which he also attributes in part to a private communication from Stallworthy. The graphs are shown in Figures (2.2.2.7.2) and (2.2.2.2.7.3). The temperature factor, F_T, can vary between 1.0 and 1.5. The pressure factor may vary between 1.0 and 1.3. The material factor, F_m, may vary between 1.0 and 2.0. F_m values are provided by both Page and Wilson and are shown in Table (2.2.2.7.1). The base case where all are 1.0 is for a 'mild steel' plant operating at normal pressure (0-100^{o}C) and pressure (1-100 psi). The cost/size factor, A, is shown in Figure (2.2.2.7.1), and is derived graphically. From the graph,

$$A = 7.30 \times 10^{-7} \times Q^{0.66}$$

where Q = plant capacity, long tons per year (2).

Stallworthy stated in his paper that his method gives results accurate to within ± 15% of the detailed estimates made for the same process. A recent study however (48 ,49) indicates that Stallworthy's method gives an accuracy of

FIGURE 2.2.2.7.1

STALLWORTHY'S COST/SIZE FACTOR[41]

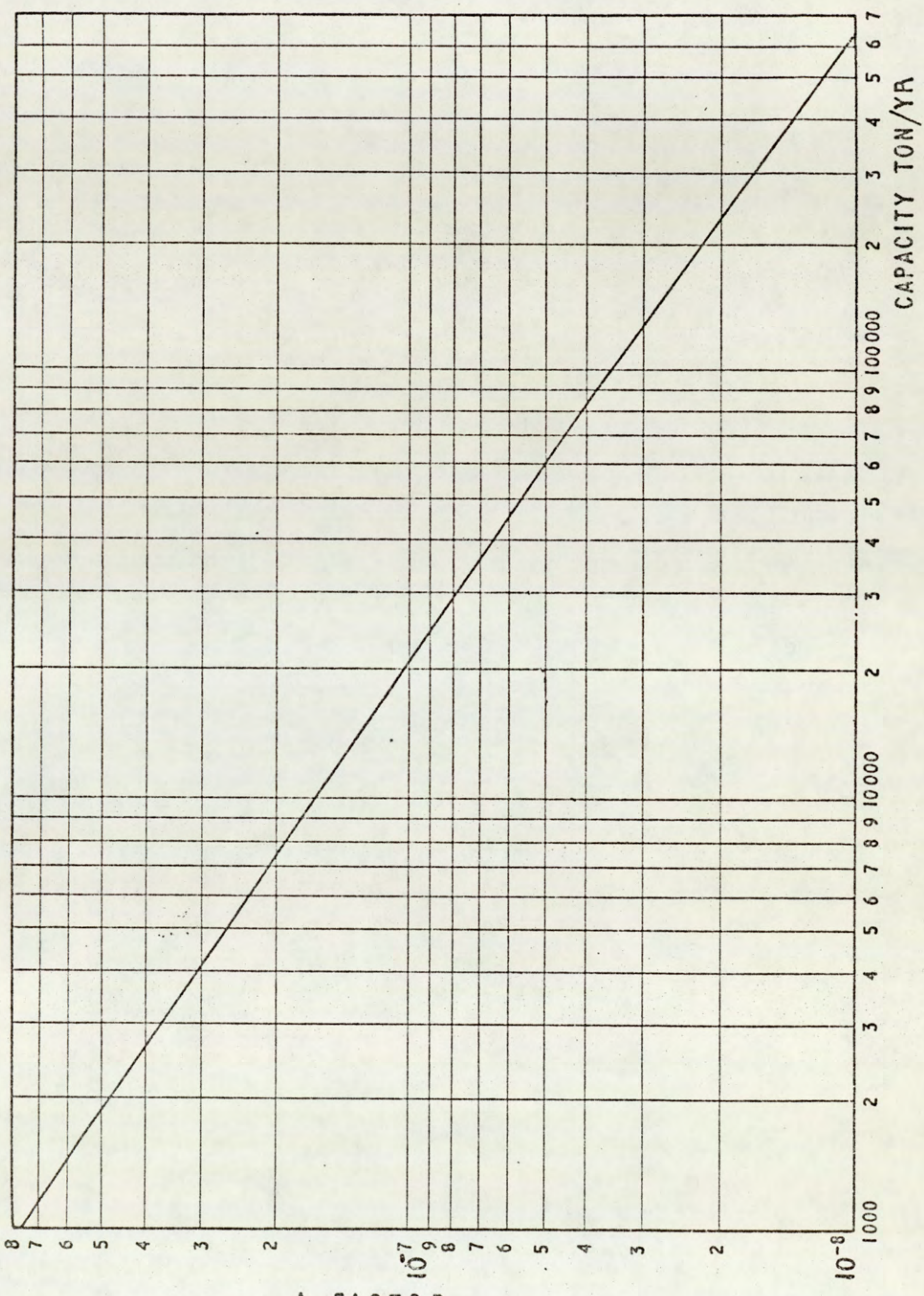


FIGURE 2.2.2.7.2.

TEMPERATURE FACTOR[41] (STALLWORTHY)

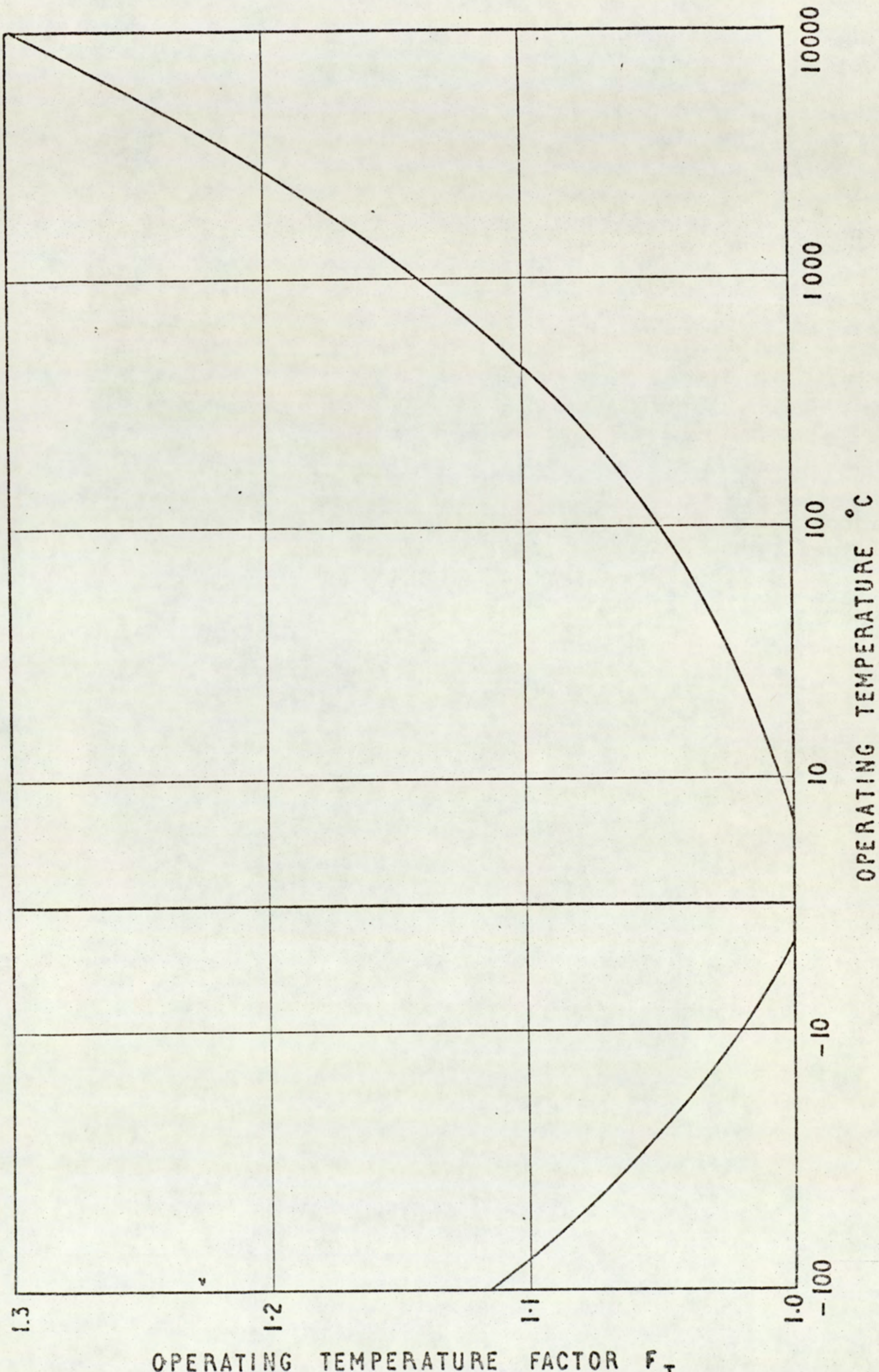


FIGURE 2.2.2.7.3

PRESSURE FACTOR [41]

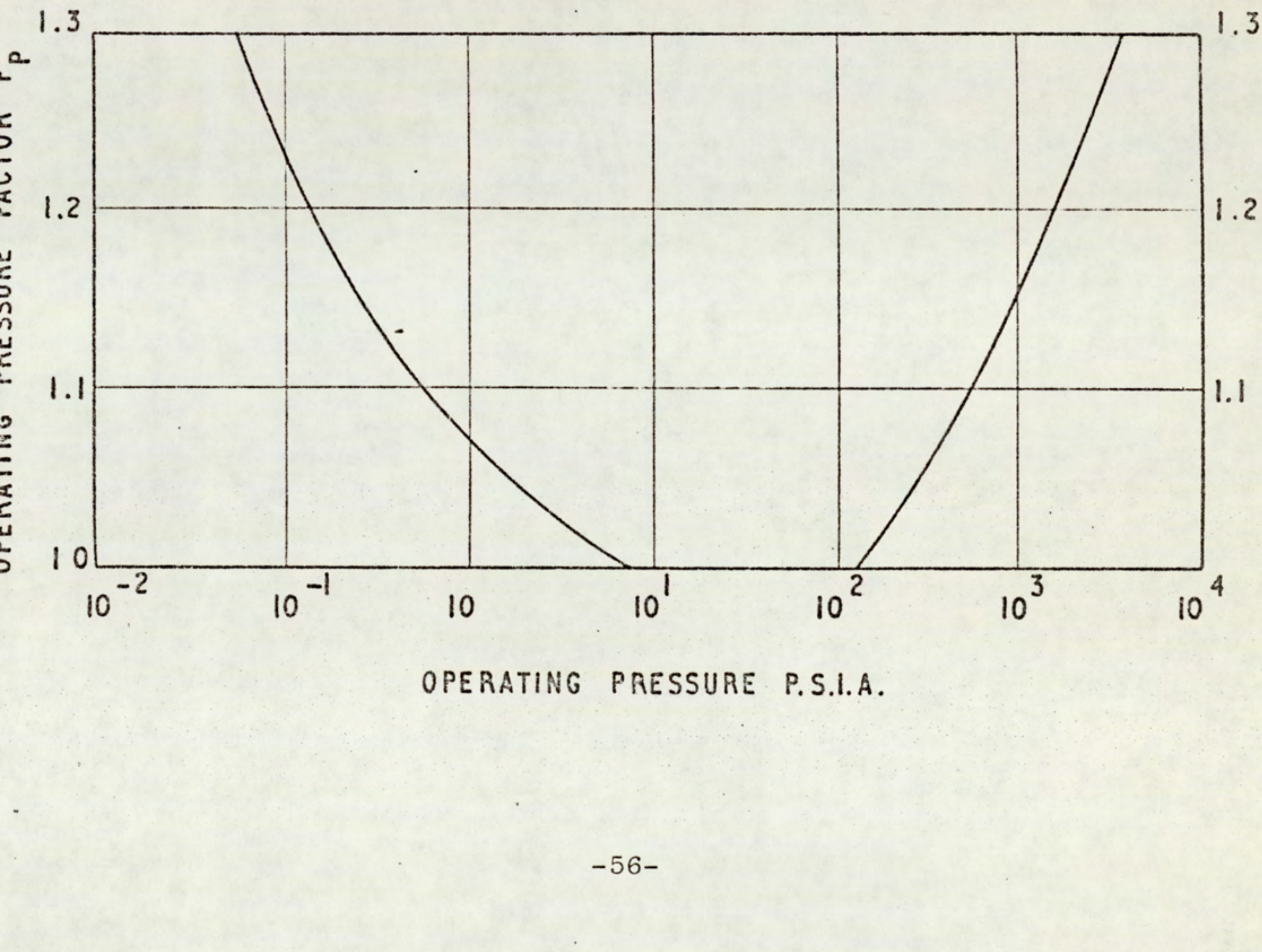


TABLE 2.2.2.7.1

STALLWORTHY'S MATERIAL OF CONSTRUCTION FACTORS (41)

Factor for Material of Construction F_m	Material
1.0	Carbon Steel (mild)
1.05	Bronze
1.065	Carbon/Molybdenum
1.075	Aluminium
1.11	Cast Steel
1.28	Stainless Steel (FDP)
1.41	Worthite
1.5	Stainless Steel (FMB)
1.54	Hastelloy C.
1.65	Monel
1.71	Nickel, Inconel
2.0	Titanium

around $\overset{\times}{\div}$ 1.25. While it is accepted that the method gives good investment results, which is not surprising considering the information needed, it should be noted that it does have a number of limitations which make it unsuitable for use in pre-design cost estimating, namely

a) Stallworthy never actually gave a definition of what constituted a process step, a serious disadvantage when trying to apply the method.

b) Detailed mass balance information is required for the process being estimated. Apart from this being a complex and time consuming procedure and so defeating the objective of 'quick costing', it is also unlikely that the information required to produce the mass balance will be available at the pre-design stage of process development. Hence, the method is restricted to estimating well established and well documented processes if it is to be used as a pre-design cost estimating method.

c) Identification of the main process stream poses a problem. Where more than one feed stream is involved, the main feed stream had to be decided upon.

d) Stallworthy attempted to take into account <u>all</u> process streams; side streams, recirculation streams as well as the main process stream. He did not want to base his method on plant capacity since he realized that this was a cause of major errors. Stallworthy's idea of using a ratio factor to compare the flows of streams to the main process stream, however, still falls into the same trap.

This is because Stallworthy defined the flowrate of the main process stream as the 'output' flowrate of the main stream of the plant. Comparing side streams and recycle streams to the output flowrate may bear very little resemblance to the actual throughputs of the equipment units which go to make up the main stream.

e) The operating temperatures and pressures and the materials of construction of the various pieces of equipment are readily taken into account using Stallworthy's graphs and tables, which were compiled from over 100 different cost samples. He does not state in his paper how the factor values were obtained. The stream factors are calculated as number-weighted mean values and so need reasonably complete process information concerning operating conditions to be known. Taking mean values of these parameters instead of their extremes, as in the previous two methods, assimilated the process plant much better but obviously involved more time and work and needs more information.

f) Stallworthy states that no account was taken of learning effects in deriving his method.

g) Stallworthy also recommends that the method should not be applied to plants of very large capacity.

2.2.2.8 Le Page's Method

Le Page of the Mond Division of I.C.I. Limited developed a quick costing technique for petrochemical plants. The method was presented in an internal ICI report (43) and has since been published (44).

His method is derived from an analysis of I.C.I. data and published data (Zevnik and Buchanan's curve). The fixed capital investment is related to the capacity of the process plant, the number of unit operation stages involved, the costliness imposed by special materials of construction and extremes of operating temperature and pressure, and by the extent to which raw materials, intermediates and final products are diluted by solvents, recycles, reflux streams and by-products in each stage. The method does not differentiate between the functions of the unit operation stages, but uses a standard statistically based cost of a 'basic functional unit' of a given throughput, multiplied by a 'costliness factor' to allow for the special complexities encountered in different chemical plants. The estimating procedure using this technique is as follows:-

1) List the chemical engineering unit operations or functional units likely to be involved in the proposed plant. Le Page (43) defines a functional unit as ' the main and ancillary plant items required to perform a unit operation or process together with its share of civil, structural, electrical equipment, instruments, piping, insulation etc.'. For classification purposes Le Page also provides a list in his report (Table 2.2.2.8.1) of the equipment that should be included.

2) Four aspects of each functional unit are then considered:-

TABLE 2.2.2.8.1

LE PAGE'S LIST OF FUNCTIONAL UNITS (43)

(i) Storage of raw materials, intermediate and final products.

(ii) Reaction Systems

(iii) Heating Systems (if more than simple steam jacketting or coils)

(iv) Refrigeration Systems

(v) Absorption, scrubbing and stripping towers.

(vi) Stills

(vii) Crushers and grinders.

(viii) Filters, screens and centrifuges

(ix) Recovery systems

(x) Effluent disposal systems

(xi) Catalyst preparation systems

(xii) Crystallisers

(xiii) Mixers

(xiv) Heat exchangers (where not part of a system)

(xv) Compressors

(xvi) Specialised equipment not included above

a) temperature extreme; the temperature reached in that particular functional unit which is furthest from ambient temperature (either above or below it)

b) pressure extreme, above or below atmospheric

c) likely material of construction to be used

d) dilution; anything that increases the size of a unit for a given throughput, such as recycle of unreacted raw material or intermediate, a high reflux ratio in a still, or dilution of reacting compounds by a solvent or an inert gas. Dilution is defined as the ratio between the main process stream and a side stream or dilution stream in a functional unit.

Depending on the value of each of the above parameters a score, S, allocated as shown in Table (2.2.2.8.2). Le Page advocated the allocation of an optimistic and pessimistic score for each unit parameter. The parameter scores (optimistic and pessimistic) for each unit are then totalled.

3) From this the number of 'basic' functional units (N) that each module represents is estimated by the following equation:-

$$N = 1.3^S$$

where N = number of basic functional units equivalent to that process module.

S = total score for the module (optimistic or pessimistic).

TABLE 2.2.2.8.2

LE PAGE'S PARAMETER SCORING (43)

Parameter	Score 0	1	2	3	4
Temperature extreme (^{o}C)	0	-25 or +500	-75 or +1,100	-125 or +1,700	-200 or +2,300
Pressure extreme (atmos absolute)	1	0.1 or 10	0.01 or 100	0.001 or 1000	
Dilution	0	0.5	1	2	4
Material Construction	Mild Steel	Nickel, Copper, Aluminium, Lead	Monel, Austenitic Stainless Steel	Inconel,	Hastelloy, Precious Metals

4) The total fixed capital required per basic functional unit (BFU) is then read off from the graph supplied (Figure 2.2.2.8.1), which shows various costs for different capacities of the proposed plant.

5) Knowing the capacity of the plant the total fixed capital investment is then obtained as the product of the fixed capital per basic functional unit x the number of basic functional units x an appropriate inflation factor (IF). i.e. FCI = fixed capital per BFU x Nx IF
Figure (2.2.2.8.1) from which the fixed capital per basic functional unit is obtained was based on 1968 prices (Marshall and Stevens equipment cost index of 273). By carrying the optimistic and pessimistic parameter scores through, an optimistic and pessimistic value for the total fixed capital investment can be calculated. The advantages of the technique are:-

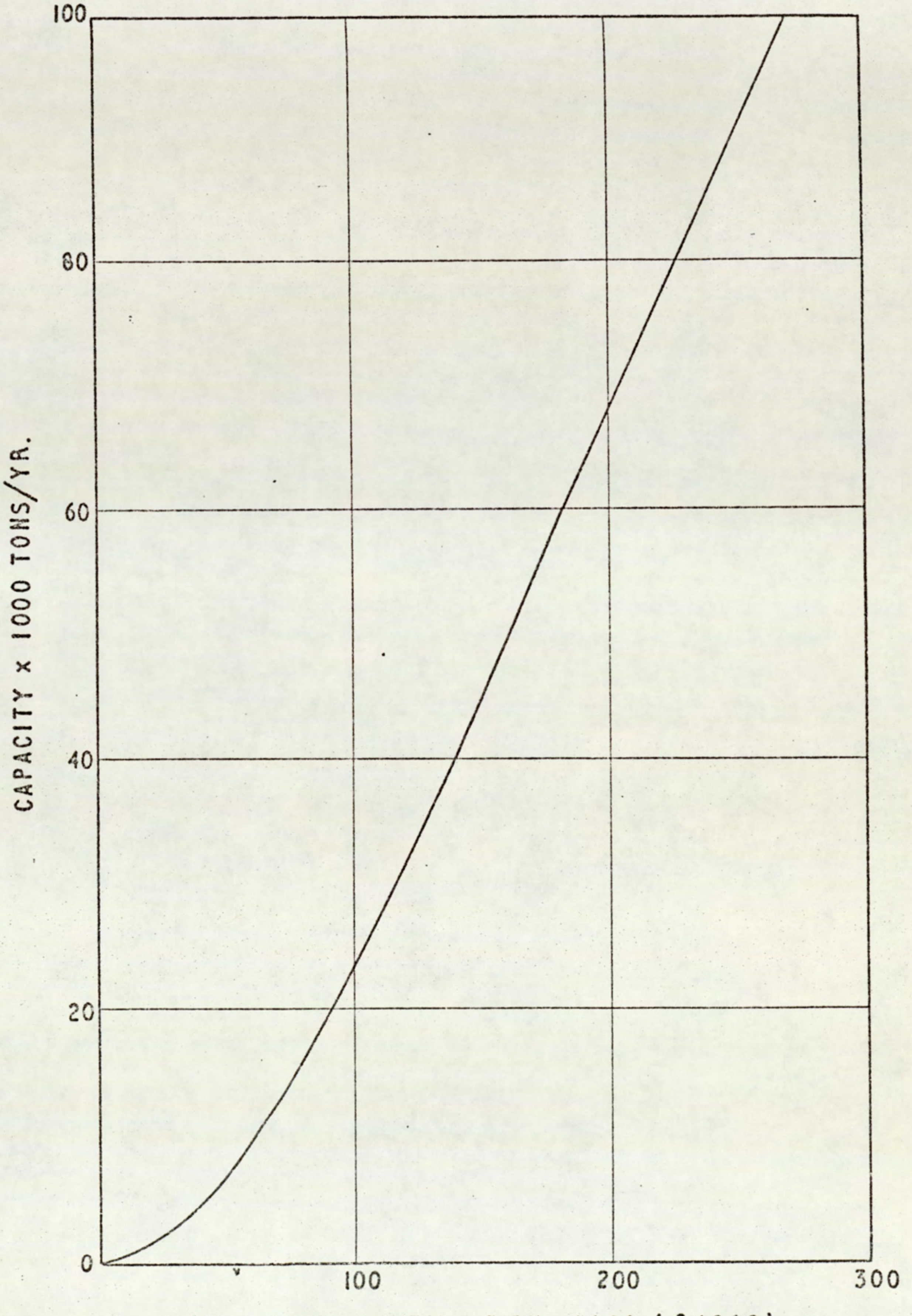
a) The term process module or functional unit is fairly well defined and a list is provided in case clarification is needed.

b) The method does take into account side streams, recycles etc. by the incorporation of the dilution factor.
However, a number of disadvantages exist when using the method:-

i) The method is difficult to apply and, like Stallworthy and Wilson, requires substantial process information. Temperature and pressure extremes, and materials of construction for each of the functional units had to be

FIGURE 2.2.2.8.1

TOTAL FIXED CAPITAL PER BASIC FUNCTIONAL UNIT 43.



known. Determination of the dilution factors for the units necessitated that a mass balance be prepared for the process.

ii) The allocation of optimistic and pessimistic scores for each unit parameter can lead to errors unless accurate data is available. In most cases it is probable that parameter values will fall between the ones quoted in the scoring table and so will have to be allocated intuitively. Page (48) found that the differences between optimistic and pessimistic estimates are so great that the usefulness of this approach is questionable when comparing different plants competing for investment.

iii) The method is derived from petrochemical cost data and its use should possibly be restricted to organic fluid phase processes since no phase factor is incorporated.

iv) A mixture of English and American cost data was used to develop the method. However, Le Page makes no mention of 'standardizing' his data to a common base.

v) Le Page failed to explain how he arrived at all his parameter scoring values and also how he determined the correlation between basic functional unit and total parameter score. He indicated however that a basic functional unit operated at atmospheric pressure and $0^{o}C$, was constructed from mild steel and had no dilution (i.e. a total parameter score of zero).

vi) Page (48) claims the method tends to overestimate and its accuracy is suspect.

2.2.2.9. Taylor's Method

Taylor's method (45) is a continuation of Le Page's work at I.C.I. Ltd, and is one of the most recent publications in this field of study. He refers to his technique as the 'process step scoring method'. The method was derived from an analysis of 45 I.C.I. U.K. projects. Accuracies of ± 36% to -26% within 95% confidence limits (standard error 15%) were obtained by the method, which is good enough for most preliminary cost evaluations. This claim has been independently supported (20).

Taylor's method is based on a system in which a complexity score accounting for such factors as throughput, corrosion problems, reaction time and similar cost relevant parameters, is estimated for each 'significant process step' and these are combined to give an overall 'costliness index'. The capital cost is then derived from a relationship between this index and process capacity.

$$\text{Capital Cost} = \text{constant} \times \text{costliness index} \times (\text{capacity})^n$$

where n = capacity exponent.

The estimating procedure is as follows:-

1) Defining the proposed flowscheme and drawing the conceptual flowsheet.

2) Identification of the number of 'significant process steps', whereby a process step refers to the operation on the material flow and not the equipment which is necessary

to perform it. No clearer definition is given but a list is provided for clarification purposes. (see Table 2.2.2.9.1).

3) Calculation of the costliness index, I, where

$$I = \sum_{1}^{N} (1.3)^{S}$$

and N = number of significant process steps

S = complexity score.

Each significant process step is 'scored' or weighted on the following variables-

a) Relative throughput; calculated as the total weight flow into the process step per unit weight of final product. Internal recycles are ignored.

b) Materials of construction; which Taylor recognized as a powerful variable influencing capital cost. He noted that several materials of construction may be present in each process step but he uses the dominant material of construction in his method.

c) Temperature; score based on the maximum temperature within the step.

d) Pressure; score based on the maximum pressure within the step. Up to 50 atmospheres the pressure was assumed to have no effect since it was assumed that the increased cost due to extra wall thickness was offset by reduced vessel size.

e) Multistreaming; which might be considered necessary for reasons such as poor reliability, limitations of

TABLE 2.2.2.9.1.

TAYLOR LIST OF PROCESS STEPS (45)

1. Chemical Reactions.
2. Neutralization/Acidification.
3. Storage/handling of a raw material, product, by-product, intermediate or recycle stream. Also effluents when these are not assumed to be discharged directly to outside battery limits.
4. Filter, screen or centrifuge.
5. Distil, evaporate, fractionate or strip.
6. Crystalise or precipitate.
7. Formulate.
8. Compress
9. Vaporise.
10. Dry or spray dry.
11. Mill or grind.
12. Scrub or absorb.
13. Packing into special containers (not sacks or drums).
14. Quench (but not normal cooling of a reaction mixture).
15. Phase separation of a reaction mixture but not when part of a still or an extraction system.
16. Extract or leach.
17. Condense when used to separate a component from a gaseous stream containing inerts (but not for normal condensation in stills, quenchers or reactors).
18. Dissolve, mix, slurry or blend when required as a specific pre or post treatment (but not when an integral part of another process step such as, say, Solvent extraction).

equipment size, uncertain market conditions. It was noted that multi-streaming was in some way a function of capacity. i.e. as capacity increases it may be advantageous to recover, or recycle, material streams.

f) Reaction time/storage time; these have different scales. The reaction time scale refers to liquid phase process steps. The cost of gas phase reactions are usually effected only slightly by the residence time and therefore no scoring scale is included for them.

g) Special conditions, which could include explosion hazards, odour, dust or toxicity problems, fractionation of materials of similar boiling points, fluid bed reactions, film evaporation and tight specifications. These are assessed on an empirical basis.

Having scored each significant process step according to Taylor's weightings the total score for each step is converted into a costliness index using the calculated factors.

4) The costliness indices for each process step are summed to give the costliness index, I, for the whole process.

5) Battery limits capital cost is estimated as

$$£C = 42\ (Q)^{0.39}\ (I).\ CCI$$

where C = battery limits capital cost of a newly built plant (including storage)

Q = design capacity in 1000 tons

I = costliness index

CCI = construction cost index.

Taylor's equation gives cost at January 1977 time base (EPE Plant cost index = 280).

Taylor states that his method has the following limitations and is unsuitable for estimating:-

- Very simple plants ($N < 5$) i.e. the more steps the better.
- Modifications to or extensions of existing plants.
- Fully batch operated plants of abnormally high capacity (3000 ton/year or more).
- Plants involving appreciable solids handling on a large scale (more than 5000 ton/year).
- Plants involving special operations such as electrolysis, fibre spinning, extrusion etc. when these are likely to represent a substantial proportion of the cost.

Preferably the method should be used for complete new plants of capacities ranging from 300 to 250,000 ton/year.

Apart from the limitations Taylor himself imposes with his technique, other disadvantages can be identified:-

Detailed process information is required, which limits its use to the later stages of process development. Taylor's conception of a 'significant process step' is a novel and interesting one. Unlike others in this field who have attempted to define what is and is not a significant process step in relation to process capital cost, Taylor simply relates cost to the number of process functions or operations without any regard to the actual hardware in the process. His scoring approach then attempts to 'weight' or standardize the process steps identified to assess the 'costliness' or complexity and so achieve a realistic cost estimate. Whilst the weighting/scoring concept is a good one it does mean that the scoring procedure does have to be very sophisticated to cope with the huge cost differences between simple process items such as flash drums and the more complex ones such as distillation columns. It is likely that Taylor had sound data from which to develop his method. However, without more information on the development of the method it is asking much of the user to assume that this method is capable of producing realistic cost estimates by the means stated. His exponent on capacity of 0.39 is surprising. However, the author is adamant about its correctness (46). Mathematically, this figure appears to be suspect since a capacity/throughput relationship exists within the costliness index, and thus the Q term in his estimating equation is not an independent variable. This could account for the unexpected exponent value, in Q.

2.2.2.10 <u>Wilson's Method</u>

Wilson's method (47) is an example of module estimating (or unit estimating). It was developed to derive the delivered equipment cost of a process and then apply an overall Lang type factor to give total capital cost.

There appear to be two levels of module estimating; one where the capital cost is a function of the average cost of each unit (48,49), which is referred to as simple module estimating; and one where each unit is treated individually (34) which is referred to as complex module estimating. Wilson's technique is an example of the former.

The following equation is proposed for the capital investment required for a process plant:-

$$C = f \times N \times (AUC) \times F_m \times F_p \times F_T$$

where C = battery limits investment (£) of the process (in 1971)

f = investment factor (analogous to Lang factor)

N = number of main plant items.

AUC = average unit cost of main plant items ($= 21\ V^{0.675}$, where V = capacity, tons per year)

F_m = factors for specific materials of construction

F_p = factor for design pressure

F_T = factor for design temperature.

A cost index would also have to be incorporated for current costs. The factors F_m, F_T and F_p appear to be those developed by Stallworthy, previously shown in Table (2.2.2.7.1), Figure (2.2.2.7.2) and Figure (2.2.2.7.3) respectively.

Wilson gives a graph of average throughput of the main plant equipment items (V tons/annum) versus their average unit cost (Figure 2.2.2.10.1). The investment factor is obtained from a graph of factor values versus average unit cost as shown in Figure (2.2.2.10.2). Wilson obtained

FIGURE 2.2.2.10.1

AVERAGE UNIT COST[47]

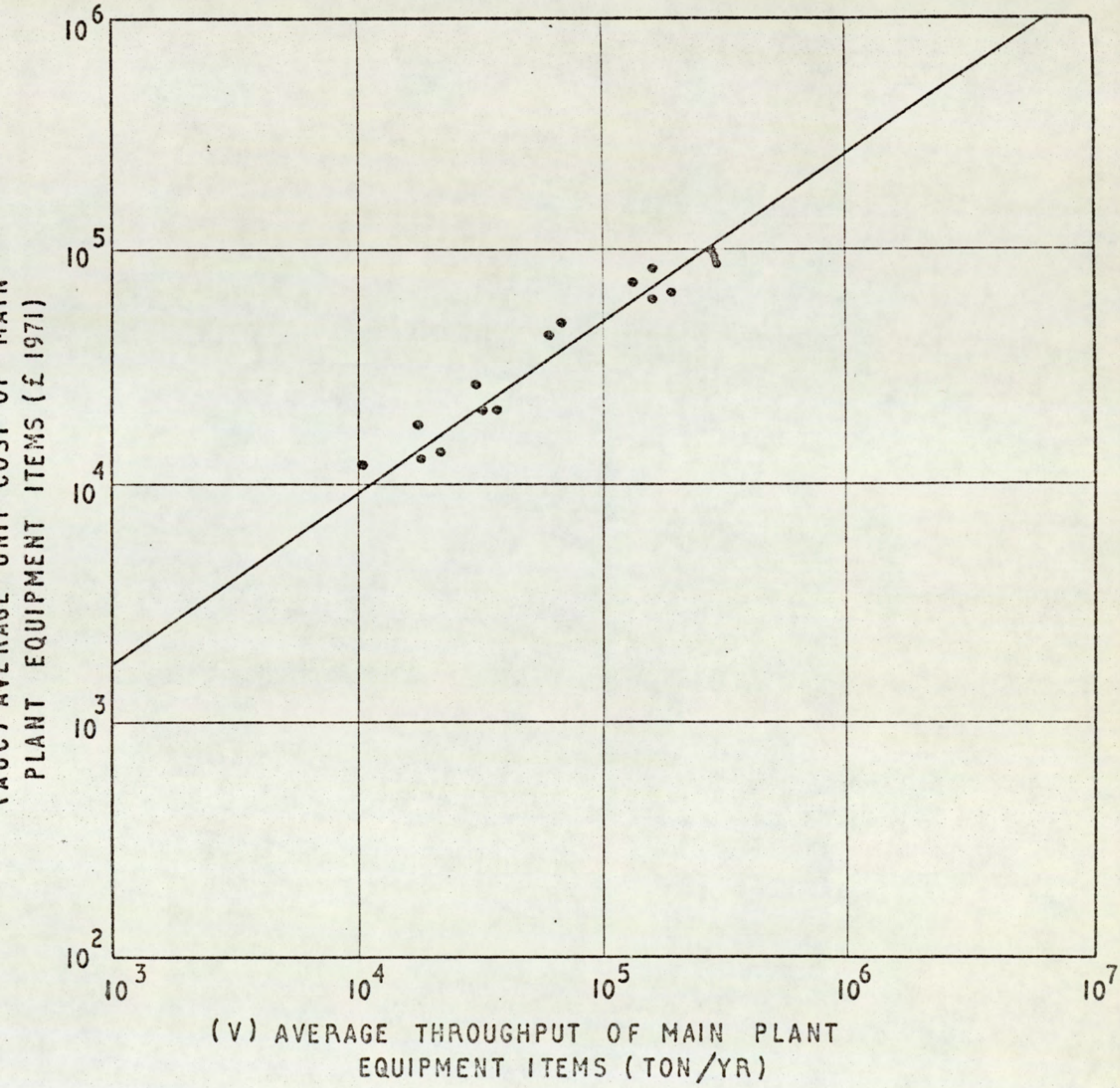
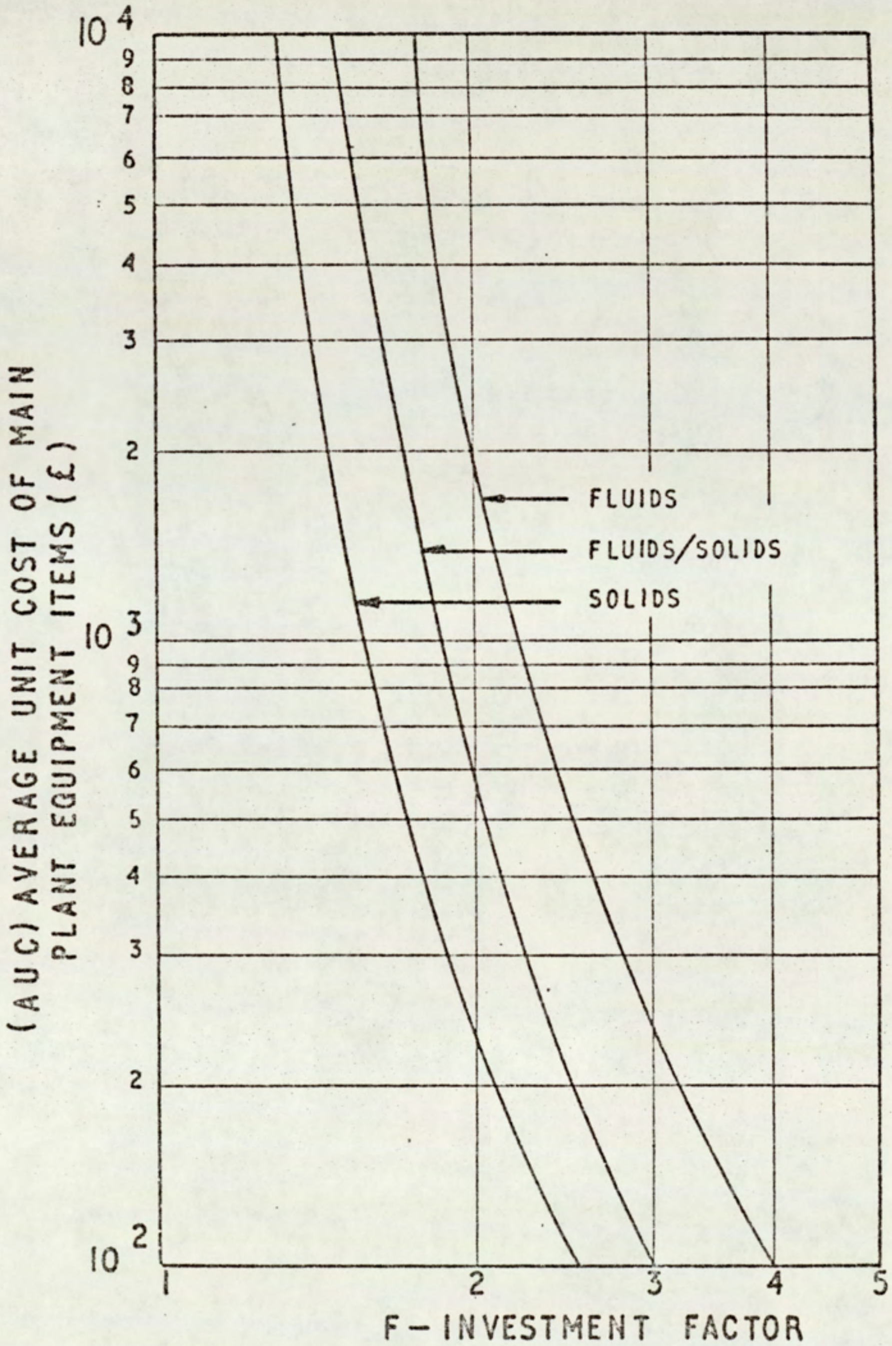


FIGURE 2.2.2.10.2.
INVESTMENT FACTORS[47]:



this graph by following the Lang factor categorization (i.e. fluid, solid/fluid, solid), so making the investment factor a function of the dominant phase, to which he applies Miller's idea that the factor is a function of plant size, or rather average unit cost (34). The number of main plant items is taken to include all major pieces of equipment involved in the process plant, such as reactors, columns, heat exchangers etc. Pumps are excluded. Wilson considered 16 plants in arriving at his correlation. He aimed at an error range of ± 30% and of the plants considered 13 fell within the range. The average throughputs of the various processes ranged from 10^4 to 10^6 tons/annum and Wilson stated that extrapolation above or below this range may not be justified. The method does have recognized advantages:-

- Calculation of the number of main plant items is quite straightforward (remembering not to include pumps).
- Plant throughput is used to measure plant size (rather than output).
- Wilson defined plant throughput as the sum of all the process streams entering an equipment item (service streams excluded).
- The investment factor is related to both the type of plant (process phases) and average throughput.

However, the method does also have disadvantages:-

- Substantial process data is required

a) to calculate the average unit cost of the plant items the average throughput per item, V, has to be determined. Hence detailed mass balance information is required. Plant capacity alone does however appear to be an acceptable alternative (20).

b) assessment of the temperature, pressure and material of construction factors requires the value of these parameters to be known or estimated for each plant item.

Only 16 data sets were used to derive the method, which suggests the method should be treated with caution. The accuracy of the method is suspect according to Page (48) who states that low and inconsistent results are obtained. Bridgwater (20) also notes that the method gives low results but that it does stay within its stated ± 30% accuracy range.

2.2.2.11 <u>Page's Method</u>

Page's method (48) is similar in principle to Wilson's, being based on an estimate of delivered equipment cost, but is more detailed. It requires more information and is much more complex to use than other rapid cost estimating techniques (20). It is however claimed to give a more accurate result. As with Wilson's method, the approach is different from the process step or functional unit concept, in that Lang type factors are required, and concentrates on a main plant item approach. He developed his proposal from eight sets of cost data published in the chemical engineering literature and claimed an accuracy of $^+_-$ 25% to - 20% from the results he obtained. The following procedure for estimating capital cost is proposed; the calculation being performed in two parts:-

1) Calculation of the Delivered Equipment Cost

$$DEC = N \times BIC \times SF$$

where DEC = delivered equipment cost

N = number of main plant items

BIC = basic item cost

SF = state factor

All pieces of equipment shown on the process flowscheme were taken to be a main plant item. It did not matter what the equipment was as long as it was shown on the flowsheet. Page included pumps, for example. The state factor, SF, is used to take account of the temperature, pressures and materials of construction encountered in the process and was defined as

$$SF = (F_t)_{max} \times (F_p)_{max} \times (F_m)_{mean}$$

where SF = state factor

F_t = temperature factor

F_p = pressure factor

F_m = material of construction factor.

Similar temperature, pressure and materials of construction factors are employed to those developed by Stallworthy and used by Wilson (see Figure (2.2.2.7.2), (2.2.2.7.3) and Table (2.2.2.7.1)). The F_t and F_p factors were based on maximum process temperature and pressure as it was thought that any advantage in assuming values for all main plant items in order to derive a number weighted mean value, did not justify the effort involved. For F_m however, a number weighted mean value was thought to be practical and better.

Basic Item Cost (BIC) is the effective cost of a 'standard' main plant item for the plant being estimated i.e. before adjusting for operating conditions. It is a function of the throughput variable, TP, which is defined as

$$TP = Q \times FF \times PF$$

where TP = plant throughput

Q = total plant feed (excluding utilities and services) measured in lb moles/annum, (i.e. volumetric basis)

FF = flow factor, to account for the effect of recycle, reflux, and side streams of a process (in the absence of a mass balance) and defined as

$$FF = \frac{\sum_{1}^{N} \text{Number of input and output process streams of each main plant item}}{\text{Number of main plant items (N)}}$$

The more complex the process flowscheme the greater the number of streams and the larger the flow factor.

PF = phase factor, to account for the phase of operation of the individual equipment items, and thus equipment size and cost, and defined as

$$PF = 0.0075 + \frac{\text{Number of Volume items } (V_I)}{\text{Total number of main plant items (Volume } (V_I) \text{ + weight } (W_I) \text{ items)}}$$

where PF = phase factor

V_I = volume item; defined as those operated in gas, gas/liquid or gas/solid phase

W_I = weight item; defined as those operated in the solid, liquid or solid/liquid phase.

The constant 0.0075 represents the ratio of the densities of a 'typical' hydrocarbon gas to its corresponding liquid feedstock and ensures that in the limiting case of a plant with no volume items the value of the throughput would be reasonable.

Having defined and calculated the throughput term for his correlation the basic item cost, BIC, is derived from a cost-throughput graph (Figure 2.2.2.11.1). However, instead of using only one line for all plants, Page derived a cost/throughput relationship for each separate plant using his self-developed 'cost weighted capacity exponent method'. He argued that since capacity exponents for process plants varied from 0.4 to 0.9 depending on the process under consideration, any attempt to use a single line in cost/capacity graphs was the source of large errors. Hence, he decided to calculate the overall plant exponent as a function of the individual exponents of the equipment items. It was proposed that the overall exponent was equal to the cost-weighted mean value of the individual equipment exponents.

$$\sum_{i=1}^{N} (C_i . R^{Exp}) = \sum_{i=1}^{N} (C_i . R^{ei}$$

where C_i = base cost of main plant item, i

R = ratio of new to base capacity

Exp = plant exponent

e_i = exponent of main plant item, i.

which rearranges to give

$$EXP = \frac{\log \sum_{i=1}^{N} (C_i . R^{ei}) - \log \sum_{i=1}^{N} (C_i)}{\log R}$$

which simplifies to

$$\sum_{i=1}^{N} (C_i . e_i) / \sum_{i=1}^{N} (C_i)$$

The equipment exponents given by Guthrie (53) were used to cost a wide range of equipment at comparable levels of performance, namely an input of 50000 lb/h and time base of 1968. Using this data (see Table 2.2.2.11.1) EXP was calculated, with the aid of a computer, for the overall process. Having calculated the EXP value and also, having plotted the cost-throughput relationships of his plants (see Figure 2.2.2.11.1) Page noted that all the cost/throughput lines came close to passing through a point corresponding to a throughput value of 2.5×10^6 lb mole/annum and a basic item cost of \$7000. Hence, a cost-throughput relationship was established whereby this value may be scaled up using the calculated EXP value. The procedure was thus developed for finding the basic item

TABLE 2.2.2.11.1

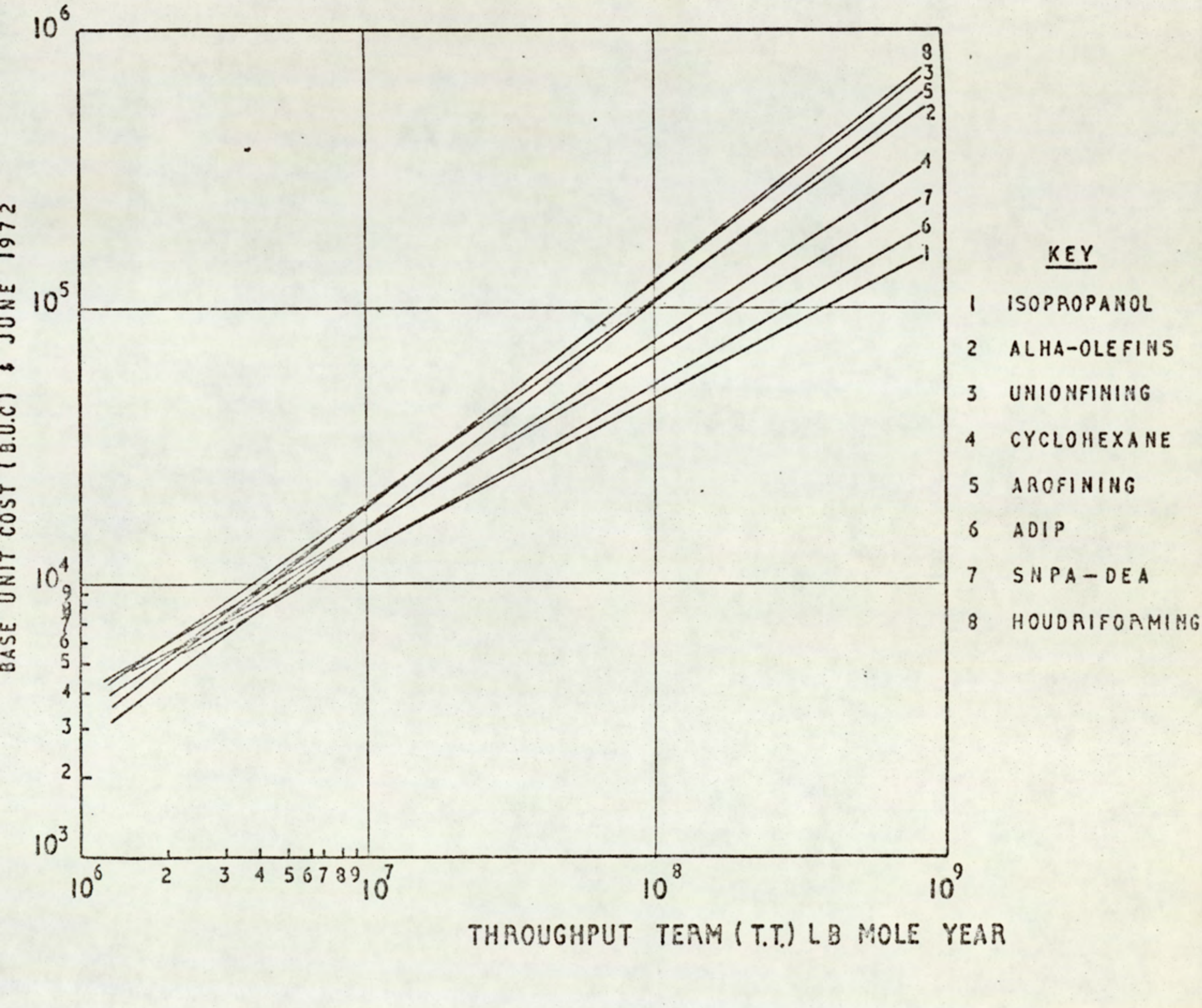
EQUIPMENT CAPACITY EXPONENTS AND COST WEIGHTINGS (48)

Equipment Item	Capacity Exponent	Cost Weighting
Process Furnace	0.85	135
Direct Fired Heater	0.85	103.5
Boilers (industrial) - 15 psig	0.5	92
150 psig	0.5	101.2
300 psig	0.5	115
600 psig	0.5	138
Packaged Boiler Unit	0.7	60
Shell & Tube Exchanger	0.65	6.5
Kettle, Reboiler Exchanger	0.65	8.8
U Tube Exchanger	0.65	5.5
Cooler	0.66	6.8
Cooling Tower Facilities	0.6	9.9
Tower with Trays	0.73	33.5
Tower with Packing	0.65	35.2
Pressure Vessel - Vertical	0.65	7.6
Horizontal	0.6	5
Storage Tank	0.3	6
Horizontal Pressure Storage Vessel	0.65	4.8
Spherical Pressure Storage Vessel	0.7	8
Centrifugal Pump - Centrifugal/Motor	0.52	1.5
Centrifugal/Turbine	0.52	3
Reciprocating Pump - Reciprocating/motor	0.7	6
Reciprocating/steam	0.7	1.1
Process Gas Compressor - 1000 psig	0.82	85
Air Compressor - 125 psig	0.28	36.5
Crushers - Cone	0.85	12
Gyratory	1.2	3
Jaw	1.2	4.7
Pulverisers	0.35	23.4

TABLE 2.2.2.11.1 (continued)

Mills -	Ball	0.65
	Roller	0.65
	Hammer	0.85
Evaporators -	Forced Circulation	0.7
	Vertical Tube	0.53
	Horizontal Tube	0.53
	Jacketted Vessel	0.6
Hoppers -	Conical	0.68
	Silos	0.9
Blowers and Fans		0.68
Crystallisers -	Growth	0.65
	Forced Circulation	0.55
	Batch	0.7
Filters -	Plate and Press	0.58
	Pressure Leaf (wet)	0.58
	Pressure Leaf (dry)	0.53
	Rotary Drum	0.63
	Rotary Disk	0.78
Dryers -	Drum	0.45
	Pan	0.38
	Rotary Vacuum	0.46

FIGURE 2.2.2.II.I.
COST/THROUGHPUT GRAPH[48.]



cost for a new plant and for adjusting it to find the plant delivered equipment cost.

2) Calculation of the Battery Limits capital cost from the Delivered Equipment Cost.

Page found over 20 references giving the breakdown of total project capital costs into its various component costs. From each of the references the delivered equipment cost was determined as a percentage of the fixed capital investment and an average value of 23.4% was obtained for grass roots investments and 31.2% for battery limits investments. When calculating the delivered equipment cost percentages only fluid and solid/fluid plants were used. These percentage factors calculated by Page were in good agreement with those developed by Haselberth and Berk (64) whose study covered 70 U.S. Gulf Coast area sites for all types of plants and thus it was decided to adopt their factors. The estimating procedure is thus described, the final step in the method being to update costs using the Marshall and Stevens cost index with June 1972 reference values of 331.

- The capacity range over which the technique is applicable is restricted by the plants considered in the development i.e. $9 \times 10^6 < \text{FEED} > 220 \times 10^6$ lb moles/annum which corresponds to $15 \times 10^6 < \text{THROUGHPUT} > 340 \times 10^6$ lb moles/annum.
- The graphs used to calculate F_t and F_p are Stallworthy's and their accuracy at values approaching the specified

range limits is questionable. He postulates that it is likely that the graphs were obtained by extrapolation rather than from actual data.

- The processes used in deriving the method were all petrochemical processes and thus the phase factor theory proposed is untested due to no weight items being present. Therefore strictly speaking the method is only applicable to petrochemical processes.

Other disadvantages or limitations also exist;

- Considering that one of the major objectives of his study was to produce a pre-design estimating method the information required for his method is formidable. Detailed process flowsheets are required to estimate the number of main plant items and flow configurations and also, the process feed (in lb mol/annum) is required. Such information is rarely available without detailed mass balance calculations. Also, for complex feed mixtures of unknown molecular weight calculation of feed in lb mols will be difficult.
- As can be seen from the description the method is complex and time consuming and this together with its information requirements do not make the method plausible for use as a quick screening estimating method by process engineers.
- The correlation is developed from a data base which is insufficient to support the proposed theory and draw specific conclusions from. Only eight processes were

used to test the theory and yet far more data is available in the published literature.

- Errors will arise when calculating battery limits costs from delivered equipment costs when applying general or 'typical' percentage factors, especially when it is considered that the factors proposed were based on a 1960 plant survey and that subsequent technological advances in construction may make them invalid. However, users with access to their own cost files can substitute their own data for Page's to overcome this problem.

To summarize Page's proposal then it can be said that he postulated some interesting theories, particularly the use of individual equipment exponents to derive an overall plant exponent value, and conducted his research in a rigorous manner, using recent cost data and clearly defining all the terms used in his correlation. However, it is considered that the information required to use his method is in many ways incompatible with the information usually available for pre-design cost estimating, and as such it is difficult to see just how and where the method will be of practical value, except possibly as a final checking estimate immediately prior to detailed design. Furthermore, the application and testing of his theory was disappointing and more information is required as to how the method copes with a wider range of processes before any degree of confidence is attached to this method.

2.2.3. Preliminary Estimates

Preliminary class capital cost estimates have a probable error of up to ± 20%. It is very difficult to define a clear dividing line between methods that are used for study estimates and those used only for preliminary estimates. Study estimating methods can often be applied to preliminary estimates depending on the confidence the estimator can place on the available data.

Preliminary estimates, sometimes called budget authorisation estimates, are prepared from carefully evaluated flowsheets, detailed equipment lists and good site and structure information. Such estimates, with an error range of the order of ± 20% or better, are often the basis for the original budgeting of capital funds.

Preliminary estimates, together with definitive estimates (see section 2.2.4) are generally derived by detailed factorial techniques. All the methods reviewed from this point on are sophisticated techniques developed to give high accuracies. However they all require considerable detailed information in order to use them and as such they are not suitable for pre-design cost estimating. They are reviewed here in order that the techniques of detailed estimating and the accuracy-information relationships which exist, may be understood to assist in defining research aims.

2.2.3.1 PERCENTAGE OF DELIVERED EQUIPMENT COST

In this method the delivered equipment cost must be determined first, from which the fixed capital investment can be estimated. The other items included in the total direct plant cost (see Table 2.2.5.1) (9) are then estimated as percentages of the delivered equipment cost. The remaining items which go to make up the fixed capital investment are based on average percentages of the total direct and indirect plant costs.

i.e. $C = (f_1E + f_2E + f_3E + \text{---}) (1 + f_I)$.

where C = fixed capital investment

E = delivered equipment cost.

f_1, f_2, f_3 = Multiplying factors for installation, instrumentation, piping etc.

f_I = indirect cost factor.

The percentages used in making a preliminary class estimate are determined on the basis of the type of process involved, design complexity, required materials of construction, location of plant, past experience and other items dependent on the particular unit under consideration. Average values of the various percentages have been published for typical chemical plants by numerous authors (27), (28), (32), (48), (49), (50), (51).

TABLE 2.2.5.1

CHECK-LIST OF FIXED CAPITAL INVESTMENT ITEMS FOR A CHEMICAL PLANT (9)

Direct Costs

1. Purchased equipment.

   All equipment listed on a complete flow sheet.
   Spare parts and non-installed equipment spares.
   Surplus equipment, supplies and equipment allowance.
   Inflation cost allowance.
   Freight charges.
   Taxes, insurances, duties.
   Allowance for modifications during start-up.

2. Purchased equipment installation.

   Installation of all equipment listed on complete flow sheet.
   Structural supports, insulation, paint.

3. Instrumentation and controls.

   Purchase, installation, calibration.

4. Piping.

   Process piping - carbon steel, alloy, cast iron, lead, lined aluminium, copper, asbestos-cement, ceramic, plastic, rubber, reinforced concrete.
   Pipe hangers, fittings, valves.
   Insulation - piping, equipment.

5. Electrical equipment and materials.

   Electrical equipment - switches, motors, conduit, wire, fittings, feeders, grounding, instrument and control wiring, lighting, panels.
   Electrical materials and labour.

6. Buildings (including services)

   Process buildings - substructures, superstructures, platforms, supports, stairways, ladders, accessways, cranes, monorails, hoists, elevators.
   Auxiliary buildings - adminstration, medical or dispensary, cafeteria, garage, product warehouse, parts warehouse, guard and safety, fire station, change house, personnel building, shipping offices and platform, research laboratory, control laboratory.
   Maintenance shops - electrical, piping, sheet metal, machine, welding, carpentry, instrument.
   Building services - plumbing, heating, ventilation, dust collection, air conditioning, building lighting, elevators, escalators, telephones, intercommunication systems, painting, sprinkler systems, fire alarms.

TABLE 2.2.5.1.(continued)

7. Yard improvements.

   Site development - site clearing, grading, roads, walkways, railroads, fences, parking areas, wharves and piers, recreational facilities, landscaping.

8. Services Facilities.

   Utilities - steam, water, power, refrigeration, compressed air, fuel, waste disposal.
   Facilities - boiler plant incinerator, wells, river intake, water treatment, cooling towers, water storage, electric substation, refrigeration plant, air plant, fuel storage, waste disposal plant, fire protection.
   Non-process equipment - office furniture and equipment, cafeteria equipment, safety and medical equipment, shop equipment, automative equipment, yard material-handling equipment, laboratory equipment, locker-room equipment, garage equipment, shelves, bins, hand trucks, house-keeping equipment, fire extinguishers, hoses, fire engines, loading stations.
   Distribution and packaging - raw material and product storage and handling equipment, product packaging equipment, blending facilities, loading stations.

Indirect Costs.

1. Engineering and Supervision.
   Engineering costs - administrative, process, design and general engineering, drafting, cost engineering, procuring, expediting, reproduction, communications, scale models, consultant fees, travel.
   Engineering supervision and inspection

2. Construction Expenses.
   Construction, operation and maintenance of temporary facilities, offices, roads, parking lots, railroads, electrical, piping, communications, fencing.
   Construction tools and equipment.
   Construction supervision, accounting, timekeeping, purchasing, expediting.
   Warehouse personnel and expense, guards.
   Safety, medical, finge benefits.
   Permits, field tests, special licenses.
   Taxes, insurance, interest.

3. Contractor's fees.

4. Contingency.

This preliminary estimating method yields most accurate results when applied to chemical process plants similar in configuration to recently constructed plants. Peters and Timmerhaus (10) state that 'for comparable plants of different capacity, this method has sometimes been reported to yield definitive estimate accuracies'.

2.2.3.2. Waddell's Method

In 1961 Waddell (33) presented a paper in which he described a cost estimating system developed by Du Pont Limited. The system was based on factoring technique and made use of 'expansion factors' which were peculiar to Du Pont Limited. For this reason the expansion factors and equations of the system were not given in the paper.

The system was similar to other factoring systems of the time except for several basic differences:-

1) Waddell emphasised the need to provide factors which took account of the strong effect of size.

2) It was statistically rather than empirically based.

3) It was applicable for any arrangement of four or more differing types of equipment.

4) It was stated to be accurate to within $\pm$ 10% of the actual value.

It is impossible to verify the accuracy claim since no precise information was given on the practical application of the system. However, since the method was based on reliable, company cost data, it has been grouped in the preliminary estimating class.

2.2.3.3. Miller's Method

Miller (34) published a method in 1965 based on factoring technique. The method he proposed may be described as a modified Lang factor method but he overcame many of the problems associated with the Lang factor method and as such achieved much better accuracies. Hence, the reason for his method being described and discussed under the preliminary estimate heading.

In addition to taking into account the nature of the process plant phase of operation (i.e. solid, fluid or solid/fluid) he also considered the effect of the size of the equipment, materials of construction, and operating pressures involved in the plant.

i.e. Total Capital Cost = factor(s) x delivered equipment cost.

If the size of the process equipment gets larger, the overall factor becomes smaller. If the equipment is made from materials such as stainless steel, inconel etc., the factor again becomes smaller. If the operating pressures increase it is also known that the overall factor decreases.

Miller suggested that to a considerable degree, all these items could be taken into account by one number, the 'average unit cost' of the process equipment, which he defined as:-

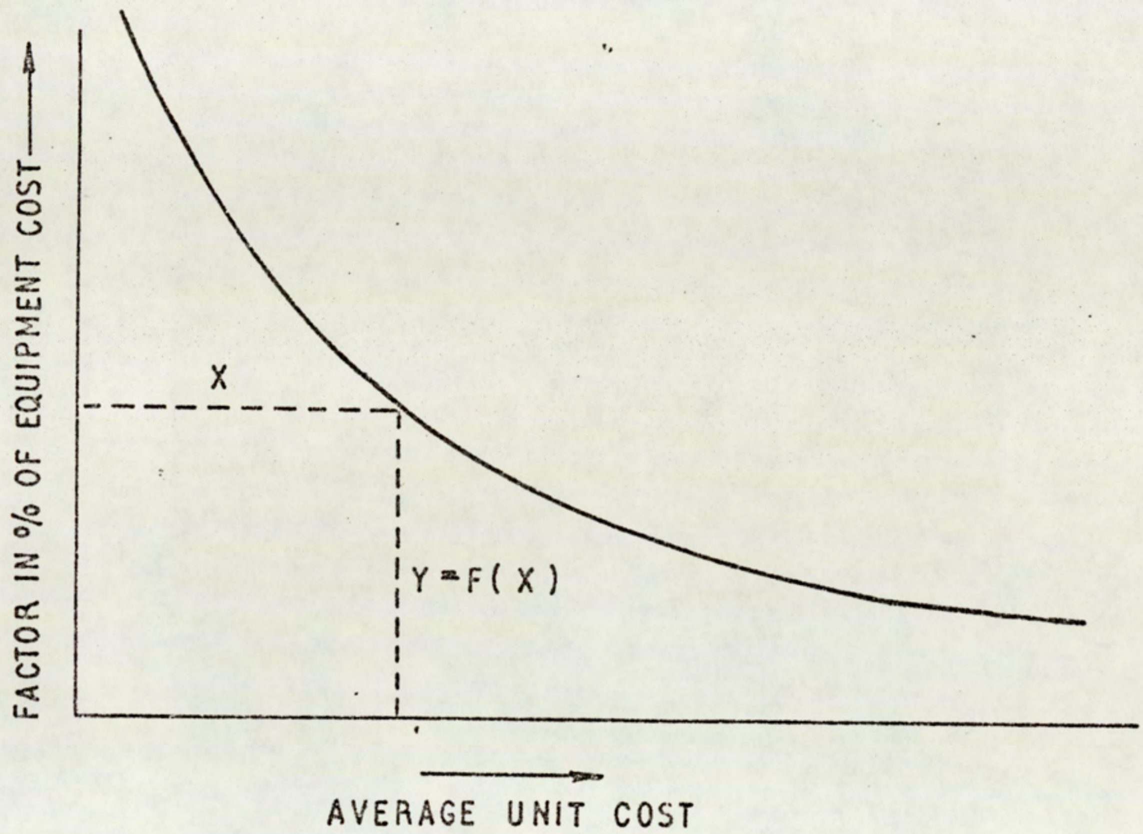
$$\frac{\text{Total Cost of Process Equipment}}{\text{Number of Equipment Items}}$$

This was a unique step made by Miller in as much as he combined two estimating concepts into his method, i.e. factoring and step counting, in order to achieve greater accuracies, and thus he provides us with an example of complex module estimating which was developed by other workers in later years. (47,48).

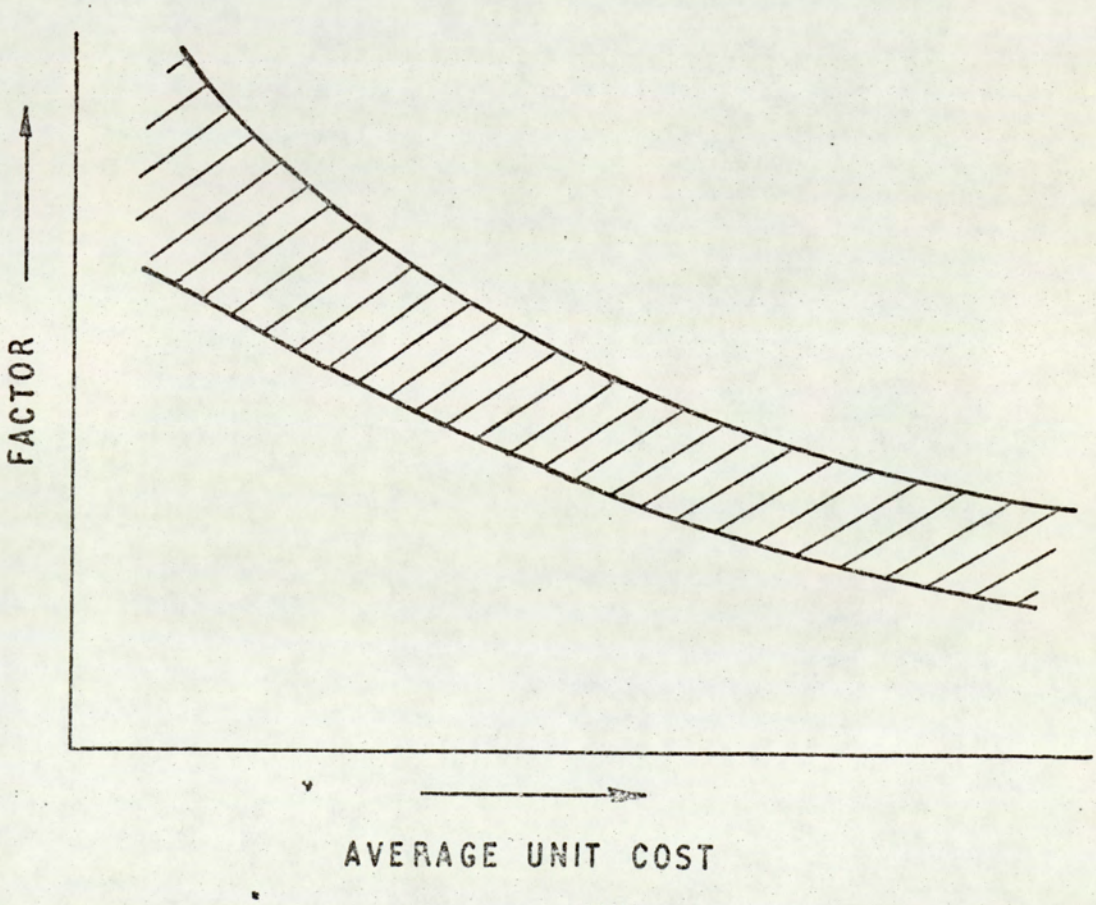
Miller (34) gave two graphs which illustrated how his average unit cost method could be used (Figure 2.2.3.3.1). If the size of a plant is increased, the equipment becomes larger and the average cost per item increases. Thus, the point on Miller's curve is farther to the right and the corresponding factor is lower. Alternatively, if the equipment has been carbon steel and is changed to stainless steel, the average cost of each equipment item is increased and the factor is reduced. Similarly, if the operating pressure is increased from atmospheric to a high pressure operation the average cost is again increased and the resulting factor lower. Hence, it follows that regardless of what issues cause variations in the factors, the average unit cost approach has a narrowing effect on the differences.

FIGURE 2.2.3.3.1

MILLER CURVES[34]



IN PRACTICE, THE CURVE BECOMES A BAND



Although when used as a pre-design estimating method Miller's method suffers from the major drawback of requiring considerable information, most of which may not be available in the early stages of process development, his technique does have many advantages for specific applications and high degrees of confidence are attached to the accuracy and reliability of the estimates obtained by his method. Major reasons for this are that Miller developed his method from analyzing the feedback from a great many chemical plants, thus providing a reliable data base, and derived factors for all cost components of the fixed capital investment. i.e. foundations, piping, erection etc. The study was based on a wide range of chemical processes. For each factor a high, probable and low value range is given and the precise factor selection depends on knowledge of the project and the experience of the estimator. This is a major limitation of the method because it restricts use to experienced cost engineers. The factors given by Miller result in the calculation of the direct cost of a battery limits chemical plant, the fixed capital investment for the plant being estimated as a function of this cost. Miller also quotes factors for estimating green field site costs and plant extension costs.

The accuracies claimed by Miller when applying his method were $\pm$ 10% for equipment estimates and $\pm$ 15% on battery limits required for preliminary class estimating.

It is believed that estimating procedures based on this approach are employed by many contractors and built into some of their computer packages.

2.2.4. Definitive Estimates

Definitive estimates of capital cost are prepared from completely specified equipment lists, finished engineering flowsheets, plot plans, and general arrangements, together with reasonably complete site and auxilliary facilities information. There estimates have an error range of $\pm$ 10% and are in sufficient detail to be the basis for sound job cost control. They are also used for securing capital authorization while keeping engineering design costs to a minimum.

The methods employed to give definitive estimates are generally based on the factorial concept of estimating. They are detailed factoring methods generally developed by experienced and well informed company cost engineers and exist in various forms of sophistication and refinement. Not surprisingly details of these methods tend to be highly confidential.

2.2.4.1 Detailed Factorial Estimating

It can be seen from the literature survey that, since Lang, the factorial concept of estimating has been gradually advanced to provide increasingly sophisticated estimates. Since the original concept of applying a single factor to convert the equipment to an installed cost a number of different factor methods have been developed such as:

i) Application of a series of factors to different types of equipment (e.g. exchangers, pumps, columns) - as per the Modified Lang and Hand methods.

ii) Preparation of a definitive estimate for the installed cost of major 'non-standard' items such as cooling towers, boilers or special equipment and then above method (1) to the remainder of the equipment - as per the Percentage of Delivered Equipment Cost.

iii) Breakdown the flowsheet into a number of standard sub-systems (e.g. distillation operation). Apply a factor to convert the equipment cost of the sub-system to an installed cost, and then obtain the overall plant installed cost by adding up the individual sub-systems including an allowance for any none standard sub-systems - as per Miller method.

However, the most advanced form of factorial estimating in use today is the unit cost technique. Provided accurate records have been kept of previous cost experience the method gives good definitive estimates. The method requires initial design and preparation of flowsheets and the costing of major process equipment items, either from vendor quotations or index-corrected cost records and published data. Equipment installation labour is obtained as a fraction of delivered equipment cost. Costs for concrete, steel, pipe, electrics, insulation, instrumentation and painting are evaluated by take offs from drawings and applying unit costs to the material and labour requirements. Units costs are also applied to engineering man-hours. Field expense, contractors overheads and profits, and contingency are estimated by applying factors calculated from past projects.

Bauman (9) summarises the method with the following equation:-

$$C = \{\Sigma(E+E_L) + \Sigma(f_xM_x + f_yM_L) + \Sigma f_eMH_e + \Sigma f_dD\}\ f_F$$

where C = fixed capital investment

E = purchased equipment cost

E_L = purchased equipment labour cost

f_x = specific material unit cost

M_x = specific material quantity (in compatible units)

f_y = specific material labour unit cost per man-hours

M_L = labour man-hours for specific material

f_e = unit cost for engineering

MH_e = engineering man-hours

f_d = unit cost per drawing and specification

D = number of drawings and specifications

f_f = field expense factor.

The major limitation of detailed factoring is the calculation time involved. Numerous, though simple, time consuming calculations are needed to produce the final estimate.

However, this disadvantage was countered with the advent of computerisation and its associated benefits of easy data storage and retrieval and calculation speed. Although considerable effort is required to develop a sophisticated computer estimating package (56,57), many companies have thought it worthwhile to do so. Recent surveys by Liddle and Gerrard (56) and Bressler and Kuo (57,58) have noted that many packages are now available. They vary widely in intended scope and use. Some are highly sophisticated and give reliable and accurate results. Examples of these are ICI's Factest (54) and Exxon's Investment Technology (56) systems. Others are based on crude estimating techniques and tend to be integrated with process design programs (55).

2.2.5. Detailed Estimates

A detailed or firm estimate of capital investment is made from final drawings, specifications and site surveys. An error of up to ± 5% is expected on such an estimate. This class of estimate is the type made by a contractor on a lump sum bid. It's major use is in providing an accurate basis for performing cost control on the construction project as well as providing the basis for a contract.

2.2.5.1 Detailed Item Method

Table (2.2.5.1) shows a checklist of all the direct and indirect costs included in the fixed capital investment for a chemical plant. A detailed item estimate requires careful determination of each individual item on the checklist. Process equipment cost is calculated from firm delivered equipment quotations (or reliable current cost data if quotations are not available). Material needs (concrete, steel, pipe, electrics etc.) are determined from finished specifications and drawings, preferably supported by firm bids from contractors. Estimates of installation costs are derived from accurate labour rates, efficiencies and man-hour calculations. Accurate calculation of engineering, draughting and field supervision man-hours is also necessary. Complete site surveys and soil investigation data must be available to minimise errors in site development and construction cost estimates. Field expense also requires detailed determination. Quotations from vendors are obtained whenever possible in this type of estimate and extensive use is made of in-house expertise.

2.3 Identification and Selection of Research Aims

Having completed the literature review of capital cost estimating techniques it is necessary to consider the conclusions arrived at from the analysis of each individual method to help direct the research aims for the project. The primary objective is to produce a rapid step counting method for capital cost estimation and two questions now have to be considered:-

a) What needs have to be satisfied by the method(s) produced?

b) How can these needs or requirements be best achieved?

The second question (b), is discussed in Chapter 3 which describes the development of theory for proposed cost estimating models. Techniques noted in the literature survey are analysed and their shortcomings and strengths identified. From the analysis, existing theory is adapted (if suitable) or rejected and new theory proposed to enable the cost models to achieve the research aims.

The first question (a), - identification of needs, is the result of a separate ancilliary study of how and when rapid capital cost estimates are used and the limitations which exist to govern their use and accuracy.

Rapid cost estimates are used in a variety of applications in the process development function. They provide engineer/managers with cost information concerning conceptual/ballpark costs, location comparisons, and process alternate studies.

For all these applications three basic requirements exist:-

i) The estimates should be accurate to within limits that are acceptable for the situation they are being applied to so that decisions of quantifiable risk may be taken.

ii) They should be quick and easy to use so that fast decisions can be made and minimum time is spent on process and cost engineering of unprofitable proposals.

iii) They should be consistent. The variety of situations requiring cost estimates in the early stages of process development indicates a need for a series of cost estimating models to provide a consistent flexible package for producing rapid estimates from the variety and quality of information available during process development.

These points are discussed in detail.

<u>Estimate Accuracy</u>

What are acceptable accuracy limits for rapid cost methods derived from limited pre-design information?

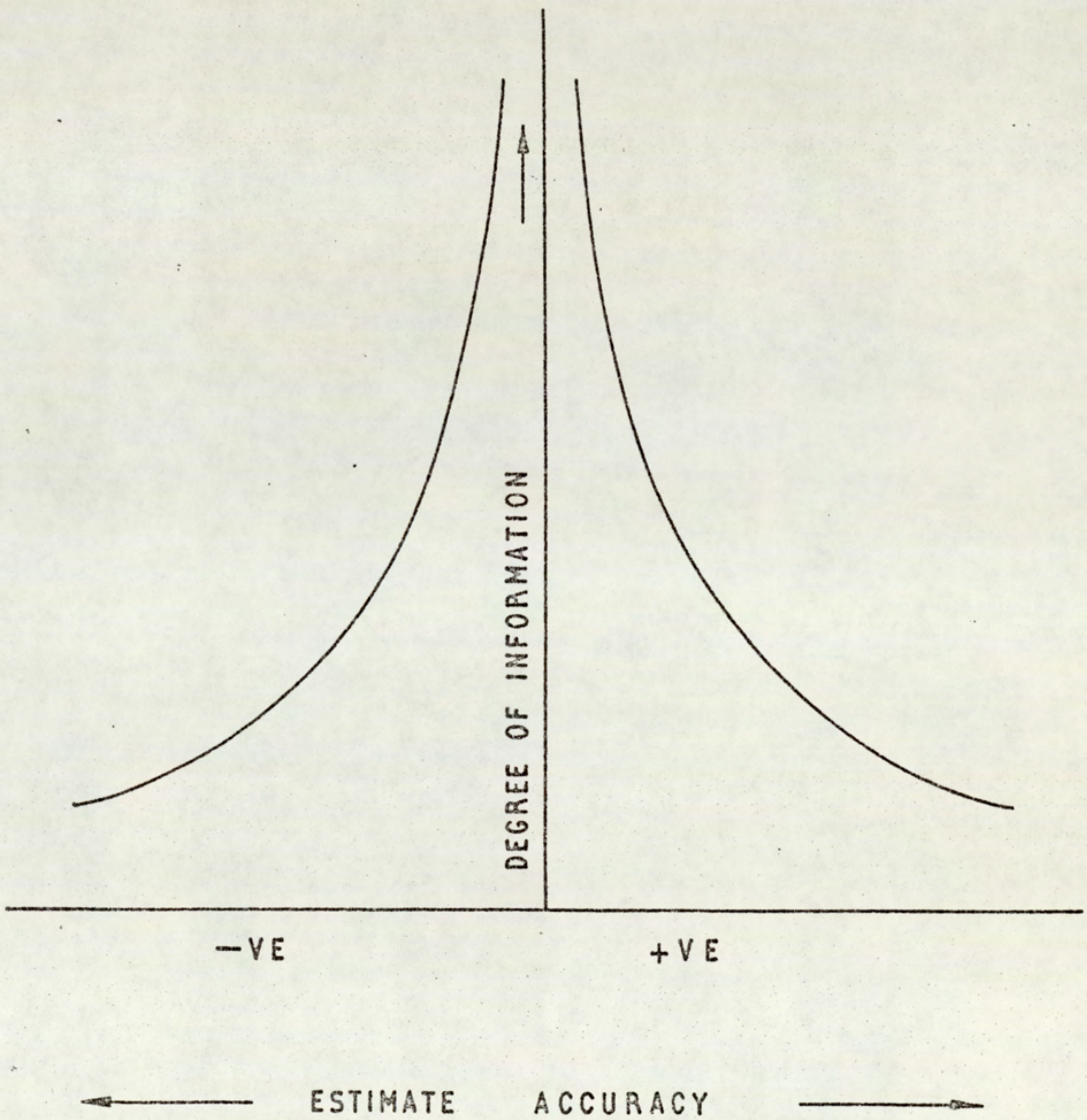
Nichols (23) has considered accuracy as a function of available information and showed graphically the relationship between probable accuracy and the quantity and quality of available information. Bauman later published a chart which was adapted from Nichols' work which related accuracy and available information for the five levels of estimate as defined by A.A.C.E.

Like Nichols, Bauman also published a plot of statistical accuracy, or limits of error, to show an envelope of variability (see Figure 2.3.1). The most probable cost is shown centered in the accuracy range, implying the probability of equal percentage variations over and under the most probable. Peters and Timmerhaus (10) state that the validity of the variable envelope has been verified within 95% confidence limits. By plotting the accuracy ranges of the estimates defined as an envelope of variability it was implied that the probable accuracy of each type of estimate could be improved proportional to the quality and quantity of information available.

Nichols (23) concluded a point very relevant to this research by noting that a large probability existed that the actual cost would exceed the estimated cost where information is incomplete or in times of rising cost trends, both situations which exist for this research. For such estimates the positive spread is likely to be wider than the negative; for example +40% to -20% for a study estimate. Nichols explained this phenomenon by the logical assumption that the positive inaccuracy in novel projects or early process development, was due to inadequate definition of the system being estimated. This conclusion highlights the need in this research for sound and consistent cost data. Since significant errors will arise from omissions of components of total project cost efforts will be made to ensure that cost

FIGURE 2.3.1

NICHOL'S ACCURACY v INFORMATION ENVELOPE.[23]



data used in the research is comparable and consistent. A checklist of the items comprising total cost is an important aid in estimating and several lists have been published in the literature which can be used as a guideline. (8,9,10,62). Table (2.2.5.1) (9) shows a comprehensive list.

The studies by Nichols and Bauman gave a first indication of the accuracy levels which may be achieved at the study level of estimating being attempted in this thesis, +40% to -20% being quoted. However, it is considered that developments since their publications no longer make these limits valid and +30% to -20% is a more realistic and achievable target.

From the literature survey it was noted that the discrepancy between positive and negative errors still persists in recent methods. Although this was originally explained by Nichols as being due to inadequate system definition, a further explanation has been proposed by Page (48) who challenges the assumption that positive and negative percentage errors are of equal importance.

He considered that because the limiting error cases for positive and negative errors were infinity and zero respectively a positive error (X_1) was of relatively less significance than a negative error (X_2) and so proposed a method for converting negative errors to their equivalent positive value and standardising the two. He proposed two equations:

$$X_1 = 100\,X_2/(100 - X_2)$$

$$X_2 = 100\,X_1/(100 + X_1)$$

- to convert a positive or negative error to its opposite equivalent. His results are presented in Figure (2.3.2).

It can be seen that at low errors X_1 and X_2 are approximately equal and at higher errors their values are widely different. From his work on accuracy measurement, Page proposed that estimate accuracies should not be represented as (+) or (-) values but rather as $\overset{\times}{\div}$ X (where X is the positive error of the estimate). The logic behind this approach is accepted here and this method will be used in this thesis to represent estimate accuracies as it enables clearer comparisons between different methods to be made.

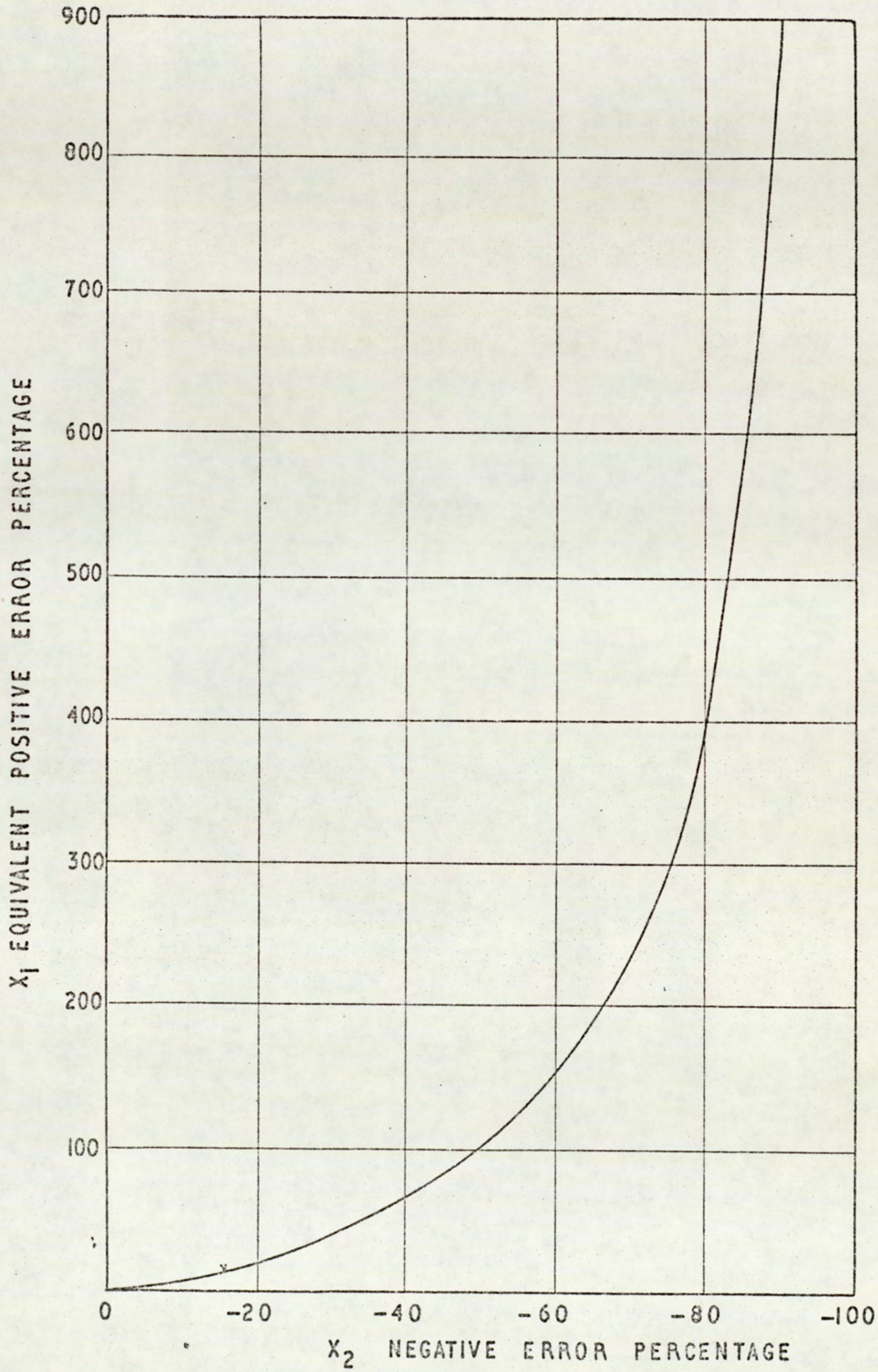
Estimate Production and Calculation

The methods derived in this research should be quick and easy to produce so that fast decisions can be made with little engineering man-hour expense spent and thus minimal cost incurred. These conditions are important in ensuring the efficient operation of the process development function and are salient points to be considered when developing rapid estimating methods.

These requirements are best met by estimates that are calculated as simple mathematical models. Step

FIGURE 2.3.2

THE EQUIVALENT POSITIVE ERROR[48.]



counting methods or functional unit methods, have proved to be well suited to this form of representation. Apart from being quick and cheap to produce when necessary (well within the guidelines given by Nichols for estimate production costs - see Table (2.3.2), they are also simple to use without the need for a high level of expertise.

Estimate Consistency

Because cost estimates are needed throughout the early stages of process development in varying accuracy and sophistication, a need exists for a package of cost estimating models derived from the same source to be developed. Such a package, ranging from very simple models used at process conception to more sophisticated ones used prior to design, would provide engineers with a consistent basis for estimating costs at different information levels.

At present no such package exists since previous workers have restricted themselves to producing singular methods applicable to a given situation. Hence, process engineers have to use different estimating methods throughout process development as their needs and available information change. The inherent fault in such an approach is that estimates derived from different sources and applicable to different situations have been compared against each other, and unless

TABLE 2.3.2

TYPICAL AVERAGE COSTS FOR PREPARING ESTIMATES * (23)

Cost of Project	Less than $1,000,000	$1,000,000 to $5,000,000	$5,000,000 to $50,000,000
Order of Magnitude Estimate	$500	$1,000	$2,000
Study Estimate	$2,000	$5,000	$8,000
Preliminary Estimate	$7,000	$16,000	$24,000
Definitive Estimate	$12,000	$35,000	$45,000
Detailed Estimate	$50,000	$125,000	$250,000

* Based on 1968 cost information.

sufficient cost engineering expertise exists to achieve compatibility between the different methods, a situation of comparing "apples and pears" exists.

Hence, the need for a consistent package of models applicable throughout the early stages of process development is an attractive ideal.

3 THEORY AND DEVELOPMENT OF NEW CAPITAL COST ESTIMATING TECHNIQUES

3.1 Introduction

Having reviewed the literature and concluded the research aims and requirements of the project it is necessary to develop the theory and estimating techniques capable of achieving those aims.

The capital cost of a process must be related to a number of variables which are considered to have a significant influence. The development of the capital cost estimating models may therefore be clearly defined and sequenced accordingly:-

i) Identification of all possible factors influencing the capital cost of a process.

ii) Selection of those factors demonstrated to be significant and capable of inclusion in a capital cost estimating model.

iii) Discussion of the selected factors; their relationship with capital cost and subsequently how they should be defined and represented for use in the estimating models.

iv) Experimental work to test the proposed models (see Chapter 4).

3.2 Identification of the Factors which Influence Plant Capital Costs

Before any attempt can be made to derive a capital cost estimating model it is essential that all the factors influencing capital costs are identified and their merits for inclusion in a costing model considered.

It is known that a great many factors exist which affect the capital cost of a plant. However, since the purpose of this research is to produce rapid pre-design cost estimating models, a restriction exists to produce models as a function of those factors which exhibit the following properties:-

- They should have a simple, basic mathematical relationship with capital cost and not be so complex and/or random in nature as to make modelling the relationship impractical.
- They should be capable of being easily defined and measured.
- They should be easily recognized and known at the early stages of process development (and so be consistent with the minimum information concept).

A further constraint exists in the selection of factors for use in developing the models - data availability. Data must be available in the literature whereby selected factors and capital cost information, in some form, are presented together so that the relationships involved can be defined and correlated in the cost estimating models.

These constraints and limitations were considered in the selection of factors for use in cost estimating models and after a thorough survey of the available cost and process data in the literature it was concluded that the conditions listed above could be met only by known fundamental and process parameters and as such any further work in the development of estimating models will be confined to examining the cost relationships of those factors which may be included in this category.

3.3 Selection of those factors considered significant in influencing capital costs

From the previous section it was concluded that any proposed cost estimating models must be based on the relationships between capital cost and known fundamental process parameters. Based on conclusions drawn from the literature survey, which considered the proposals of other workers in this field and also the constraints applicable to the available data in the literature (as listed in the previous section), it was considered that the following process parameters exhibited the necessary characteristics for inclusion in a rapid pre-design cost estimating model:-

- Phase(s) of Operation.
- Number of process steps/process complexity.
- Plant size.
- Process materials of construction.
- Process operating conditions (temperatures and pressures).

Each of the above listed variables has the required properties to enable it to be included in a rapid cost model. Each will be fully discussed so that their relationship with capital cost is understood and with that understanding a decision made if, and how best, they may be practically incorporated into a cost estimating model.

3.4 Discussion and Description of Selected Factors

Having identified and selected the variables that are to form the basis of the capital cost estimating models it is necessary to discuss how and to what extent they influence capital cost and how they should be defined in such a way that they be correctly represented in the models and accurately predict their influence on capital cost. Each of the variables is discussed in turn.

3.4.1 Process Phases of Operation

The considerable influence of process phases on plant capital costs has been noted by some of the previous workers in the field of rapid capital cost estimating and the relationship is beginning to be studied more closely as time progresses.

Because process phases determine so much about the actual processing hardware used in the plant it is probably one of the most significant governing parameters on plant costs. For example, gas or liquid phase processes would be characterised by what would be considered as standard chemical processing hardware such as towers, drums, pumps, piping. Whereas solid phase processes would be characterised by conveyor systems, and crushing. Hence, it can be seen that different phases have substantially different processing equipment requirements, to such an extent that there are essentially two quite different groups of equipment. Apart from determining equipment 'type'; phases also govern equipment volumes and it is thought also that they determine to what extent other variables such as plant size and operating conditions influence costs.

Aston (59) made a study of temperature and pressure effects on costs for different phase processes and concluded that their influences were more significant in gaseous phase dominated processes. It is also believed that the extent to which plant size - probably the most important process variable - influences plant costs is closely related to process phases. This is discussed in more detail in section 3.4.3.

Because of this considerable influence that process phases have on plant capital costs, especially in determining the equipment hardware, previous workers have reached the inevitable conclusion that in order that any estimating methods they produce have any chance of achieving their desired accuracies they must either:-

Limit their methods to estimating processes of a particular phase or 'group' of phases which employ similar equipment to process them,

or: Attempt to define or classify the different types of equipment in such a way that they could be equated.

Not surprisingly nearly all of them opted for the first alternative and limited their methods to one particular phase system.

Hill, Zevnik and Buchanan, Stallworthy, Wilson, Le Page, Page and Gore all limited their techniques to predominately gaseous phase processes, whilst Bridgwater produced a correlation for predominately solid phase plants. Only Taylor (45) fails to stipulate a phase constraint for his method by claiming that his process step scoring technique relates capital costs directly to the process chemistry without considering the type of equipment required. However, he does also state that his method is not suitable for plant involving appreciable solids handling on a large scale (above 5000 tons/year).

Page (48) however, did attempt to classify equipment by phase and so give flexibility to his method. He proposed two classes of equipment, volume and weight, which were defined basically as those items containing gas and those that did not. The classifications were decided upon because he reasoned that the presence of a gas in any unit would significantly affect its size and because gas, gas/liquid and gas/solid units are designed on gas velocities. Thus, by taking predominately gas phases or volume items, he was accounting for the worst operating conditions. Having defined his phase categories he then defined a process phase factor which was used to adjust plant throughput and account for the different equipment volumes associated with different phases.

The factor goes some way to representing phase influence on plant size and thus cost, although it must be stated that Page never fully tested his theory since no 'weight' items were present in the processes he used to test his method on. However, the factor does not attempt to define and describe the essential difference between the two identified equipment types. Whilst a laudable objective, it was considered to be too ambitious due to the difficulties that would be encountered in modelling the complex relationships involved and also in obtaining the necessary quantity and quality of information required to test any such theory proposed.

Hence, it is concluded that any estimating models produced in this research should initially be restricted to estimating processes which contain similar phase processes. It is however intended to propose a phase approach to estimating and the research will be structured to accommodate this principle. It is considered that three distinct phase groups are apparent and for the purposes of this thesis these have been identified and defined as follows:-

- Volumetric phase - processes or units which contain gas, gas/liquid or gas/solid phases.
- Liquid phase - which contain liquid or solid/liquid phases (capable of being pumped).
- Weight phase - containing solid or solid/liquid phase material (requiring solids handling equipment for transport).

For reasons already discussed in this section the above classifications have been selected because of phase influence on:-

- the main and ancillary equipment type
- equipment size
- the extent to which other variables (size exponent, temperature and pressure) influence capital cost.

Ideally cost estimating models would be derived for each of the three phase groups. Having derived these correlations it would be possible to develop a 'sub-process' approach to cost estimating such that it will be possible to estimate the cost of any process by splitting it up into its component phase units and applying the relevant correlations.

3.4.2 Functional Units

The literature review has identified problems with functional units in terms of definition and application. As this is a crucial area of study it is thought that a review of the functional unit concept would be advantageous in order that their development up to the present day be understood and the problems associated with their use fully identified.

3.4.2.1 Development of the Functional Unit Concept

At the conceptual stage of process development little information about the process is available. However, basic information which is usually available for use is the block diagram flowsheet showing the number of process steps. The functional unit technique was developed to make use of this information, and so enable cost estimates to be made without the need for equipment specification which is a basic requirement for factoring techniques.

The functional unit or process step concept was first introduced by Wessel (7) who produced a correlation to estimate labour costs as a function of the 'number of process steps' and other factors. A similar idea was developed by Hill (35) in 1956 who assigned capital costs to 'standard units' which considered complexity by having two levels of unit. In 1963 Zevnik and Buchanan (37) first introduced the term functional unit which they defined as 'a step in a chemical process' (including all ancillary equipment required for the operation of that step). The definition was difficult to interpret and apply but it served as the basis for numerous later studies and definitions produced by subsequent workers who only tampered with and modified the wording of it. Stallworthy (41) only modified the definition to 'a significant process step in the main stream or process side streams', whilst Gore (39) expanded it to describe a functional unit as

(1) all that equipment which is necessary to achieve a chemical or physical transformation of the major process stream.

or alternatively,

(2) a significant piece of equipment which carries out an operation on the main process stream.

It can be seen that the problem common to all of the above proposals for defining the functional unit is the definition of the phrase 'significant step'. What is a significant step and how is it identified? Although later workers published step counting methods, namely Le Page (43) and Taylor (45), no overall definitions were proposed. Instead lists of process steps were given. The definition of the functional unit however, has been developed via research continued at Aston

University to the following definition and understanding of a process functional unit.

3.4.2.2 Most Recent Definition and Understanding of a Functional Unit

A functional unit is all the equipment necessary to perform a significant physical or chemical transformation or the main process stream, recycle stream(s) and side stream(s). The following qualifications apply:-

(i) A functional unit describes all the equipment required to perform a unit operation.

It is not a main plant item but a block of process equipment consisting of a main plant item, ancillary equipment items (such as pumps and exchangers) and non-equipment items such as piping, instruments, civils, electrics, steel and insulation, and as such encompasses all the equipment, materials and associated costs required for the successful operation of a unit. It is usually defined as a major process step or unit operation and usually represented as a block in the conceptual flowsheet.

At the current understanding (2) examples would include distillation, absorption, reaction, but exclude pumps, heat exchangers (unless of a very high duty) storage of liquids and gases, cyclones or any equipment which is subsidiary to a unit operation. Equipment which performs mechanical operations such as crushers and grinders are included if they constitute systems in their own right and cannot reasonably be built into a unit operation. So also are mechanical feeding and discharging items but not cyclones or simple gravity settlers without mechanical gear or moving parts. Heat transfer equipment which cannot be integrated into a unit operation, with high duties or unusual heating media such as quench towers would be included, but equipment which is dependent on local economics is excluded.

(ii) Multi-stage operation only constitutes one functional unit. An early point of contention was the problem of multi-stage operations (multi-effect evaporation, multi-stage solvent extraction) and how they should be accounted for in functional unit estimating techniques. It is considered that multi-stage operations only constitute a single functional unit, which agrees with conventional thinking. This is explained by the fact that any increase in plant cost due to multiplication in one unit is not large enough to warrant that multiplication being transferred proportionally to the estimate of the number of functional units. It is considered that no problems would arise in assessing the number of functional units where multi-stage operation exists since 'by convention' a multi-stage operation would be represented on a plant flow scheme either as a single block or by units in parallel, or occasionally in series as in a multi-stage reactor with inter-stage heat transfer. These will perform the same operation and hence be taken as one functional unit.

(iii) The cost of a function unit represents the total installed fixed capital cost.

3.4.2.3 Conclusions on Most Recent Functional Unit Definition

From the literature survey of the work carried out so far on functional unit definition it can be seen that even with the considerable effort that has been aimed at this area of cost estimating, problems still exist which limit its use and acceptance as a cost estimating technique and the following criticisms have been levelled:-

i) The definitions proposed so far for the functional unit have been found to be vague and difficult to interpret. Because of this estimators are not able to accurately determine the number of functional units in some processes and so are liable to estimating errors. This presents a serious and fundamental problem to the functional unit technique of estimating. The listing of functional units helps to avoid confusion but by itself, this is not thought to be a completely satisfactory solution to the problem. It is however sufficiently helpful to warrant inclusion of a comprehensive list together with a more thorough explanation. However, the problem still remains as to what should and should not be included as a functional unit since some uncertainty remains.

ii) The number of functional units calculated for a process is subject to changes with time due to process development and consequent flowsheet changes. Hence, accurate calculation of the number of functional units may not always be possible at the early stages of process development. Unfortunately there is no acceptable way round this problem, which is a common one for any rapid cost method, since estimating accuracy is largely a function of available information. However, it should be remembered that the functional unit method is designed to estimate for the minimum process information situation.

iii) The definitions of previous workers are inconsistent when practically relating numbers of functional units to plant investment and so can lead to sizeable estimating errors. The principle of the estimating procedure is that any cost differences between individual functional units are accounted for by equiprobability theory, but no supporting proof has been attached to such an assumption. By differentiating between different type phase systems the heavy reliance placed on equiprobability assumptions are reduced, since no longer does a cost balance have to be justified between such diverse items as distillation columns and crushing systems for example. However, errors will still arise unless a functional unit definition is derived which attempts to take into account possible high cost imbalances between individual functional units in the three phase systems defined in Section 3.1.

It can be seen that a new functional unit definition is required which will overcome the difficulties outlined above, and the next section is an attempt to achieve this objective.

3.4.2.4 Re-Defining the Functional Unit

The two objectives for defining the functional unit have been set. They are:-

- that the definition should be clear and easy to interpret.
- as far as possible it should be cost consistent and should identify to within acceptable accuracy limits, equally significant portions of the overall plant investment.

The means of achieving these aims are discussed here. Since both objectives are closely linked they will be tackled together and the following approach will be used to produce a new functional unit definition:-

- Review and list all unit operations and unit processes.
- Identify and list the standard/known functional units and non-functional units on present day understanding.
- Having identified these units identify any common characteristics exhibited by them and proceed to re-define functional unit based on the findings.

A review of unit operations and processes was carried out by a thorough literature review. All the current chemical engineering magazines were searched as well as relevant texts. (A list is shown in Table 3.4.2.5.1.) This was followed by the identification and listing of 'known' functional units and non-functional units derived from published papers quoted in the literature review (Table 3.4.2.5.1). This presented a problem because the difficulty of defining what was and was not a functional unit was encountered almost immediately when moving away from the classic distillation and crushing type units. Attempts were made to overcome this problem and develop some sort of criteria for selecting a functional unit.

Firstly attempts were made to identify similar cost units or systems by searching the literature for equipment/unit cost data in order that similar cost groups could be identified and high or low cost items

TABLE 3.4.2.5.1

PRELIMINARY LIST OF 'CLASSICAL' FUNCTIONAL UNITS

Distillation
Evaporation
Compression
Reaction
Crushing
Milling
Crystalisation
Absorption
Solvent Extraction
Filtration
Cracking

noted. Such a search would serve to assist in achieving cost consistent functional units, and remove the reliance on equiprobability theory, such that

$$\frac{\text{COST 1}}{\text{COST 2}} \simeq \frac{\text{FUN 1}}{\text{FUN 2}} \text{ within acceptable accuracy limits}$$

- where FUN 1 and FUN 2 are different functional units with identical process parameters such as 'size', materials of construction, temperature and pressure.

If this cost information were available then it would be possible to produce cost weighted functional units:

let FUN 1 = COST 1 @ some base size etc., location & time

FUN 2 = COST 2 @ the same " " " "

etc. ad infinitum.

Then if a base case functional unit were determined as C_{BASE} = 1.0, N for the process could be calculated as

$$N = w_1 + w_2 + w_3 + \ldots w_N = \Sigma_1^N w = \text{cost weighted N.}$$

$$\text{where } w = \frac{\text{cost of functional unit being estimated}}{\text{cost of base functional unit}}$$

Such a method of identifying functional units and hence N for the process would be ideal. However, not surprisingly, it was found to be impractical at this level of estimating because of the several limitations described below:

Although some cost data could be found in the literature (60) it was not comprehensive enough to enable cost weightings to be undertaken for a large range of equipment. Even if the data was available it would be erroneous to base any selection criteria on equipment cost ratios alone, since total installed cost ratios are considered. It is known that the total cost of a functional unit is made up of two types of cost; variable, C_v, and constant, C_K. The variable costs are influenced by various process parameters such as size and materials of construction, process phases, internal contents of equipment and the amount of ancillary equipment associated with a functional unit. The constant cost factors cover items such as process design, detailed engineering, cost estimating etc. which are essential components of the overall cost but are fairly constant and independent factors. Hence, it may be seen that in comparing the cost of two functional units, FUN 1 and FUN 2, the relationship is:

$$\frac{C_{v1} + C_{K1}}{C_{v2} + C_{K2}} \quad \text{where } C_{K1} \approx C_{K2}$$

and not C_{v1}/C_{v2} which is essentially the equipment cost ratio. Hence, although it can be concluded that $\Delta C_{FUN's}$ is a function of ΔC_v's the cost difference is not in the ratio of the ΔC_v's but of the $\Delta(C_v+C_K)$'s which will always be a smaller ratio. The possibility of calculating an average C_K value for process functional units was examined and the literature was searched for C_v and C_K ratios in order to be able to establish cost ratios based on equipment cost differences (61). However, only one data source could be found and too many assumptions needed to be made from it to enable such an exercise to be meaningful. The problem of equipment cost data still remained also. Difficulties are envisaged in defining functional unit size and thus comparing different functional units to establish a base case. Hence, it was concluded that identifying functional units by cost comparisons was not feasible.

In the absence of cost selection criteria it was felt that the only approach left was to identify functional units intuitively. Previous workers in this field seem to have listed functional units based on their own intuition and assumptions such that the cost of a functional unit $\gg$ the cost of a non-functional unit. This seems to be a logical approach and quite reasonable under the circumstances.

These previously identified units were listed (Table 3.4.2.5.1) prior to producing a definition by identifying common characteristics. At this stage it is recognized that having rejected the possibility of cost comparisons the necessity for equiprobability theory to be applied to the functional unit method is admitted.

However, by producing a consistent functional unit definition it is still hoped to minimise the degree of reliance on such a concept.

Examination of the functional units defined at this stage for similar characteristics was greatly helped by the previous decision to classify functional units into three phase groups (Section 3.4.1) and the following points were noted:-

(i) All functional units consist of a system of process equipment made up of a main plant item, ancilliary equipment which either transports or transfers energy to the process materials and non-equipment items such as piping and instrumentation. The phase classification imposed earlier results in the functional units defined so far having similar types of main plant item, ancilliaries and non-equipment. For example, gas and liquid units tend to consist of cylindrical main items, piping, pumps, heat exchangers etc., whereas solid phase units tend to consist of a main plant item with high internal contents and ancilliary equipment consisting of conveyors, belts, screw, weighing devices etc. It can be seen that different phases require entirely different equipment types to process them. But, by classifying functional units by phase and identifying similar equipment systems and sizes, we are reducing the extent to which we have to rely on equiprobability theory in modelling. Although it is recognized that cost imbalances will still occur there will be diminished. In volumetric, liquid and weight type functional units other common characteristics also exist apart from them consisting of equipment blocks or systems.

(ii) They all require energy transfer to or from the process materials to perform their function which in turn requires ancilliary equipment to facilitate that transfer. It is noted that items previously excluded from the functional unit category such as cyclones and gravity settling chambers do not require this transfer of energy.

(iii) The main plant items of the functional units have a high amount of internals and tend to be complex items thus making them expensive because of the extra costs involved in material and fabrication, installation and design.

(iv) All functional units contain moving parts, either in the main plant item or as transport ancilliaries. These in turn require electrical power and possible tougher materials of construction and thus add significantly to the unit cost.

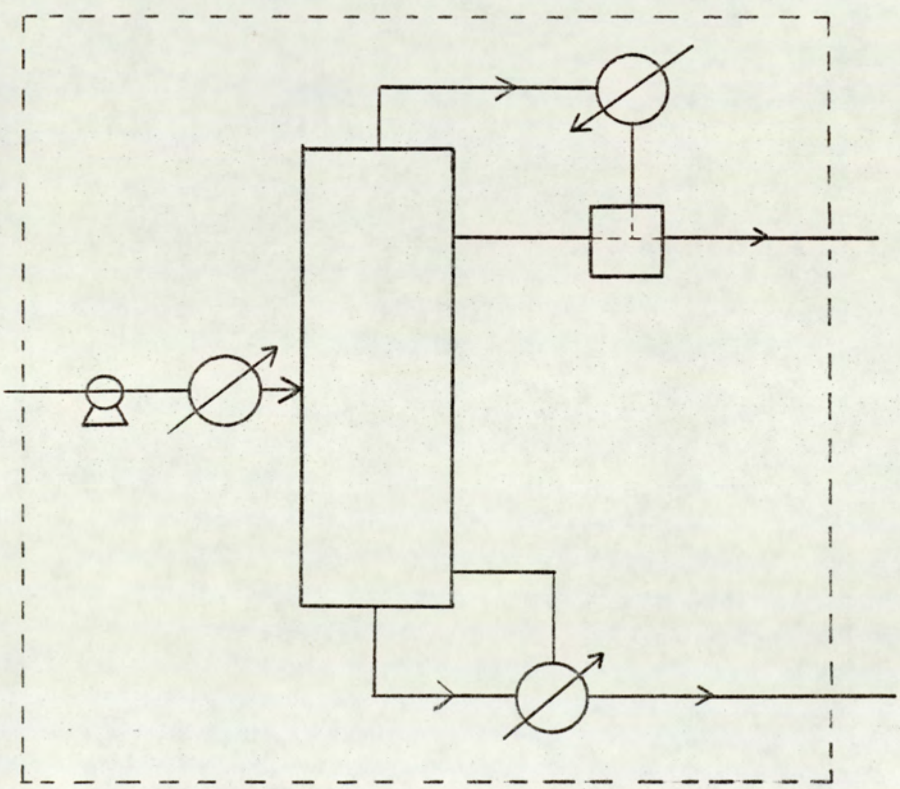
Based on these comments from the examination of functional units it is proposed to produce a new definition of a functional unit which will be used in the initial experimental work. The results from the early experimentation will be analyzed and, if necessary, the initial definition proposed in the following section modified.

3.4.2.5 New Functional Unit Definition

From the work carried out on identifying common functional unit characteristics the following definition is proposed:

> A functional unit is a significant (in terms of cost) and essential step in a chemical process which acts on the main process stream(s), recycle stream(s) or side stream(s) to produce a physical or chemical transformation (that is a change in form, substance or character) of those streams by the transfer of energy to/from the process materials. It represents the total installed unit in terms of fixed capital cost.

The above definition is illustrated by using a distillation unit as an example:-



The following points should be noted:-

(i) The distillation operation is a functional unit. The functional unit represents a block of equipment in a process and consists of all the equipment necessary for the successful, efficient operation of the unit including as the column itself, reboiler, condenser, pump and non-equipment ancilliaries such as piping, instruments, electrics, steel, civils and insulation. The cost includes all material, labour and indirect costs.

(ii) The ancilliary equipment items are not functional units but are parts of them. Although they may act on the process streams to produce a physical change in the process materials by energy transfer (e.g. preheater) they are not essential operations but subsidiary operations which improve the efficiency and control of the major operation.

(iii) Multi-stage operation constitutes a single functional unit.

(iv) Transport equipment such as pumps, blowers and conveyors does not constitute a functional unit (with the exception of compressors acting in the main process stream). Solids handling systems have previously been classed as functional units because it was considered that they were significant cost items, but the phase classification imposed results in two distinct types of unit being costed, each with similar transport ancilliaries, and so the need to include solids handling because of possible cost inbalances no longer arises.

(v) Heat transfer equipment does not constitute a functional unit when part of a system performing a unit operation e.g. reboiler, condenser, air fin cooler. However, major heating equipment which performs a unit operation and/or requires unusual heat transfer media other than normal utilities, would be a functional unit. e.g. steam cracking furnace, scraped surface heat exchanger and waste heat boiler. Refrigerated heat transfer equipment is not considered as a functional unit since standard devices using special materials of construction (usually killed carbon steel) would be used <u>but</u> the refrigerant generation equipment (e.g. expander/compressor) would be a functional unit if part of the onsite unit being considered.

(vi) Intermediate process storage equipment such as hold-up drums and tanks does not constitute a functional unit unless significant handling problems exist.

(vii) Feed and product storage are excluded from the definition here since they are consisered offsite items and for reasons described in Section 4.1.2.1. the thesis is limited to estimating battery limits costs.

(viii) The cost of the functional unit represents the total installed fixed capital cost.

It is thought that the above definition assists in achieving the set objectives of clarity and consistency. However, based on the above, Bridgwater has proposed an alternative definition which may further assist those aims:-

A functional unit is a significant step in a chemical process acting on the main process stream(s), recycle stream(s) and side stream(s) such that it either:-

a) performs a chemical reaction or mass transfer operation.

and/or b) is a separation process in which energy transfer to/from the process stream occurs.

and/or c) causes a change in particle size of the process stream.

and/or d) is heat exchange equipment utilizing fuel or unusual heat exchange media other than air, steam or water. Examples include furnaces, refrigeration and scraped surface heat exchangers,

and/or e) involves the mechanical solids other than equipment associated with any of the above (conveyance and storage (per se) being ignored).

The cost of the functional unit includes all costs and equipment necessary to perform the operation of that functional unit and represents the total installed fixed capital cost.

A brief survey of University of Aston chemical engineering undergraduates and postgraduates failed to reveal either of the two definitions shown as better and easier to understand than the other. Nearly all considered that both aided in their understanding of a functional unit.

Based on this improved understanding of functional units a list of proposed functional units for use in cost estimating models was drawn up for volumetric, liquid and weight phase processes. These are listed in Table 3.4.2.6.1.

3.4.2.6 Formulation of the Relationship Between Capital Cost and the Number of Process Functional Units

Two alternatives may be followed in establishing the relationship between capital cost (C) and the number of functional units (N) -

(a) The first assumes that $C \propto N$ and determines how good the relationship is. The relationship takes the form -

C = N x average cost per functional unit (calculated as a function of other plant parameters)

TABLE 3.4.2.6.1

DETAILED LIST OF PROCESS FUNCTIONAL UNITS

(1) GAS (VOLUMETRIC) FUNCTIONAL UNITS

Phase	Functional Unit	Non-Functional Unit
Gas	Gas-Gas Separations	Gas Storage
	Low Temperature	Gas Mixing -
	Fractionation	Jet mixing
	Adsorption	Injectors
	Hypersorption	Baffled Flow Mixing
	Fluid Char Unit	
	Gaseous Diffusion	
	Mass Diffusion	
	Thermal Diffusion	
	Reverse Osmosis	
	Gas Centrifugation	
	Gas Transport	
	Compressors	Fans/Blowers
Gas-liquid	Gas-Liquid Separation	
	Distillation	Degassing
	Absorption	Flashing
	Evaporation	Stabilisation
	Spray Tower Units	
	Venturi Scrubbers	
	Jet Scrubbers	
	Wet Walled Columns	
	Thin Film Evaporator	
Gas-solid	Gas-Solid Separation	
	Mechanical Centrifuge	Air Filter
	Electrostatic Precipitator	Cyclone
		Bag Filter
	Sublimation	Gravity Settling Chamber
	Freeze Drying	
	Gas-Solid Heat Transfer	
	Stationary Bed Systems	
	Fluidized Bed Units	
	Vacuum Rotary Dryers	
	Turbo Tray Dryers	
	Hearth Furnace	
	Shaft Furnace	
	Batch Furnace	
	Forced Convection Pit Furnace	

TABLE 3.4.2.6.1 (continued)

Phase	Functional Unit	Non-Functional Unit
	Rotary Hearth Furnace	
	Car Bottom Furnace	
	Spouted Beds	
	Flash Roasters	
	Pellet Coolers and Dryers	
	Multi-louvre Dryers	
	Moving Bed Systems -	
	Rotary Dryer	
	Rotary Kiln	
	Indirect Steam Tube Dryer	
	Indirect Rotary Calciners	
	Direct Roto Louvre Dryers	

(2) LIQUID FUNCTIONAL UNITS

Phase	Functional Unit	Non-Functional Unit
Liquid	Liquid-Liquid Separation	
	Mixer-Settler System	
	Continuous Contacting Equipment	Cyclones
	Oldshore-Rushton	
	Rotating Disc	
	Pulsed Columns	
	Centrifugal Extractors	
	Ion Exchange	
	Electrolysis, Electrodialysis and Dialysis	
		Liquid Storage
		Liquid Transport (Pumps)
	Liquid-Liquid Mixing - (with small amounts of solid phase)	
	Paste Mixing	
	Homogenizers - thickeners (mechanical) counter current decantation	
	Units - clarification (mechanical)	
	Liquid Heat Transfer Equipment	
	Incineration Furances	
	Other specialised Process Furnaces	Shell and Tube Exchangers
		Air Fins
		Simple Process and Utility Heating

TABLE 3.4.2.6.1 (continued)

Phase	Functional Unit	Non-Functional Unit
Solid-Liquid	**Solid-Liquid Separation**	
	Leaching	Gravity sedimentation
	Crystallisation	
	Ellutriation	
	Froth Frotation	
	Dense media separation	
	Filtration -	
	Cake filters	
	pressure	
	tubular	
	Continuous pressure	
	Vacuum	
	Drum	
	Horizontal table	
	Belt	
	Lifting pan	
	Centrifugation	
(3) SOLID FUNCTIONAL UNITS		
Solid	**Solids mixing/blending**	
	Single rotor mixer	
	Twin mixer	
	Tumbler mixer	
	Ribbon mixer	
	Screw mixer	
	Impact mixer	
	Turbine mixer	
	Size Reduction	**Solids Transport**
	Crushers -	Conveyor belt system
	Jaw	Screw type conveyor
	Cone	Bucket system
	Gyratory	Pneumatic system
	Smooth Roll	
	Rotary	**Solids Storage**
	Mills -	Bins
	Hammer	Hoppers
	Ball	
	Tumbling	
	Vibrating	
	Ring Roller	
	Vertical	
	Disc Attrition	
	Pin type	
	Buhrstone	
	Disintegration	
	Flash Pulverisation	
	Jet mill (fluid energy)	
	Dispersion & coltoid	
	Explosive disintegration	
	Pulverisation	
	Cutters -	
	Rotary knife cutters	

TABLE 3.4.2.6.1 (continued)

Phase	Functional Unit	Non-Functional Unit
Solid	Size enlargement -	
	Pressure compaction	
	Agglomerization by	
	Tumbling	
	Drilling	
	Sol-gel Unit process	
	Sintering and	
	Heat hardening	
	Fusion	
	Solids Separation -	
	Solids crystallisation	
	(eutectic separation)	
	Jigging	
	Screening	
	Solids Heat Transfer Equipment -	
	Solidification -	indirect heat transfer
	Table type	equipment for solids -
	belt	mainly comprises of
	vibratory	materials handling
	Rotating drum	equipment.
	Rotating Shelf	stationary tube type
	Pebble Heaters	rotating shell
	Blast furnaces	spiral/screw conveyors
		Conveyor belt

The advantages and justifications claimed for this approach are (2):-

- although the approach realises there are cost differences between different units these are averaged out by equiprobability over the process as a whole.
- with the information available at this level of estimating there is no indication of the relative costs of different items.
- the costs of the items are in any case immaterial as it is the overall cost that is being estimated.
- the approach is justified considering the accuracy level required.
- it appears to work well.

(b) Secondly, N is made an independent variable to determine if $C \alpha N$. Such that -

$C = f(N)$ and other process parameters.

There is a strong argument for adopting this approach since it is recognized that N is a crucial parameter in plant cost and thus a study of the cost v N relationship should improve estimating accuracy and show that the functional unit definition being used is consistent, such that if $C = f(N)^{1.0}$ is obtained then the definition is predicting costs consistently and equiprobability can be assumed.

The drawback to this approach however is that any errors involved in determining N may be magnified if the exponent on N is greater than 1, and therefore a sound definition and understanding of a functional unit is required. It is thought that such an understanding exists. Both alternatives are examined.

3.4.3 Process Size

The relationship between capital cost and plant size is a significant one and all of the previous workers in this field have attempted to accurately define and represent it in their methods. However, although the relationship seems apparent it is difficult to model because of problems in defining exactly what 'plant size' is, and thus representing and measuring it in a cost model. A number of alternatives exist but for pre-design estimating the only viable approach is to represent plant size as a function of the process material flows within the plant because mass balance data of varying quality is the only information which is consistently available throughout process development which serves to indicate plant size.

Literature quotes plant size based on material flows in a variety of ways; either as capacity (output), feed (input) or throughput and measures it either on a weight or volumetric basis. Hence, it can be seen that two areas of study are needed in the research to assess the best means of modelling the cost-size relationship, namely:

- the alternatives for defining plant size
- the alternatives for measuring plant size

These alternatives are discussed and recommendations made.

3.4.3.1 Discussion of the Alternatives Available for Defining Plant Size and Representing it in a Cost Estimating Model

There are three alternatives available:-

(i) Plant Capacity

(ii) Plant Feed

(iii) Plant Throughput

Each will be discussed and the relative advantages and disadvantages associated with their use identified.

3.4.3.2 Plant Capacity

Of the three alternatives listed above plant capacity has been the most widely used by previous researchers. This is easy to understand since most processes are designed to produce a given output. Thus capacity is nearly always known right from the beginning of process development and is available for use in preliminary cost estimating. This is a strong argument in favour of using plant capacity to define plant size. Since no mass balance information is required to determine its value it complies with the minimum process information constraint and is well suited to conceptual type screening study work, a primary objective of this research.

Against the benefits outlined above however limitations to using plant capacity are recognized. Early workers who used it in their correlations tended to achieve poor accuracies with the result that later researchers concentrated their efforts on the remaining alternatives of feed and throughput. They reasoned that plant capacity gave no indication of the amount of material being processed by the plant, only the amount coming out, and thus was not very indicative of plant size.

For the purposes of this thesis, plant capacity will be defined as the quantity of product(s) the plant is designed to produce in a given time (measured either on a weight or volumetric basis). The following points should be noted:

(i) Products are defined as saleable material

(ii) A standard operating year of 330 days/annum is adopted which corresponds to a 90% service factor. This time unit is the one most commonly quoted in the literature and is typical for most process plants.

(iii) For multi-product processes there are four alternatives for expressing plant capacity:

(a) main product capacity (taken as the product the plant was primarily designed to produce)

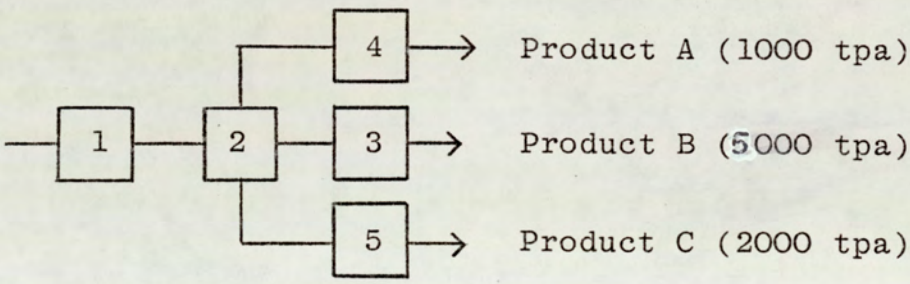
(b) total capacity (main plus by-products)

(c) functional unit weighted average of the individual product capacities

(d) average of the individual functional unit capacities.

To illustrate the weighted average definition, (c) above, an example calculation is given:

Consider the following multi-product process (theoretical);



Plant capacity is calculated as-

$$\text{Capacity} = \frac{N_A.A + N_B.B + N_C.C + N_x.Q_x}{N}$$

where N_A = number of functional units processing Q_A only

N_B = " " " " " Q_B "

N_C = " " " " " Q_C "

N_x = " " " " " Q_x "

Q_x = (in this case) = $Q_A + Q_B + Q_C$

N = total number of process functional units

$$\text{Capacity} = \frac{1x1000 + 1x5000 + 1x2000 + 2x(1000+5000+2000)}{5}$$

$$= 2300/5 = 4,600 \text{ tons/annum}$$

Hence, a representative average process capacity is calculated (which moves towards an expression of plant throughput).

An average (arithmetical or logarithmic) of the individual functional unit capacities gives average throughput. This is discounted as a viable approach however, because the mass balance information required to develop this approach would not be available at the pre-design stage of process development.

The first three alternatives a, b and c will be investigated and experimental work performed to conclude the best means of representing process capacity.

3.4.3.3 Plant Feed

Although as a general rule processes are designed to produce a specified capacity, they are sometimes specified to handle a given quantity as feed. Examples of this are waste disposal plants and gas cleaning units. Hence, it can be seen that a need exists to establish a relationship between capital costs and process feeds for such plants. However, the use of plant feed(s) to measure plant size presents problems.

In most cases, where the plant is specified to give an output of produce, the feed value is not known and so must be calculated from process mass balance information. This is not always feasible at the pre-design stage of process development. Hence, feed is not often a practical means of representing plant size generally in rapid cost methods unless:-

- process development is at an advanced stage and mass balance calculations have been performed.
- the process being considered is well established and documented mass balance data exists.
- the process is designed to handle a feed and thus the feed value is known.

Apart from calculating feed, difficulties also arise in defining and measuring it. Logically, a plant must be capable of handling <u>all</u> its material inputs; both raw material and utilities such as steam and cooling water, and so plant size should be modelled as a function of total process feed since all feed streams require storage, generation, transport and control and so contribute to the overall capital investment. However, in most cases knowledge of the process utility requirements is unavailable before process design and so would not be liable for inclusion at the level of estimating being attempted here. Hence, it is felt that utility feeds should be ommitted from further considerations. Since the amount of utility feeds is usually small in comparison with the amount of reactant process raw materials, their ommission is unlikely to make any resulting errors significant.

Hence plant feed then will be defined as a function of the process raw materials, as follows:-

- The plant feed is the total amount of reactant raw materials input to the plant per unit time (measured on a weight or volumetric basis).
- For reasons previously described the definition excludes utility feeds.
- The time unit is one year; corresponding to 330 days/annum operation.

Problems exist in measuring plant feed. If complex mixtures of unknown composition exist, determining feed value, either on a weight or volumetric measurement basis, is difficult. Examples of such feeds include crude oil, mineral ores and waste disposal streams.

Whilst noting that the above limitations exist for using plant feed as a measure of process size it is considered worthwhile to attempt to incorporate feed values into cost models for those situations which exist where plant feed is available. Experiments will be performed to assess its potential for measuring the relationship between capital cost and plant size.

3.4.3.4 Plant Throughput

The use of plant throughput to measure plant size has significant advantages over the previous two alternatives since, unlike capacity and feed values, it is a direct measure of the material flows within the plant and thus is a better indicator of actual equipment sizes. This has therefore led some of the previous research in this field to attempt to define plant throughput and use it in their proposals in order to improve accuracy (2), (46). Against the advantage of expected accuracy improvement however, there are a number of disadvantages to using plant throughput to measure plant size.

It is difficult to define just what 'throughput' is since, unlike plant capacity and feed, whose definitions are fairly straightforward, a number of alternatives exist for defining it. For example:-

(i) If individual functional unit throughputs were known and considered as the total feed and recycle flows through the unit per unit time, then plant throughput could be defined as some calculated average value (weighted, arithmetic, log etc).

(ii) In the absence of individual unit throughput data plant throughput could be calculated and expressed from a knowledge of the overall plant mass balance; feed, capacity and possibly major recycle streams.

It is considered however that neither of the options above are valid for expressing throughput in rapid cost models since both require detailed mass balance data. Such data would only be available for well established and documented processes or at the process design stage of project development, and so is exempt from use for conceptual estimating. If the required data did exist then it is likely that cost estimating would move into more sophisticated areas and techniques to make full use of it and so mass balance information should not be a requirement for rapid pre-design estimating techniques.

(iii) Having made the comments above the remaining valid alternative for defining and expressing throughput is as a function of the minimum mass balance data likely to exist; process capacity <u>or</u> process feed (not both). Some of the previous workers have attempted to do this such as Bridgwater (2) who postulated a throughput expression based on a plant reaction system such that-

$$\text{Throughput} = Q/(S^{\frac{1}{2}})$$

where Q = capacity

S = reactor "conversion efficiency" (dimensionless)

It is thought that the expression is of limited value. Equating reactor throughput to plant throughput could be misleading and represents little improvement over using capacity and feed to measure plant size. Reactor mass balance is also required to determine S.

Page (46) proposed to represent throughput as a function of plant feed and a flow factor calculated as:

$$\sum_{1}^{N} \frac{\text{Feed + Output streams for each main plant item}}{\text{Number of main plant items}}$$

on the premise that the more complex the flowscheme the higher plant throughput was likely to be. This was a valid approach but requires mass balance data for most processes to determine plant feed and complete process flowsheets if accurate throughput is to be achieved.

Finally Gore (37) obtained a throughput value by multiplying process capacity by a 'recycle factor' which was derived empirically. The theory behind the development of this factor is not described however and his work in this area should therefore be treated with caution.

All of the proposals have their limitations and although this is a difficult area due to the conflicting demands of high information requirement and low information availability, it was thought that an attempt should be made to improve, or at least equal, previous work.

3.4.3.5 Discussion of the Alternatives Available for Measuring Process Material Flows

There are two alternatives for measuring material flowrates:

(a) Volumetric measurement.

(b) Weight measurement.

Both have advantages and disadvantages as units of measurement.

Since material volume within the plant is a primary influence on actual equipment volumes and thus costs, volumetric measurement would seem to be more suitable and indicative than weight to assess relative process sizes. This is particularly true for gas phase systems where volume/mass ratios are often significantly higher than liquid or solid phases. Against this however is the fact that calculation of material volumes can be difficult in some situations when the required physical property data such as molecular weight, density and temperature and pressures is not readily available. This constraint would apply particularly if feed values were being used to measure plant size when we consider that many processes handle streams of complex and variable structures such as crude oil, mineral ores, waste gases and liquids and complex organic/aqueous mixtures. For gas phase systems there is the possible advantage of relating volumetric to molar flows.

The advantage of weight measurement is that stream weights are nearly always known without calculation since most plants are designed to handle or produce a given material weight by convention. However, as inferred earlier, weight units give no indication of relative material volume differences and so do not appear to be good indicators of process size.

From this discussion there appears to be no best method for accounting for plant size in terms of feed, capacity or throughput or for measuring it on a volumetric or weight basis.

There are good arguments for and against using all the proposed means to define and measure plant size and it is apparent that situations exist in the pre-design stage of process development where each of the alternatives would be more suitable than the others for application, depending on the information available.

As such, all the alternatives discussed for defining and measuring plant size will be explored experimentally, and a series of cost models produced to establish the relationship between capital cost and the three variables; feed, capacity and throughput.

3.4.3.6 Proposals for Modelling the Relationship between Capital Cost and Plant Size

There are two options available for modelling capital cost (C) as a function of plant size (Q) at this level of estimating. The functional unit technique is basically an extension of the exponential method of capacity adjustment (detailed in Section 2.2.1.3) in which cost is estimated as a function of plant size, such that:-

$$C = f\ (Q^a)$$

where a = plant scale factor; an overall plant exponential value defining the relationship between cost and size for the plant under consideration.

From this discussion, a number of options emerge for modelling the cost-size relationship. The first is an overall exponent method which defines the general relationship between cost and size for all plants. This is the option adopted by most previous workers and has become the established method of modelling the cost-size relationship. It does have recognised disadvantages however. It is crude and inaccurate since individual plant exponents can vary from 0.4 to 0.9 (10) and so large errors can result by applying a general average. However, it is considered that these errors can be reduced by limiting the use of an average exponent value to processes operating in the same phase. Examination of the literature revealed that, in general, equipment designed to process gaseous and liquid phase material have much lower cost-size exponents than those designed to handle solid phase material (9,10,48,62,63).

This effect can be explained by the fact that equipment handling solid phase material usually has higher proportions of internal contents and greater limits on equipment size and so any increases in size would tend to result in almost linear increases in cost correspondingly. Evidence to support this theory exists from Bridgwater's correlation for solid phase processes which predicts a much higher plant size scale up factor than those methods used to estimate fluid phase processes.

From the literature survey on exponents it was concluded that three average values for an equipment size exponent could be expected depending on the process phase of operation -

(a) Volumetric (gas) - average exponent = 0.61 (0.4-0.68)

(b) Liquid - average exponent = 0.63 (0.55-0.75)

(c) Solid - average exponent = 0.75 (0.7-0.9)

(see Table 3.4.3.5.1 for justification)

From this it can be seen that the potential errors caused by employing an overall exponent might be reduced significantly by employing scale factors applicable to the phases handled.

TABLE 3.4.3.5.1

EQUIPMENT CAPACITY EXPONENTS AS A FUNCTION OF PHASE

Equipment Items	Capacity Exponent
Volumetric Phase	
Process Furnace	0.85
Boilers	0.5
Shell & tube exchangers	0.65
Tower with Trays	0.13
Tower with Packing	0.65
Pressure Vessel - Vertical	0.65
Horizontal	0.6
Cooling Tower Facilities	0.6
Storage Tank	0.3
Horizontal Storage Pressure Vessel	0.65
Centrifugal Pump	0.52
Process Gas Compressor - 1000 psig	0.82
Air Compressor - 125 psig	0.28
Evaporators - forced circulation	0.7
vertical tube	0.53
horizontal tube	0.53
jacketted vessel	0.6
Blowers and fans	0.68
Reciprocating pumps	0.7
AVERAGE VALUE =	0.61

TABLE 3.4.3.5.1. (continued)

Liquid Phase

Filters -	Plate and Press	0.58
	Pressure Leaf (wet)	0.58
	Pressure Leaf (dry)	0.58
	Rotary Drum	0.63
	Rotary Disc	0.78
Crystallizers -	Growth	0.65
	Forced Circulation	0.55
	Batch	0.7
	AVERAGE VALUE =	0.63

Weight Phase

Crushers -	Cone	0.85
	Gyratory	1.2
	Jaw	1.2
	Pulverisers	0.35
Mills -	Ball	0.65
	Roller	0.65
	Hammer	0.85
Hoppers -	Conical	0.68
	Silos	0.9
Driers -	Drum	0.45
	Rotary	0.46
	AVERAGE VALUE =	0.75

The use of general exponents according to phase is consistent with the sub-process (by phase) approach proposed in Section 3.4.1. Average phase exponent values can be determined via experimental work to assess size effects on <u>installed</u> costs. The results can be applied to any type of process as follows;

Consider a process:

N_1 = number of gas functional units (as defined in 3.4.1)

N_2 = number of liquid functional units (as defined in 3.4.1)

N_3 = number of solid functional units (as defined in 3.4.1)

then, process exponent values might be expressed as:

$$\frac{N_1.a + N_2.b + N_3.c}{N}$$

where a = gas phase exponent

b = liquid phase exponent

c = solid phase exponent

N = total number of process functional units ($=N_1+N_2+N_3$)

Therefore the potential errors involved in using a single exponent for all processes can be reduced by utilising phase sub-processes and their calculated exponents (individually rather than collective).

In addition to the phase refinement for overall exponent values it was also suggested by Bridgwater that a further sophistication could be investigated namely, determining scale up factors for the essentially fixed and variable components of plant costs (discussed in Section 3.4.3) with size. This approach could not be developed due to inadequate data in the literature.

An alternative approach is to derive an individual exponent to define the cost-size relationship for a specific plant.

This approach was studied by Page (46) who proposed a cost weighted capacity exponent whereby he related the overall plant exponent to the individual exponents of the equipment items which made up the plant (detailed in Section 2.2.2.7). The approach does have recognised limitations when used for pre-design estimating:-

- At process conception (box diagram stage) equipment types may not be identified and so the technique could not be used.
- Although unlikely, the method could become obsolete as new equipment designs with exponents unknown reach the industrial market.

Even so, it is considered that the approach was a valuable piece of research in the area of expressing cost as a function of size. The approach is not applicable to functional unit estimating however, which is designed to estimate installed costs directly. Page's method was based on determining equipment costs (factored to installed costs) and so could utilise the mass of published equipment cost exponent data to calculate an overall plant exponent. No equivalent data exists however for relating installed equipment/unit costs to size and so the base data required to develop an overall exponent from a knowledge of the individual process units is not available. Hence, regretfully, the concept of deriving individual plant exponents based on a knowledge of process component units was unable to be carried out.

This left the first alternative of using overall plant exponents as the method of measuring the cost-size relationship and this will be adopted for this research.

3.4.4 MATERIALS OF CONSTRUCTION

Capital cost is significantly influenced by materials of construction and as such many previous workers - Hill, Zevnik and Buchanan, Stallworthy, Le Page, Wilson, Page and Bridgwater - have attempted to define a cost-materials of construction relationship in their methods. It is particularly desirable to define a relationship between installed capital cost and materials of construction for use in rapid cost models because the process materials of construction are usually known, or can be "guesstimated" from a knowledge of the chemical and physical properties of the process materials, in the early stages of process development. The relationship between installed capital cost and materials of construction are discussed here with the aim of producing a means of incorporating the relationship in a capital cost estimating model.

3.4.4.1 Representation of Process Materials of Construction in a Cost Estimating Model

There are four options available for representing a material of construction effect in a capital cost estimating model:-

(i) as a function of the chemical properties of the process materials handled.

(ii) as a function of process operating conditions (i.e. temperatures and pressures).

(iii) as a function of both of the above.

(iv) as a function of the actual materials employed.

Each of the options is discussed.

(i) Aston (59) postulated an empirical relationship to derive a materials of construction effect as a function of the corrosivity of the process materials, such that;

$$Fm = a + b.Fc$$

where Fm = material of construction factor

Fc = corrosivity factor ; defined as a function of the process materials

- Organic/Neutral, $F_c = 0$
  Alkaline/Aqueous, $F_c = 1$
  Acidic, $F_c = 2$

a,b = constants (undefined)

No explanation is given as to what the theory and parameter values are based on and how the above relationship may be used in a cost model and as such the correlation is thought to be of little value.

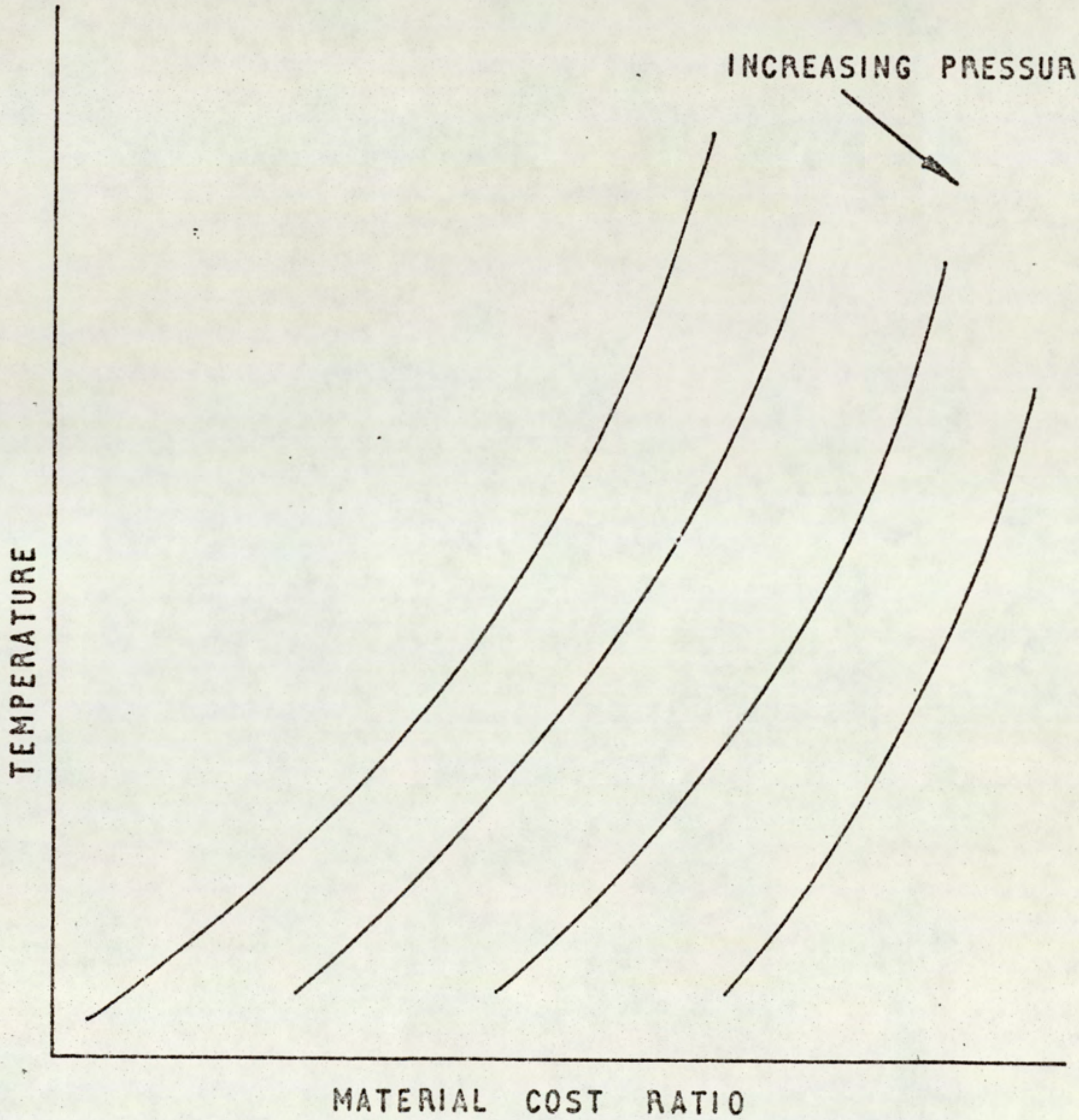
(ii) Hill, Zevnik and Buchanan, Stallworthy, Wilson and Le Page attempted to relate material of construction to process temperatures and pressures by adopting a multiplicative factor approach. The validity of this method was questioned by Page (48) who recognised that the three variables; materials of construction, temperature and pressure were not independent of each other and so a change in one could produce a change in the other i.e. inter-dependence of variables. If temperature and pressure were altered beyond certain levels it may cause a material of construction change but the change is usually a step change whereas the basic assumption of using a multiplicative approach is that the variables are continuous functions. Also if temperature and pressure are included as separate variables in a cost model should ΔP and ΔT be accounted for here, since by doing so they are accounted for twice. This is an argument for not including a material of construction factor at all if the above situation existed. However, other variables such as corrosivity would be neglected by such an ommission.

Having identified the above limitations, Page considered an additive approach but this was not developed and he reluctantly proposed his own multiplicative approach in which he referred materials of construction to a common base using a material cost ratio defined as:

> the cost of the plant being estimated divided by the cost of the same plant constructed in mild steel and operating at 1 atm and 0-100^{o}C. No explanation is given for defining the base case as such.

A graph of material cost ratio v Temperature, Pressure was proposed (see Figure 3.4.4.1) so that the three variables find their own inter-relationships instead of imposing a relationship upon them as other methods had done. This was rejected as detailed information of

FIGURE 3.4.4.1
MATERIAL RATIO[48.]



(FOR CONDITIONS ABOVE THE BASE CASE ONLY) process temperatures, pressures and materials of construction would be required.

From his study in this area, Page concluded by adopting Stallworthy's approach. He recognised that it involved all the limitations he had identified but concluded that it was superior to anything else presented in the literature because it was based on a sound and comprehensive data source.

(iii) It is considered that both of the previous options discussed are inadequate and essentially incorrect since <u>both</u> corrosivity and operating conditions are equally significant in determining process materials of construction and so determination of a material of construction factor must be based on both variables. Either one in isolation is impractical. However, to attempt to define materials of construction as a function of both variables mathematically is extremely difficult due to the many complex relationships present and random step changes involved. Substantial and reliable data is necessary which is not freely available and hence this approach has not been adopted.

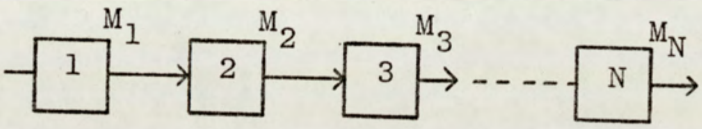
(iv) The premise upon which previous research has been based is that detailed materials of construction are not known in the early stages of process development. Hence, the attempts to represent and define material of construction effects as a function of more fundamental, known process variables. It is considered that if any progress is to be made in this area that this basic premise should be altered. To deduce a realistic material of construction factor from a knowledge of process material chemical properties and operating conditions would require far more process information and/or guesstimation than would be required for making a simple assessment as to what materials of construction will actually be used in the process.

Hence, it is proposed to develop material of construction factors from a knowledge of the actual process materials employed. Such factors would have advantages over the previous approach since, instead of being based on obscure mathematical relationships, they would be based on actual material cost ratios. It is intended that material of construction factors should be derived to assess the effect of different materials of construction on the installed capital cost of a process. This would make them easy and simple to apply to rapid cost estimating models. The derivation of such factors can be achieved (though in crude form) by utilizing available information in the published cost engineering literature, namely;

1. Material cost ratios for equipment.
2. Lang type factors.

Their development is described here:-

If the process being estimated is considered as follows;



where M = material of construction of functional unit

N = number of functional units.

Then, if the material of construction for each functional unit is known, it is possible to represent the effect of materials of construction on installed capital cost using the following procedure:-

· Collection of relative material of construction cost data to deduce;

$$f_m = \frac{\text{DEC FUN}_m}{\text{DEC BASE}_m}$$ i.e. comparison of material cost

where f_m = material of construction cost ratio for equipment.

DEC FUN_m = delivered equipment cost of the functional unit being estimated.

DEC BASE_m = delivered equipment cost of the same functional unit constructed in a chosen base case material.

The base material will be taken as carbon steel since it is the most common and cheapest material used in the processing industries. Minimum alteration for materials of construction will thus be required for the majority of processes. For the base case of carbon steel, $f_m = 1.00$. Material of construction cost data is available in the literature (52), for a range of materials.

The next step is calculation of a material of construction factor for the functional unit. Using f_m, a material factor, F_m, to assess the effect of materials of construction on the total installed cost can be developed by relating delivered equipment cost to total installed cost using Lang factor theory. Considering the base case, a functional unit of carbon steel

$$TIC_{BASE} = DEC_{BASE} \times L$$

where TIC_{BASE} = total installed cost of functional unit

DEC_{BASE} = delivered equipment cost of basic unit

L = appropriate lang factor for functional unit.

and so, $DEC_{BASE} = TIC_{BASE}/ L$

starting from the following relationship

$$TIC_{BASE} = DEC_{BASE} + \text{other costs (which are considered independent of fm).}$$

then, $$TIC_{BASE} = \frac{TIC_{BASE}}{L} + \left(TIC_{BASE} - \frac{TIC_{BASE}}{L} \right)$$

For functional units with other materials of construction

$$TIC_{FUN} = DEC_{FUN} + \text{Other costs (which remain constant)}$$

$$= \frac{TIC_{BASE} \times fm}{L} + \left(TIC_{BASE} - \frac{TIC_{BASE}}{L} \right)$$

$$= TIC_{BASE} \left(\frac{fm}{L} + (1 - \frac{1}{L}) \right)$$

Therefore $$\underline{\underline{\frac{TIC_{FUN}}{TIC_{BASE}} = Fm = \frac{fm}{L} + (1 - \frac{1}{L})}}$$

where Fm, the installed cost material of construction ratio, is defined as the ratio of the totalled installed cost of a functional unit compared to its cost if constructed in carbon steel (the base case).

This principle may be extended to deduce an overall process factor by employing an arithmetical averaging technique such that:-

$$MOCFAC = \frac{Fm(1) + Fm(2) + \ldots + Fm(N)}{N}$$

where MOCFAC = overall process installed cost material of construction ratio.

Fm 1,2..N = Installed cost material of construction ratio of the individual functional units in the process.

N = number of process functional units

$$= \frac{Fm(1)}{L} + (1-\frac{1}{L}) + \frac{Fm(2)}{L} + (1-\frac{1}{L} + \ldots \frac{Fm(N)}{L} + 1-\frac{1}{L}$$

$$= \frac{\Sigma_1^N \frac{Fm}{L} + N\,(1-\frac{1}{L})}{N}$$

Therefore $$\underline{\underline{MOCFAC = \Sigma_1^N \frac{Fm}{L.N} + (1 - \frac{1}{L})}}$$

Hence, MOCFAC may be mathematically defined.

3.4.4.2 Conclusions and Proposals for Modelling the Relationship between Capital Cost and Process Materials of Construction

It is recognized that the above theory is somewhat crude but it is thought that the concept involved is sound and worthy of experimentation. The modelling procedure will be as follows:-

i) Identification of all process materials of construction present in the process being estimated.

ii) Identification of dominant material of construction for each functional unit.

iii) Using published fabricated material cost ratios, (50), fm, calculate Fm (the functional unit material of construction factor) as described, and subsequently MOCFAC (the overall process material of construction factor) to use in cost models via Lang factors. Different Lang factors would ideally be used according to the process phases. To derive an overall process material of construction factor the process being estimated should be split up into phase sub-processes such that:-

$$\text{MOCFAC (volume phase)} = \sum_{1}^{N} \frac{Fm}{fvNv} + \left(1 - \frac{1}{fv}\right)$$

where Nv = number of volumetric (fluid) functional units

fv = volumetric (fluid) phase lang factor.

Similarly, for liquid and weight phase items.

Hence, $\text{MOCFAC process} = \dfrac{\text{MOCFAC volume + MOCFAC liquid + MOCFAC weight}}{3}$

The procedure proposed is considered feasible but limitations are recognized:-

. The approach is lengthy and involves a lot of pre-estimating calculation.

. Substantial information is required to calculate MOCFAC the overall process material of construction factor, namely -

process materials of construction - which may not always be available. Guesstimation may be necessary which may lead to estimating errors.

Material cost ratio data - available in the literature for commonly encountered materials (50, 52, 60). Exotic materials (e.g. tantalum) may cause problems as cost data is scarce.

Lang type factor data - which is freely available in the literature.

- The method has many of the disadvantages associated with Lang factor theory.
- The method assumes that fm and Lang factors are accurately known.

Due to time constraints and the absence of sufficient and reliable data on liquid and solid phase processes, only gas phase processes were investigated in the experimental work. Since most of the cost data referred to petrochemical processes the Lang type factors, (L) supplied by Hand (27) are applicable to the research (see Table 2.2.2.2.1). It was decided to adopt a value of four for the experimental work. Gallagher (52), Jelen (50) and Popper (60) publish material cost ratio (fm) data (see Table 3.4.4.1).

Particular attention was given to determining an accurate ratio for stainless steel as this was the second most common material encountered after carbon steel. From a literature search an average value of 2.15 was obtained (see Table 3.4.4.2). Using the above data for L and fm the installed cost ratio (Fm) was calculated (see Table 3.4.4.3) and subsequently the overall process material factors (as quoted in the cost data tabulation in Appendix III).

TABLE 3.4.4.1

GALLAGHERS RATIO OF ALLOY/CARBON STEEL COSTS (fob price) (52)

Material of Construction		Ratio
Carbon Steel		1.0
Cast Steel		1.0
Stainless Steel,	Type 410	2.1
	Type 405	2.25
	Type 304	2.5
	Type 316	3.0
	Type 310	3.25
Monel		4.0

TABLE 3.4.4.2 (50) (52) (60)

STAINLESS STEEL (316)/CARBON STEEL MATERIAL COST RATIOS

Data Reference	Equipment	Ratio
Popper p.87 (60)	Box Heater	1.75
p.87	Cylindrical Heater	1.5
p.88	Heat Exchanger	1.54
p.89/158	Airfin Cooler	1.93
p.90	Process Vessels	2.45
p.91	Trays	2.7
p.92	Pumps (centrifugal)	1.93
p.92	Pumps (reciprocating)	2.1
p.106	Tanks	3.2
p.112	Agitators	1.17
p.114	Pan Dryer	1.6
p.114	Rotary Dryer	1.6
p.114	Filter	1.25
p.150	Shell	2.6
p.158	Pumps	1.92
p.158	Pumps	1.64
Jelen (50)	Piping	3.77
Gallagher (52)	Heat Exchanger	3.0
Gallagher	Fabricated equipment	3.0
	AVERAGE VALUE =	2.15

TABLE 3.4.4.3

INSTALLED COST MATERIAL RATIOS (Fm)

Calculated using $Fm = \left(\frac{fm}{L} + \left(1 - \frac{1}{L}\right)\right)$

Material	fm	Fm *
Carbon Steel (base case)	1.0	1.0
Stainless Steel 410	1.5	1.13
304	1.8	1.2
316	2.15	1.29
Monel	2.5	1.38
Aluminium	1.5	1.13
Titanium	7.89	2.72
Glass lined carbon steel	3.17	1.54
Rubber lined carbon steel	1.64	1.16
Phenolic lined carbon steel	1.23	1.06

This concludes the discussion of the cost-material of construction.

3.4.5 PROCESS OPERATING CONDITIONS

The severity of the process operating conditions has an impact on capital cost. Most of the previous workers in this field have attempted to measure this impact by defining a relationship between cost and the process temperatures and pressures. This has proved difficult and not entirely satisfactory because a number of complex and contrasting inter-relationships are involved. The two variables, temperature and pressure, are discussed here in order to understand how and to what extent operating conditions affect cost, and so how these effects can be modelled in a realistic manner.

3.4.5.1 Effect of Temperature on Cost

Temperature of operation affects capital cost in a number of ways:

(1) Predictable Effects - whereby the temperature effect on cost is predictable such that as the temperature increases costs may be expected to increase (and vice versa) due to:

- increased insulation requirement
- possible change in materials of construction

. possible increases in instrumentation for safety reasons

. the amount and complexity of the ancilliary heat transfer equipment will increase.

From this it is possible to tentatively conclude that as a general rule capital cost will increase as temperature increases. However, this relationship is complicated by a second set of effects.

(2) Variable Effects - which are complex, not generally predictable and are governed by the following functional unit characteristic:

. the type of unit operation:

Depending upon the operation, temperature fluctuations within the unit can affect cost in different ways. For example in most reactions reaction rate increases with temperature increase. This reduces the reactor size required but inversely may necessitate more expensive materials of construction. Some reactions may behave differently with analogous consequences. In gas-phase systems increased temperature causes either increased volume and hence increased size and cost, and/or increased pressure and hence reduced size and cost but additional pressure vessel costs. These effects are different again for liquid phase and solid phase systems. The effects are therefore complex and often unpredictable.

- <u>the size of unit operation being considered</u>:

  Gore (39) postulated that plant size influenced the cost-temperature relationship due to the costs of insulation and installing heat recovery equipment in large plants being greater than for small plants. This would result in temperature exerting greater influences on cost as plant size increased. No supporting evidence was given to this theory.

- <u>phase of operation</u>:

  The extent to which temperature influences cost is dependent on a unit's phase of operation because temperature fluctuations/conditions will influence unit pressures and volumes to different extents (assuming restricted volumes). Gaseous phase units would be most susceptible with the temperature influence on pressure and volume affecting size, material thickness, possibly necessitating a change in material of construction and increasing instrumentation requirements for safety reasons. Liquid phase units would be much less susceptible to these effects until liquid boiling point was reached when a step change would occur and the above effects would become noticeable. Negligible temperature effects due to pressure increase would be expected for solid phase units even at melting points and over, since little change in material volume would result.

In conclusion, although the constant temperature effects listed earlier would apply to all units, the extent to which temperature influences cost is also a function of phase. This is supported by Aston's study of cost-temperature relationships in which he concluded that temperature had greatest influence of gas phase processes and least on solid phase (59) (see Figure 3.4.5.2.1.1). This further supports the initial decision to adopt a phase approach to cost estimating. Cost-temperature relationships should be defined for volumetric, liquid and weight phase processes as this, together with the resulting localisation of temperature conditions by sub-processes in multi-phase processes giving greater temperature profile representation, will give improved estimating accuracy.

. <u>Ancilliary Heat Transfer Equipment</u>

Heat transfer equipment is common to many functional units and so can have a significant influence on unit cost. The cost of this equipment is a function of the actual temperatures and temperature changes needed in

FIGURE 3.4.5.2.1.1

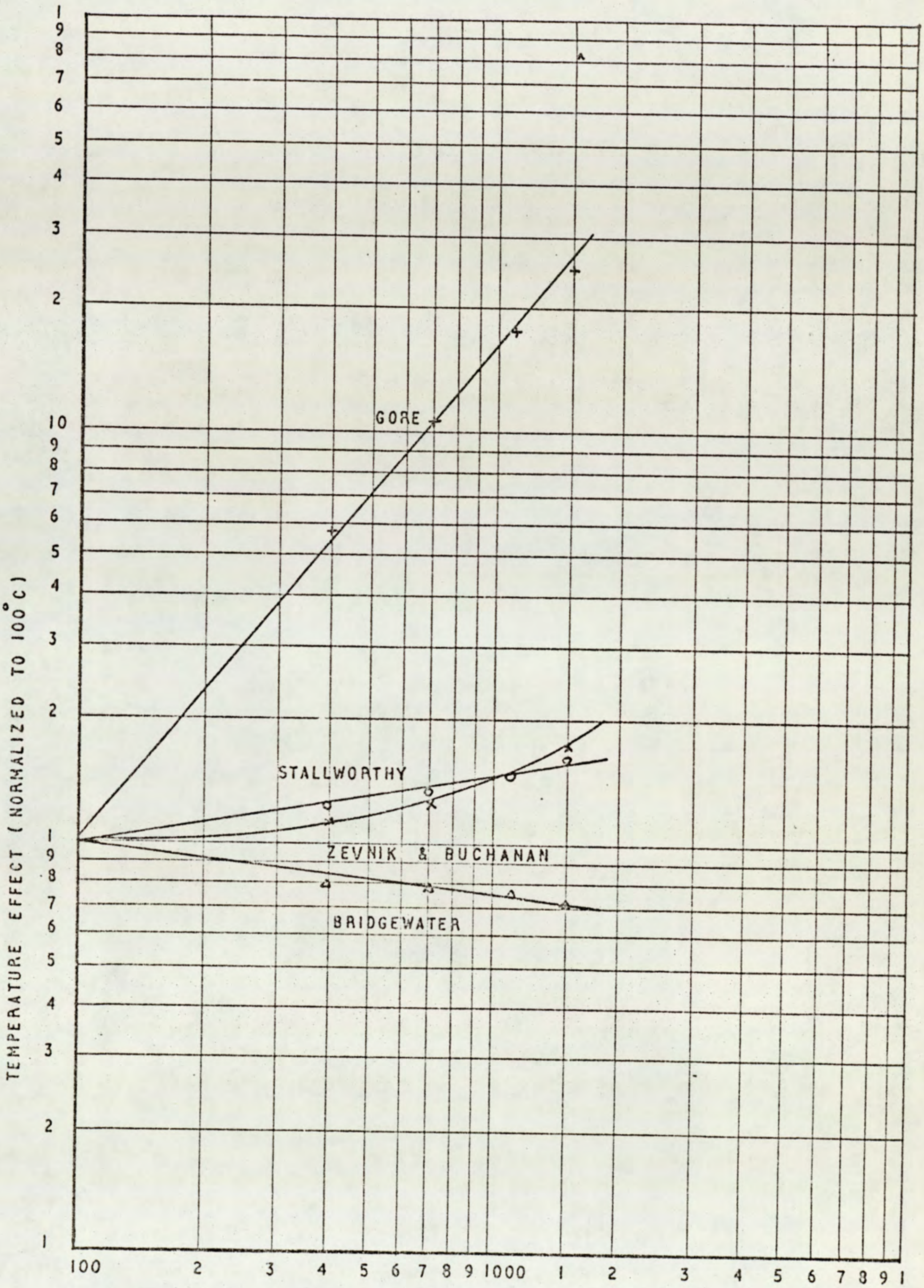


FIG 2

COMPARISON OF TEMPERATURE EFFECTS [59]

the unit. Hence, it can be seen that the temperature influence on cost may be largely due to unit temperatures determining the heat transfer equipment requirement.

The cost of the ancilliary heating equipment is a function of both the maximum and minimum temperature of the process materials. This governs the type of equipment required. High temperatures will require specialised equipment such as furnacesto attain them, whilst normal/standard temperatures would require only standard shell and tube type equipment. Sub-ambient temperatures would require refrigerant generation equipment (usually outside battery limits) but the heat transfer equipment required would be standard shell and tube type with special materials of construction (usually Killed Carbon Steel), or, in the case of gas refrigeration, compression and expansion facilities. The latter type equipment is not considered here.

From these considerations then, two equipment ranges have been identified which are a function of maximum/minimum temperature. The lower range will be defined as equipment designed to operate at temperatures of less than 200°C. This temperature range is selected because standard heat transfer equipment ancilliaries operate within it (e.g. shell and tube, G fins, Air fins) using process utilities such as steam and cooling water as heat transfer media. Such utility equipment would be an integral part of a functional unit. The upper range refers to equipment operating above 200^{o}C which generally requires unusual heat transfer media (non-utility) to achieve the required temperatures. (e.g. furnaces). Such equipment is specialised and sophisticated requiring exotic materials of construction, sophisticated controls and generally large heat transfer surface areas and thus size. Such equipment has already been classified as a functional unit in section 3.4.2. Refrigeration heat transfer equipment is not classed in the upper range but refrigerant generation is (if part of on-site battery limits unit.

A second effect is the required temperature change in the process materials (ΔT). In each of the equipment ranges the cost of the equipment would primarily be a function of the required heat transfer surface area and thus equipment size which is a function of the required ΔT. (and heat transfer coefficients).

This completes the discussion of the cost-temperature relationship. From the comments made it can be seen that the relationship is a complex one involving numerous process variables. However it is possible to conclude that capital costs tend to increase as temperatures and temperature differences increase. What follows is an attempt to simplify those relationships so that they can be modelled and the relationship between cost and temperature described.

3.4.5.2 Representation of Temperature Effects on Capital Cost

Having discussed the cost-temperature relationship it is necessary to investigate ways to accurately represent the relationships described in a cost estimating method. This requires the temperature variable to be defined and measured in such a way that it achieved this aim. The previous discussion examined the effect of a functional units temperature profile on its cost and concluded that two temperature variables; maximum process material temperature and temperature changes, were the most influential on cost. Both variables must be examined and cost modelled as a $f(T_{max}, \Delta T)$.

Having concluded this, a method for representing these variables in cost models has to be derived so that the cost-temperature relationship can be measured. A number of alternatives exist for doing this but two limiting constraints have to be considered. These are:-

- Which T_{max} and ΔT to use? In a functional unit a number of Tmax's and T's exist as, by definition, it consists of a main plant item and ancilliaries.

. The amount of information available. This has a direct bearing on the above. It is unlikely that the information required to Tmax or T of ancilliary equipment would be available in pre-design work and so Tmax and ΔT measurements will have to refer to main plant item equipment. Having concluded this however the problem still exists of how to define Tmax and ΔT for a whole process plant. The literature survey revealed that Tmax and ΔT data was not available for individual functional units in a process. If it was average, process values for Tmax and ΔT could be derived (either arithmetic, log, geometric or harmonic) to give a fair representation of the process temperature profile. In the absence of the required data, guesstimation of unit temperatures was considered by using available physical property data such as boiling points, but this was rejected, because such data in isolation would not be capable of predicting process material temperatures, as other variables, such as pressure, had to be considered.

It was concluded therefore that the temperature variable definition was restricted by the minimum information concept being applied and that process temperature conditions had to be represented and defined without knowledge of individual unit temperatures but from other information which was consistently available - maximum and minimum process temperatures. The same conclusion was reached by previous workers.

The alternatives which exist for expressing representative Tmax/ΔT values from this information are limited and consist of:-

- using maximum ($Tmax_p$) and minimum ($Tmin_p$) process temperatures such that:-

  Tmax term = Tmaxp

  ΔT term = Tmaxp - Tminp

- using simple average values of Tmaxp and Tminp to define:-

  Tmax term = (Tmaxp-Tminp)/2

  ΔT term = expressed possibly as (Tmaxp-Tminp)/2 - Tminp

- using weighted average values (as proposed by Bridgwater (2)) such that:-

  Tmax term = Tmaxp x (n/N)

  ΔT term = expressed possibly as Tmaxpx(n/N) - Tminp

  where n = number of functional units operation at > Tmax/2

  N = total number of functional units.

The problem of expressing temperature conditions from Tmax and Tmin was not studied seriously as it is considered that none of the above methods offers a practical solution to the problem and the author can propose none. The use of Tmax and Tmin or some simple average of the two would be an innaccurate statement of the temperature profile for most processes whilst the weighted average approach, though logical, has serious problems (discussed in Section 2.2.2.6). To use any of these methods could lead to significant estimating errors. As such, it was concluded that no accurate way of representing temperature conditions from minimum information existed. The options available for doing so are inadequate and so, combined with the further complications of modelling cost as a function of Tmax and ΔT together, plus the problem of units of measurement for temperature, it would seem that modelling the cost-temperature relationship in a significant manner is not feasible.

At this point it was considered whether this was as significant a problem as it appeared. It was thought not to be because:-

. It has been decided to adopt a phase sub-process approach to estimating. Since similar cost-temperature relationships are thought to exist in each phase group it is likely that the absence of a temperature variable in cost models may not be as significant as would appear. Previous workers' results on the cost-temperature relationship would seem to support this assumption.

- Previous workers' results also indicate that temperature influence on cost is nowhere near as powerful as the other variables of size, number of process steps and materials of construction which are being included in the cost models.
- Abnormal temperature conditions are already being accounted for since the heat transfer equipment required to produce them has been classed as a functional unit. Hence, it is likely that the introduction of a temperature variable would lead to over-estimating the effect of temperature on cost.
- Temperature influence on materials of construction has been accounted for.

3.4.5.3 Effect of Pressure on Cost

Since temperature and pressure are being considered under the common heading of process operating conditions the same structure for discussing the cost-pressure relationship and its subsequent representation in a cost model will be adopted for pressures as was used for temperature. As with temperature, the relationship between cost and pressure is a complex one and it is possible to identify the same predictable and variable effects on cost, listed as follows:-

1) Predictable Effects - whereby the pressure effect on cost is predictable such that as pressure increases costs may be expected to increase due to;

. increased equipment thicknesses or possible change to more costly material to cope with increased pressure.
. increased control for safety reasons.
. possible necessity for compression equipment to attain high pressures not obtainable through increased temperatures or pumping.

However, unlike temperature, it is not possible at this point to conclude tentatively that pressure increases out as a general rule to increase costs, since as unit pressure is increasing the unit volume of the process materials are decreasing and so act to reduce plant costs. Apart from this initial complexity in the cost-pressure relationship variable effects exist which act to complicate the relationship further.

2) Variable Effects - which are random and governed by

. the type of unit operation

Depending on the type of unit operation a pressure increase can either increase or decrease cost. For example - reaction. By Le Chatelier's principle a pressure increase may drive a gaseous reaction away from the optimum which would result in larger recycle streams and so increase cost. Conversely, the reverse effect could be noticed.

. <u>phase of operation</u>

Gas phase systems are more sensitive to pressure changes than liquid or solid phases which are incompressible and so not subject to the same volume fluctuations. Hence, the extent to which pressure influences cost is a function of phase, as evidenced by Aston's study (59) (see Figure 3.4.5.3.1.1).

. <u>ancilliary pressure equipment</u>

As with temperature, the cost of ancilliary pressure producing equipment, i.e. compressors, is a function of two variables; maximum pressure and the system pressure differentials (ΔP). Again two ranges seem to exist; a lower one where small centrifugal compressors, gas turbines, fans and blowers can be used and an upper one where large centrifugal and reciprocating compressors which are systems in their own right are required. These latter types have been classed as functional units. Unfortunately no cost data could be found in the literature to enable the lower and upper ranges to be defined. In the absence of this data assumptions have had to be made as to whether or not some of the compression systems encountered in the experimental work were functional units.

FIGURE 3.4.5.3.1.1

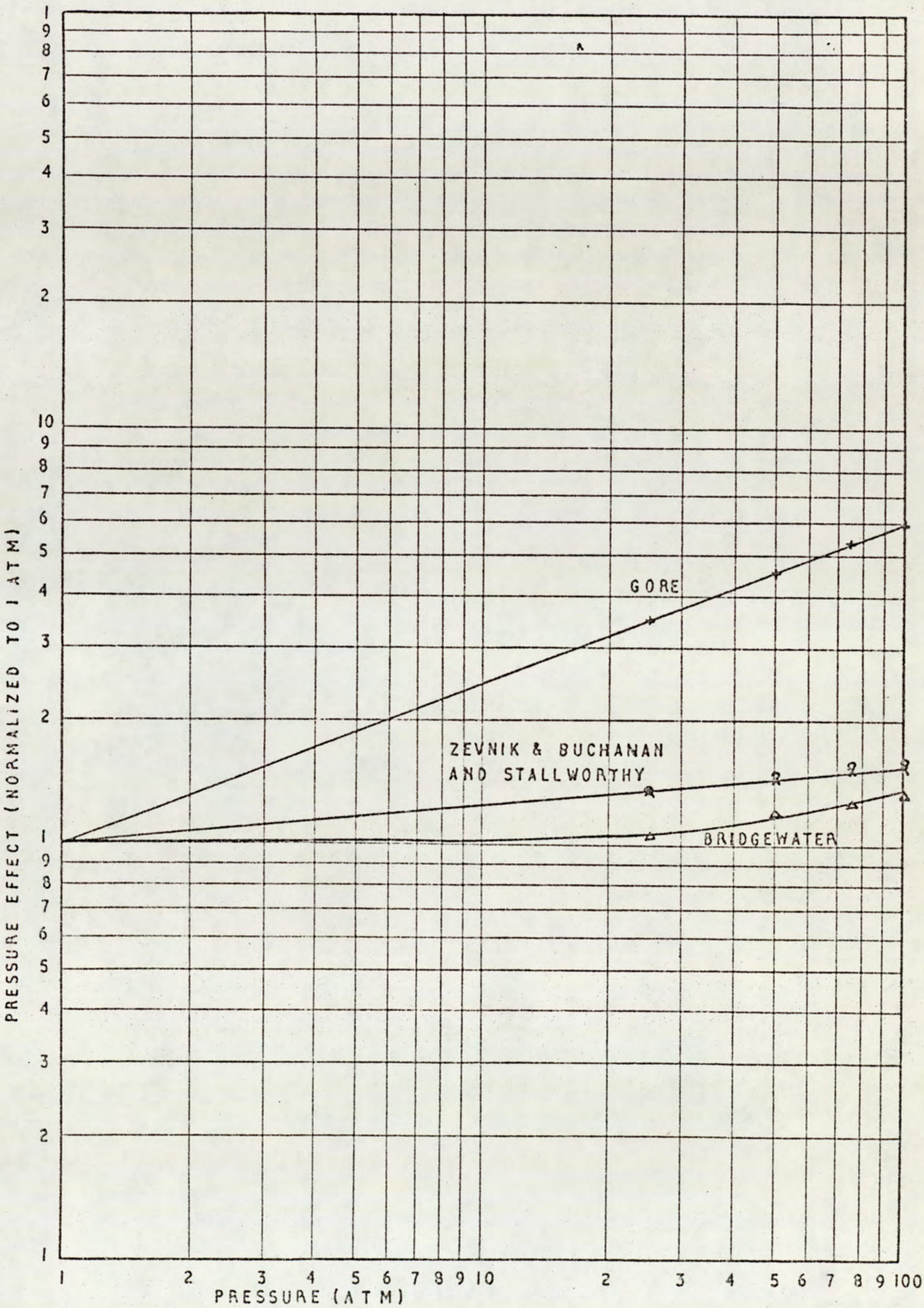


FIG 2

COMPARISON OF PRESSURE EFFECTS[59]

This concludes the cost-pressure relationship discussion. It can be seen that the relationship is so complex that it is not possible to conclude a general cost trend, either way, which could be expected as process pressures were increased. However, based on previous workers' results pressure can be a significant influence on cost depending on phase. For gaseous systems Gore reported a significant effect ($P^{0.4}$) whilst for solid phase processes Bridgwater concluded that it was less so ($P^{0.14}$). Hence, the cost-pressure relationship should be investigated, particularly for gaseous phase processes.

3.4.5.4 Representation of Pressure Effects on Capital Cost

The same problems exist for modelling pressure as noted for temperature and the same conclusions apply. That is, no realistic way of representing and measuring process pressures and their effect on capital cost exists. As with temperature, this was not considered to be a significant problem because:-

a) Pressure influence on materials of construction has been accounted for.

b) Previous workers results indicate that pressure is not a significant influence on installed cost, except for possibly gaseous processes where pressure effects size. However, this effect is compensated for by using the phase sub-process approach.

c) Abnormal pressure conditions are accounted for by the fact that the equipment required to produce them has been classed as a functional unit.

3.4.5.5 Conclusions

The study of the relationships between capital cost and process operating conditions and the subsequent work on how those relationships could be defined and represented in a rapid cost estimating model has revealed that it would be impractical to attempt to produce sophisticated models of capital cost as a function of temperature and pressure. For reasons previously described this is not considered a significant problem since many of the effects they exert on cost are being accounted for indirectly by phase sub-process approach and other process variables used to define capital cost. Also, previous work indicates that their influence on installed capital cost is not considerable and intuition tells one that temperature and pressure influence on cost is restricted largely to equipment cost and not installed cost.

However, it is not possible to conclude that <u>all</u> the temperature and pressure effects are accounted for by the above. For example, temperature influence on insulation is not covered. Therefore, some investigation of their effects is still necessary and so cost as a function of operating conditions will be examined. It is believed that no expected accuracy improvement in the models will result from this and no serious effort is considered necessary as this would only detract efforts from more worthwhile areas of study previously identified. Hence, only preliminary models of cost as a function of the maximum process temperatures and pressures will be obtained and the results studied to assess if further work is warranted.

CHAPTER 4 - EXPERIMENTAL WORK

Having considered and derived the relevant theory to produce a series of rapid cost estimating models the next step is to devise an experimental program that will enable the theory to be tested.

The experimental program consisted of two parts:

4.1 Cost Data Collection, Screening and Adjustment

One of the major criticisms levelled at previous workers in this field is that very often inadequate cost data was used to derive their methods. To ensure that reliable and accurate cost estimating models are derived it is considered essential that a sound cost data base is used in the experimental work.

The first requirement is for the collection of a large quantity of cost data which can be screened and analyzed to assess its reliability and also to identify the limitations and inherent errors in published cost data which exist to cause possible variances between different sources. Thus, cost data, after selection for reliability, has to be normalised to a common base to make it consistent and comparable. The procedures used to satisfy these requirements are discussed.

4.1.1 Data Collection and Screening

So that a large data set could be obtained a complete survey of the chemical engineering literature was undertaken to find reasonable quality cost data. The following journals were used in the search - Hydrocarbon Processing, Chemical Engineering, Process Engineering, Chemical Engineering Progress, European Chemical News, British Chemical Engineering, The Chemical Engineer, American Institute of Chemical Engineers Journal, Chemical Engineering News and Industrial and Engineering Chemistry.

The following constraints were used to select cost data for use in the research and so form the screening process used on the cost data found in the literature.

- Any cost data used must be accompanied by sufficient corresponding process information so that modelling can be reliably performed.
- All other references associated with the cost data reference were checked to ensure that the cost quoted could be accurately placed in time.
- Where costs were quoted with no corresponding time base the time base was set to be at 6 months prior to the source publication date. This was assumed to be a reasonable time lag.
- Only recent cost data between 1966 and 1976 was used in order to minimize the errors associated in updating costs using cost indices.
- Costs were collected only for processes built in the industrialized western world. It was considered that cost data relating to plants constructed in the Iron Curtain countries and

Third World locations could not be realistically compared. This is a conclusion shared by Yen Chen Yen[65] in his study of plant costs in developing countries and previous workers in this field.

Using the above constraints 103 sets of cost and process data were obtained from the literature survey. It was then necessary to meet the second requirement of standardizing it to a consistent and comparable base. This is discussed next.

4.1.2 Cost Data Standardization

In order to achieve the required standardized cost data base the collected cost data was studied and analyzed to identify factors which exist (apart from the process parameters discussed in Chapter 3) to cause cost differences in apparently identical process plants. From the analysis it was concluded that the following variables had to be studied and the extent of their influence quantified to enable cost data to be accurately compared -

Project Cost Definition
Year of Construction
Location
Learning Effects
Market Effects

Each of the above points is discussed in detail so that their influence on cost is understood and recommendations can be made to measure and take account of their influence. As a result of the above study the standard cost base will be defined and procedures for the standardization of the cost data be described.

4.1.2.1 Project Cost Definition

Capital costs published in the literature are quoted in a variety of forms and there are numerous types of investment to be defined and estimated. Examples of these different investment definitions include grass roots investment, fixed capital investment, turnkey costs, revamp/expansion costs and battery limits investment. The terms are fully defined and explained in Appendix 2.

It has been decided however that battery limits investment is to be the standard cost definition used for this research as most of the cost data in the literature is defined as battery limits and so errors involved in relating different cost definitions are minimized. Furthermore, it was apparent from the initial experimental results and cost data analysis that different interpretations of the battery limits cost definition were being used by different sources in the literature and so a further requirement for standardizing battery limits cost data existed also.

Valuable assistance was given to solving this problem or relating different definitions of project cost by Kay[46], who explained the problem of project cost definition and supported it with data showing the approximate breakdown and components of project cost which were achieved at the ICI site at Teeside. The discussion with Kay is summarized as follows -

Total Investment =	Fixed Capital Investment	+
(Grass roots or	Start up Charges	+
Greenfield site)	Working Capital	

where the Fixed Capital Investment was defined as -

the battery limits investment (cost of the production unit) + storage + land + utilities + services + contingencies

At ICI the battery limits investment was defined to include -

DIRECT COSTS	- delivered to site purchase costs and installation costs of major equipment, piping, electrics, civils, steel and insulation
+ INDIRECT COSTS	- engineering, procurement, licensing, construction supervision, spares, administration charges

A breakdown of this cost was given for a 'typical' project as follows -

DIRECT COSTS (Material and Labour)

Item	%
Equipment	35
Piping	25-35
Instrumentation	12-15
Civils	10-12
Insulation	3
Others	12-15

INDIRECT COSTS

Item	%	
Engineering	15-20	(25 for small projects)
Construction Management and contingencies	10-15	
Start Up	5-7½	
Spares	2	
Administrative Charges	1-1½	
	133-146	(this factor is mostly a function or plant size)
average = 140		

Given this information then (whilst recognizing its limitation of being a single source applicable to 'typical' projects) together with other sources in the literature[(64)] (see Table 4.1.2.1.1) it was considered that enough information was available to make approximate assumptions about the relationships between the different cost definitions quoted in the literature and to derive approximate cost ratios for use in normalising

TABLE 4.1.2.1.1 BATTERY LIMITS PERCENTAGE FACTORS (64)

	FLUIDS						SOLID/FLUID						SOLIDS					
	SMALL Instal-lation			LARGE Instal-lation			SMALL Instal-lation			LARGE Instal-lation			SMALL Instal-lation			LARGE Instal-lation		
DIRECT PROCESS COSTS	L	A	H	L	A	H	L	A	H	L	A	H	L	A	H	L	A	H
Delivered Equipment Cost	18	25	36	19	27	39	27	36	44	28	36	46	27	41	55	27	41	54
Installation of Equipment*	7	11	14	8	12	16	11	13	17	11	14	18	11	16	22	11	16	22
Process Piping (installed)	10	18	24	10	21	32	12	13	20	13	17	22	3	5	11	3	7	12
Instruments (installed)	2	5	10	3	5	10	1	3	7	3	4	8	2	3	7	1	4	8
Electrical (installed)	1	2	4	1	3	4	1	3	5	1	3	6	2	3	7	1	3	7
Process Buildings	1	2	18	1	3	19	1	3	11	1	3	14	1	2	7	1	3	8
Average Total Process Costs		63			71			71			77			70			74	
INDIRECT COSTS																		
Engineering, etc.	14	16	19	10	13	16	11	13	16	8	11	13	11	14	16	11	12	14
Contingencies	18	21	27	15	16	23	12	16	19	10	12	17	14	16	19	12	14	18
Fixed Capital Investment		100			100			100			100			100			100	

* Includes painting and insulation. Small/Large Division $\$10 \times 10^6$

costs to a standard battery limits level.

The following assumptions were used to adjust cost data definitions to the base case, defined as the battery limits investment (see Appendix 2 for full definition) and considered to be equivalent to the ICI battery limits base of 140.

1. Where costs were quoted simply as 'battery limits' a cost base of 120 was assumed. This is an average of the 'maximum' cost ICI definition (140) and the 'minimum' cost contractor definitions (100). Considering the lack of definition given in the literature this cost level seemed to be a reasonable one to take.

2. Battery limits costs quoted by contractor sources were generally found to be lower relative to other sources by early experimental results, a conclusion shared by Kay, and unless detailed definition data existed a cost base of 100 (i.e. direct costs only) was assumed for data from these sources.

3. Where details of battery limits cost were quoted, for example an investment was quoted to include direct costs and engineering, the investment figure was adjusted to the standard 140 using assumptions based on ICI data of battery limits cost breakdown.

The above assumptions were the result of the discussion with Kay which was concerned with standardizing battery limits cost data. However, other definitions exist (see Appendix 2) which have to be related to the standard battery limits base (of 140) and the following assumptions were used to do this -

4. Costs quoted as fixed capital investment (i.e. onsites + offsites investment) were divided by 1.40. The literature [(9),(10),(64)] indicates an average value of 1.3 (see Table 4.1.2.1.2). However, Kay stated that the ratio of fixed capital investment to battery limits investment may vary from 1.2 to 1.5 depending on the type and size of process being considered. Since the data being used referred mostly to mostly large scale petrochemical process it was thought that a relatively high ratio would be applicable[(64)] and hence the 1.4 value was decided upon.

5. Total/Grass Roots investments (onsites + offsites + working capital) were divided by 1.55. This assumed that the working capital was 10% of the fixed capital investment (a reasonable value for large scale processes).

6. Turnkey battery limits investment was considered equivalent to the 140 base.

7. Plant expansion investments on existing sites were considered equivalent to a base of 150 (to account for the added complexities in engineering and installation.

This concludes the work on relating different cost definitions. Although it is realised that the work is in many cases based on assumption, and thus can only

TABLE 4.1.2.1.2

GRASS ROOTS PERCENTAGE FACTORS (64)	FLUIDS						SOLID/FLUID						SOLIDS					
	SMALL Instal-lation			LARGE Instal-lation			SMALL Instal-lation			LARGE Instal-lation			SMALL Instal-lation			LARGE Instal-lation		
DIRECT PROCESS COSTS	L	A	H	L	A	H	L	A	H	L	A	H	L	A	H	L	A	H
Delivered Equipment Cost	15	21	30	15	21	30	20	27	33	20	26	33	20	30	40	20	30	40
Installation of Equipment*	6	9	12	6	9	12	8	10	13	8	10	13	8	12	16	8	12	16
Process Piping (installed)	8	15	20	8	16	25	9	10	15	8	12	16	2	4	8	2	5	9
Instruments (installed)	2	4	8	2	4	8	1	2	5	2	3	6	1	2	5	1	3	6
Electrical (installed)	1	2	3	1	2	3	1	2	4	1	2	4	1	2	5	1	2	5
Process Buildings	1	2	15	1	2	15	1	2	8	1	2	10	½	1	5	½	2	6
Average Total Process Costs		53			54			53			55			51			54	
OTHER DIRECT COSTS.																		
Utilities (installed) **	5	8	12	3	12	20	6	9	15	3	13	20	3	5	10	3	7	12
General Services ***	1	2	3	1	3	6	1	3	3	½	3	6	4	7	10	3	5	8
Buildings, General ****	2	3	10	2	3	14	4	7	12	3	6	10	8	9	12	6	8	12
Storage, receiving and shipping (Installed)	2	4	10	2	5	12	3	6	10	2	6	12	2	6	10	3	6	10
Average Total Direct Costs		70			77			78			83			78			80	
INDIRECT COSTS																		
Engineering, Overheads, etc.	12	13	16	8	10	12	8	10	12	6	8	9	8	10	12	8	9	10
Contingencies	15	17	22	12	13	18	9	12	14	7	9	12	10	12	14	9	11	13
Fixed Capital Investment		100			100			100			100			100			100	

give approximate values, it is considered that the ratios listed above will greatly assist in achieving a standardized cost data base.

4.1.2.2 Year of Construction

The effect of time on costs is well appreciated and is measured by a cost index. A cost index (C_I) is merely a number for a given point in time showing costs at that time for a specific geographical location relative to a defined base year. If the costs at some time in the past for the location are known the equivalent cost at the present time at that same location can be determined by multiplying the original cost by the ratio of the present index value to the index value at the time the original cost was obtained, as follows -

$$\text{COST @ present day} = \text{COST @ year of construction} \times \frac{C_I \text{ present day}}{C_I \text{ year of construction}}$$

Cost indices are compiled such that components of the overall plant cost are weighted according to their supposed significance (derived statisticaly), the total of the weightings being unity. As the component costs rise the affect is accounted for in the construction cost index and costs may be updated.

Many cost indices are available for use in the chemical engineering literature, the subject having received much attention over recent years. Unfortunately, all cost indexes are artificial and two indexes covering the same type of projects may give results that differ considerably[(66),(67),(68)] depending on their derivation. The most any index can do is to reflect average or hypothetical changes. In times and/or locations of economic stability this effect is less marked and cost indices tended to rise in similar proportions such that it is of no great importance which index is used. However, over recent years a period of considerable economic instability has occurred with widely differing inflation rates and corresponding exchange rate fluctuations throughout the world. As a result large discrepencies between different cost indices have been noted.

Because it is necessary in this research to be able to accurately relate plant costs at different times to each other and standardize them to a common time base and location the choice of cost indexes is important. All indices have disadvantages and advantages when compared to others since they all are designed to suit different applications. Some of the more widely quoted cost indices are reviewed below.

Marshall and Stevens Index	- USA
Engineering News Record Index	- USA
US Department of Labour Index	- USA
Chemical Engineering Index	- USA
Nelson Refinery Index	- USA
Process Engineering Index	- UK
Engineering and Process Economics Index (EPE)	- UK,USA, and others

(The problem of relating costs at different locations will be covered in the next section).

1. Marshall and Stevens Index(es)[69]

There are two categories of this index -

(a) the all industry equipment index is the average of the individual indexes for 47 different types of industrial, commercial and housing equipment.

(b) the process industry equipment index is a weighted average of eight of these with the waiting based on the total product value of the various process industries.

The percentages used for the weighting are as follows -

Industry Equipment	%
Cement	2
Chemicals	48
Clay Products	2
Glass	3
Paint	5
Paper	10
Petroleum	22
Rubber	8

These indexes consider the cost of machinery and major equipment, fixtures, office furniture and other minor equipment.

The Marshall and Stevens indexes have two disadvantages which discount it for use in this research.

They are concerned with equipment costs only not installed costs and they are based on an index value of 100 in 1926, a time base so distant that it seems inevitable that changes in equipment specification must affect the credibility of the index which cannot adequately cope with such changes. It is generally considered to be a good index however when applied correctly although it may tend to give low figures.

2. Engineering News Record Index[69]

This index was primarily designed to assess the cost changes in civil engineering works that use large quantities of unskilled labour and as such is heavily dominated by labour costs.

It employs a composite for structural steel, lumber, cement and common labour and is usually reported on one of three bases -

(a) An index value of 100 in 1913
(b) An index value of 100 in 1926
(c) An index value of 100 in 1949

It was never intended for use in the chemical industry and is not suitable for use in this research.

3. US Department of Labour Indexes[9,10]

The US Department of Labour publishes monthly statistics giving material and labour indexes for various industries. The labour index is reported as the average earnings in cents/hr. for workers involved in the manufacture of durable goods compared to a 1926 time base. The material index applies to the metals and metal product industries and is also compared to a 1926 base level. By obtaining a total labour and material index, the values can be used to estimate the present cost of purchased equipment from the cost of similar plant at a past date. For process equipment made of conventional materials of construction a material/labour split of 50/50 may be assumed. If special materials are used the material/labour split may be 65/35.

Again, this index is not considered suitable for this research. Apart from the distant time base, estimation of material/labour ratios for process plant have to be made which can lead to sizeable errors. Also, the index is suitable only for predicting equipment cost changes.

4. Chemical Engineering Index[69]

This index was an attempt to make a tailor made index for the chemical plant construction industry in the U.S. It was designed to be a highly flexible index and included many sub-indices which made it capable of being modified for all purposes. The weighted composite was thoroughly researched, well documented and easy to understand and is as follows -

Component	%
Equipment	22.6
Machinery	12.8
Piping	12.2
Steel	6.1
Electrics	3
Instrumentation	4.3
Erection Wages	22
Buildings	7
Indirect Costs	10
	100

The index has an inbuilt productivity adjustment of 2.5%/annum, but this could be over-correcting.

The Chemical Engineering Index is probably one of the best construction cost indices applicable to the chemical process industries today and will be considered as a possible candidate for use in this research.

5. Nelson Index[69]

The Nelson Index was designed specifically to measure cost changes with time for the petroleum industry and refineries in particular. It has no labour productivity correction factor built into it as such. In 1966, the True Refinery Index was introduced, which is supposed to be corrected for productivity and technological/learning improvements. It is aimed specifically at refinery construction and due to the very high costs incurred therefore it would be risky to use it for any other application. As far as chemical plants are concerned the original Nelson Index is a rough guide, but tends to give high figures. It will not be considered for use in this research.

6. <u>Process Engineering Index</u>[(70),(71)]

Originally the Chemical and Process Engineering Index, the name was changed as the magazine publishing it changed its name to Process Engineering. The index was introduced in 1965 and was designed to measure cost changes in the UK[(70)].

The original index incorporated productivity allowances and employed a weighted composite based on mechanical, civil and electrical components of process plant construction. The index was re-examined in 1973[(71)] with the aim of increasing its sophistication and thus prediction accuracy. Cran, who was responsible for the development of the index from the start, proposed a new index based on the following weightings -

$$I = 0.37IM + 0.08IE + 0.1IC + 0.19IS + 0.26IO$$

where I = process engineering index

IM = mechanical component of project cost

IE = electrical component of project cost

IC = civil component of project cost

IS = site engineering component of project cost (project labour)

IO = overheads component of project cost (engineering, design, supervision etc)

The component indices are in turn made up of a composite of industrial indices.

Cran warns that the weightings of the components used in the index are for a 'typical plant' and are not necessarily independent of plant type and location. He suggests that a user may wish to input his own factors based on his own judgment. For example, a refinery would have a relatively large mechanical component. However, he states that he is fairly sure of the coefficients in his weighted index being representative for the majority of plants and quotes a study carried out in the U.S.[(86),(87)] which analyzed a wide range of plant costs and concluded that the relative weightings of component cost were independent of the type of plant and independent even of the country in which a project is launched.

The index is probably the best indicator of plant costs relative to time in the UK in use today and will be considered for use in the research. Its main advantage is that the data required for update is readily available from regularly published government sources.

7. Engineering and Process Economics Index(es)[(72)]

In 1976, Cran established a new series of cost indices for use in the chemical process industries in 16 different industrialized western countries. Setting a base of 1970 = 100 for each country a cost index was produced for each location showing the change in plant costs for that location through a period from 1970 to 1976. The escalation for each country was a measure of plant cost increases measured in the local currency. The indexes for the different locations are shown in Table (4.1.2.2.1).

The EPE index was not sophisticated as lack of data for the different locations meant that a complex multi-component cost index could not be derived. A two component index based on steel and labour cost data was derived:

$$I = 0.4\ I_S + 0.6\ I_L$$

where I = EPE cost index for a particular location

I_S = steel price index for the particular location

I_L = labour price index for the particular location

The index was derived as such from an analysis of previous multi-component indices which were simplified to the above by the theory of elimination. The resulting index provides a fairly representative index for the chemical construction industry and can be easily deduced from data that is readily available from published sources (e.g. the United Nations Monthly Bulletin of Statistics).

When tested the two component index was found to compare very well with the established published indexes for the different locations with only slight differences in evidence. This is possibly the best justification for his approach.

It was decided that the EPE construction cost index should be adopted for use in this research. Whilst it is recognized that the EPE index for a

TABLE 4.1.2.2.1 EPE COST INDICES

EPE plant cost indices (1970 = 100). Annual indices

	Australia	Austria	Belgium	Canada	Denmark	France	West Germany	Ireland	Nether-lands	South Africa	Spain	Sweden	U.K.	U.S.A.
1971 J	106	107	105	107	109	109	110	113	108	112	112	99	113	104
1972 J	116	114	117	115	116	120	118	126	129	115	124	103	126	107
1973 J	125	126	133	123	137	141	132	145	138	128	142	105	132	113
1974 J	143	156	162	135	170	168	144	176	154	147	175	108	148	119
1975 J	186	199	188	165	204	197	176	240	173	188	217	154	205	149
1976 J	214	195	201	189	213	219	174	280	191	219	280	171	245	149
1977 J	245	221	236	208	232	256	186	319	202	263	345	202	285	163
1978 J	272	247	259	244	251	275	191	356	210	272	411	232	295	175

EPE plant cost indices (1970 = 100). Monthly indices

	Australia	Austria	Belgium	Canada	Denmark	France	West Germany	Ireland	Nether-lands	South Africa	Spain	Sweden	U.K.	U.S.A.
1975 J	186	199	188	165	204	197	176	240	173	188	217	154	205	149
F	187	187	190	168	207	196	174	245	173	188	219	155	205	149
M	191	193	179	170	210	195	173	249	173	190	220	155	213	150
A	192	198	182	172	212	200	171	254	174	191	232	156	214	145
M	198	222	184	173	209	203	171	258	175	193	244	157	224	142
J	198	236	190	175	206	206	172	263	175	214	254	159	227	143
J	199	230	192	175	203	207	172	267	182	215	267	161	229	144
A	202	216	194	176	205	209	172	271	182	216	279	164	230	144
S	208	205	195	179	207	210	171	274	182	216	276	158	231	146
O	211	246	195	184	209	212	171	275	184	216	273	162	232	146
N	212	240	197	187	209	213	172	276	185	217	270	164	242	146
D	212	213	199	185	211	215	173	277	178	217	274	170	243	148
1976 J	214	195	201	189	213	219	174	280	191	219	280	171	245	149
F	225	197	201	189	214	221	177	283	192	220	285	174	251	149
M	227	199	220	191	216	223	180	287	193	221	291	176	251	150
A	229	202	220	193	217	230	185	290	195	221	296	179	261	150
M	234	204	222	196	219	231	185	293	196	232	301	181	273	151
J	234	206	224	199	221	233	186	296	197	233	307	184	274	151
J	237	208	225	198	222	242	191	299	198	246	312	187	276	152
A	238	210	227	199	224	244	191	303	198	249	317	189	276	158
S	238	213	229	201	225	246	191	306	199	252	323	192	283	158
O	240	215	231	203	227	253	192	309	200	252	328	194	284	158
N	245	217	233	204	229	255	190	312	200	254	334	197	284	161
D	245	219	234	207	230	256	187	315	201	262	339	200	284	163
1977 J	245	221	236	208	232	256	186	319	202	263	345	202	285	163
F	246	223	238	210	233	255	187	322	203	264	350	204	284	163
M	253	226	240	210	235	255	188	325	203	265	356	207	286	164
A	254	228	242	210	237	257	189	328	204	266	361	210	285	165
M	257	230	244	211	238	261	190	331	205	264	367	212	287	166
J	258	232	245	214	240	263	191	335	205	264	372	215	288	169
J	260	234	249	217	241	264	192	338	206	264	378	218	288	169
A	265	237	251	217	243	266	190	341	206	264	383	220	288	170
S	265	239	253	220	245	268	190	344	207	270	389	222	289	171
O	269	241	254	220	246	270	191	347	208	272	395	225	289	173
N	271	243	256	222	248	271	191	350	209	272	400	227	298	173
D	274	245	258	223	250	273	191	353	210	271	406	229	294	174
1978 J	277	247	259	224	251	275	191	356	210	272	411	232	295	175
F	279	249	261	226	253	277	191	359	211	273	417	234	296	177
M	282	251	263	227	254	278	192	362	212	274	422	237	297	179
A	285	253	265	229	256	280	192	366	213	275	428	239	298	180
M	288	255	267	230	258	282	192	369	214	276	431	241	299	181
J	290	257	268	232	259	284	192	372	214	276	434	243	300	182
J	293	259	270	233	261	286	192	374	215	277	437	246	301	184
A	296	261	271	235	262	288	193	377	216	278	440	248	302	185

particular country may not necessarily be as accurate as some of the more well established indices for that country there are some major advantages.

The index has been recently developed having a time base of 1970 = 100. This is an advantage because the inherent defects in any index which cause it to be in error as time proceeds from the time of its development are minimized. Since the cost data to be used to derive cost estimating models is of the same time period as the data used to derive the index the EPE index would seem to be the most compatible index to use in this research as learning effects are minimised.

It is necessary in this research to gather a large amount of cost data which inevitably means that data from many countries must be used. The series of EPE indexes produced by Cran for the different countries provide a consistent measure of cost changes with time for the different countries listed. This is an important point, particularly to enable use of data from countries with no other cost index available.

The EPE index then can relate costs through time using the individual indices for the different locations but knowing that each index has been derived in the same manner and thus has common and consistent errors and faults. This consistency allows comparison and relates cost data from different times and locations safely.

Cran also comes to the same conclusion and notes that since the indices are comparable they can furnish some of the information needed to produce a location index and determine location effects on plant costs. This subject is discussed in the following section.

Other European cost indices exist. For example, in the UK the Association of Cost Engineers publishes an index and in Germany the Kolbel and Schulz index is widely used. However, these are generally poorly documented and the literature available on them is sparse.

4.1.2.3 Location Effects

As mentioned in the discussion on cost indices, a cost index only provides a measure of cost changes with time for one particular location or country. Since cost data is being collected for different locations in this research, and because many factors exist to make cost levels in these locations different at any time, it is necessary to attempt to standardize the cost data to take account of these location differences.

Unfortunately, because there are many factors which contribute to location effects on project costs some of which are not always well defined - such as shortage of skilled labour, industrial disputes, adverse weather and climatic conditions - and because these phenomena are not readily predictable, a reliable analytical approach to the problem of location effects is virtually impossible. Therefore, reliance must be placed upon historical data if location effects are to be quantified. Two alternatives exist for quantifying location effects and thus establishing location indices for use on the project cost data, namely:

1. Derivation of factors from historical cost data
2. Making use of published location factors

The first alternative was considered by Cran in 1973[73],[74] who by an analysis of US and UK cost data derived a location index for the two countries US/UK of 1.2-1.3 for 1973. This factor was derived by comparing standardized average unit costs of similar process plants for the two locations, which were calculated as mean values from the available data. To eliminate or reduce the effects of random error it was necessary to collect sufficient data from which mean values could be calculated. It was also considered that by assuming average values the effects of learning and other influencing factors could be minimized. In conclusion Cran states that due to problems in standardizing his data for project cost definition, learning and plant size (which required accurate size exponents) his results should be treated with caution. He also noted that the location index varies as a function of the process being considered.

From this work it was necessary to consider the feasibility of deriving location factors from the project research data using his technique. That is, by comparing similar processes and standardizing costs at locations to the same time base and plant size using size exponents, could location indices be established? However, the same limitations which applied to Cran's results would apply, insufficient good quality data was available on which to base the study, and the introduction of a further unquantifiable factor might invalidate the research.

Having rejected the first alternative then the second was examined. Apart from Cran's work on US/UK location comparisons only three other references could be found which covered the problem of location indices. Mendel[75] published a paper in which he described how plant costs could vary at different locations within one country, the US, mainly as a function of wage rates and labour productivity. Johnson[76] expanded the subject and produced a list of location indices applicable to most of the major western industrialized countries in 1969 (see Table 4.1.2.3.1). Considering the importance of the subject, however it was disappointing to find that no other comparable data existed to check Johnson's indices.

It is considered that the use of Johnson's factors is not warranted or advisable for this research. Apart from it being single source data, and hence not verifiable (except for the US/UK index of 1.1 compared to Cran's value of 1.2 and Bridgwater's[77] 1.03-1.1) other complications exist which need to be considered before using any published location index. Because the major factors which influence location indices

TABLE 4.1.2.3.1.

JOHNSON LOCATION INDICES (76) (1969 base)

Country	Index
USA/Spain	1.0
UK	0.9
Scandinavia	1.1
Germany	0.85
France	0.9
Japan	0.82
Italy	0.84
India, etc	1.35

are constantly changing relative to each other with time - productivity, wage rates, inflation, exchange rates etc. - then it is to be expected that location indices will also change with time and thus are applicable for short terms only. This is to be more expected since 1972, since when inflation rates throughout the world have differed significantly and currency exchange rates have been allowed to float and with wide fluctuations.

The only recent studies of location indices have been made by Bridgwater[(77),(78)]. In 1976, he published a study showing a comparison of US/UK cost levels from 1970 onwards and drew some interesting conclusions. From an analysis of inflation rates predicted by the US and UK, EPE cost indices and exchange rate fluctuations between the location currencies he noted that prior to 1972, in a period of fixed exchange rates, a variation of the location index existed probably due to the artificial monetary exchange system. After 1972, a reasonably constant value is obtained which shows UK plant costs remaining at a level of between 90-97% of the US costs, which he indicated was a result of the two exchange rates being allowed to achieve realistic comparative levels and ultimately acting towards minimising variations in location effects.

At the time of experimentation in this research the author was of the same view and considered that no location effects existed, or if they did they were negligible. Although no direct evidence can be offered to support this view an analysis of the research cost data, standardized to June 1976 time base and converted to US $ using the current exchange rate, showed no indication of unit costs for each location (expressed as cost per functional unit) exhibiting any measurable location effects whatsoever, and unit costs from all locations showed up as either positive or negative relative to the US location unit costs taken as the base. Hence, it was thought that no location indices need be applied to the cost data. Costs at location were simply converted to the US equivalent using the June 1976 exchange rates.

4.1.2.4 Learning Effects

Any operation can be carried out better the next time. It is now accepted that the pattern of improvement can be sufficiently regular to be predictable. Such a pattern characterizes not only individual performance but also the complete performance of many individuals organized to accomplish a common task.

In 1964, Hirschmann[79] published a paper in which he maintained that industrial learning curves could quantify such performance. In a study of the aircraft industry it was found that the number of manhours needed to manufacture an aircraft declined at a regular rate over a wide range of production time. Such continuing improvement is so common in the aircraft industry that it is now the normal expectation.

Hence, production and other measurements of performance are customarily scheduled on some basis of progressive betterment.

Although learning curves have been found in other industries, they have yet to realise equivalent acceptance. Instead costs are expected to remain at a relatively constant level (after inflation has been taken into account). Nevertheless, progress through experience and technological improvements constantly occurs, and since this progress reflects increases in efficiency, the paths traced by the progressive improvement may be claimed as learning curves.

In his paper Hirschman shows a learning curve which was evolved in the aircraft industry from the plotting of unit assembly labour against the cumulative number of units manufactured (Fig. 4.1.2.4.1), the improvements being expressed in terms of the ratio of unit manhours everytime production is doubled.
He states that learning curves will have different ratios depending on:-

(a) the operation involved

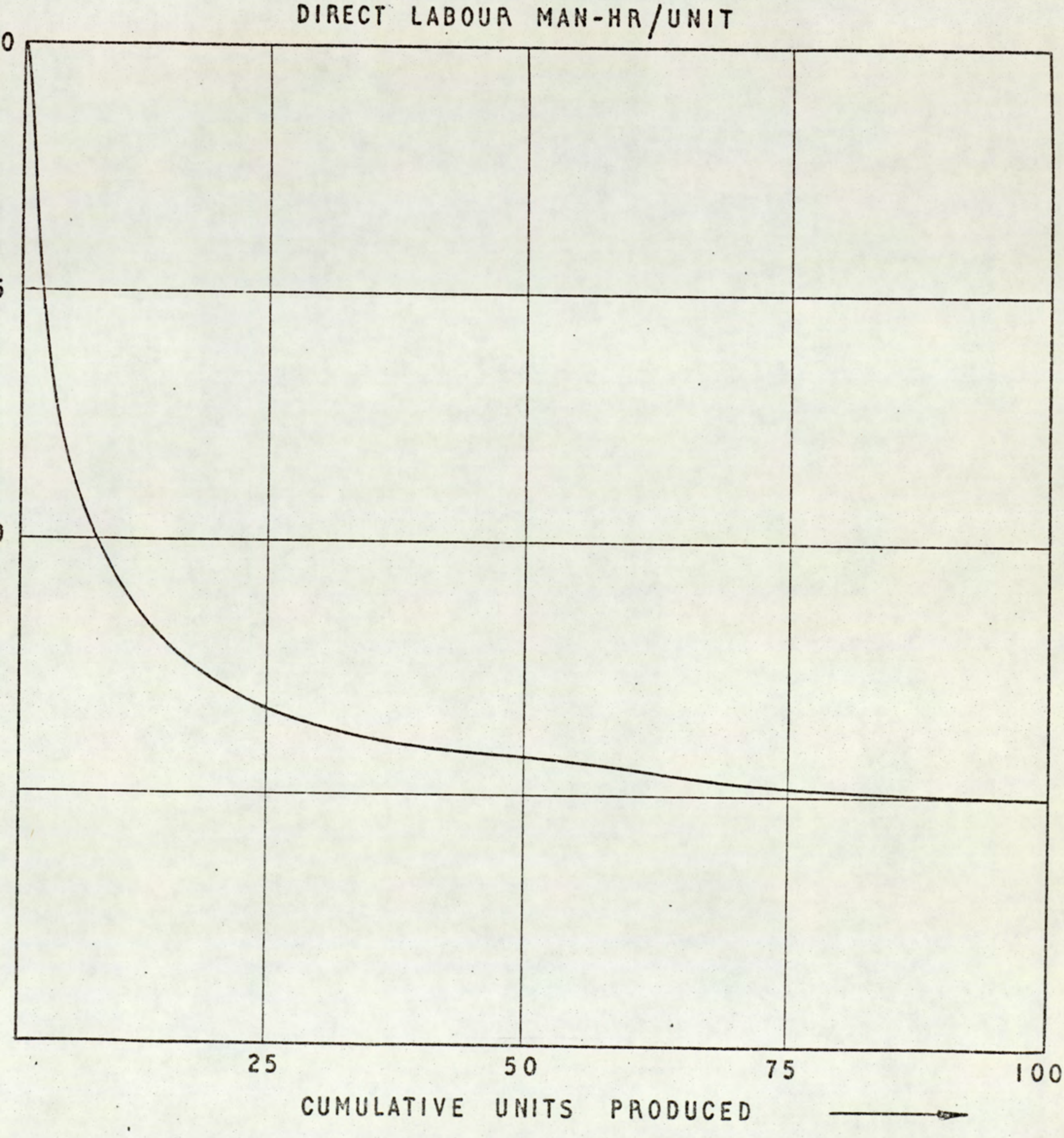
(b) the proportion of direct labour for assembly to that for machine work

(c) the repetitive nature of the operation

When plotted on log-log paper, the learning curve is a straight line that is represented by an equation of the type

FIGURE 4.1.2.4.1

AN 80% LEARNING CURVE SHOWING A RAPID INITIAL DECLINE[79]



$$\frac{C_2}{C_1} = \left(\frac{X_2}{X_1}\right)^{-b}$$

where

C_2/C_1 = the ratio of unit labour or unit cost

X_2/X_1 = the ratio of production

b = the exponential relationship between the two ratios

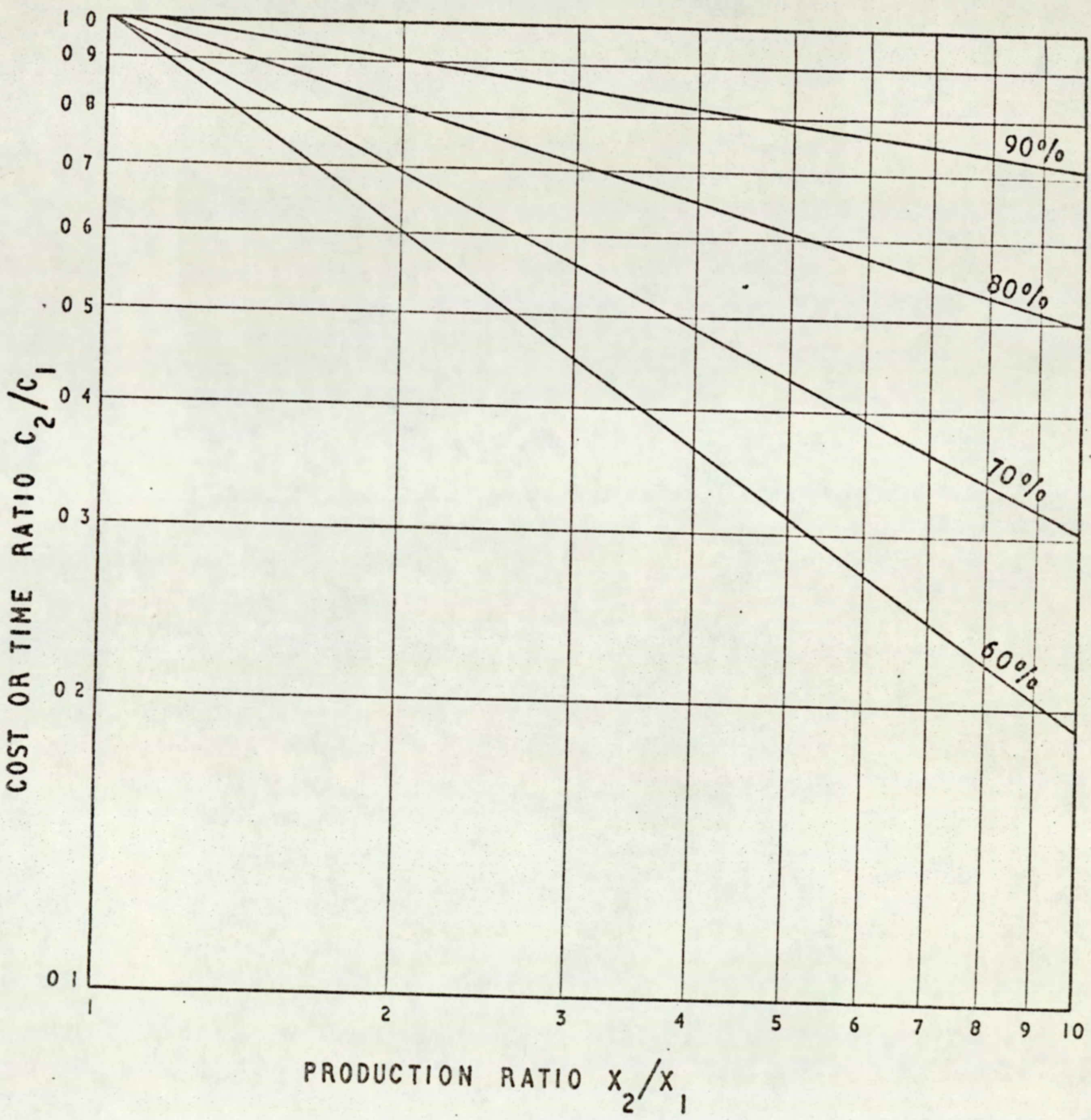
Fig (4.1.2.4.2), published in Hirschmann's paper, shows learning curves for the aircraft industry based on a decrease in unit costs of 90%, 80%, 70% and 60% between doubled quantities. The commonly expected slope of the learning curve in the air frame industry is 80%.

As a follow on from Hirschmann's initial analysis of the learning effect, Nelson[80],[81],[82] started to publicise the influence of learning on the cost of process plants making comparisons between the costs one would expect from the normal cost index indicators and the actual increases observed. He concluded that a slope of 75% could be expected in the petrochemical and chemical industry over four years on plant already in commercial production (1962-1966).

Hirschmann also looked into the learning effect on chemical plant[83] and concluded that the capital cost of similar plants decreased (relatively) with the number of plants built. He indicated learning slopes of between 70% and 80% for the industry, which were compatible with Nelson's findings.

FIGURE 4.1.2.4.2

TIME RATIO v PRODUCTION RATIO.[79]



Therefore it can be seen that the learning effect is significant enough to be considered when analyzing historical cost data and that it would be prudent to include it in any potential estimating methods. It can help prevent erroneous figures and increase the accuracy of cost estimates.

Having decided this, however, it is necessary to consider how such learning effects, or factors, can be sufficiently quantified for this research. An initial analysis of the problem reveals that serious constraints exist to make the successful calculation of learning effects improbable. A great deal of information and research would be required before any such learning factors could be accurately establisehd as learning effects would vary for each individual process and as a function of the time span they were measured in. Hence, to establish learning factors even for one single process it is necessary to know plant costs, sizes and the number built, where and when the process was introduced, and have sufficient knowledge to be able to standardize this data to a common time, location and plant size base accurately. Only with this information could a comparison be obtained of actual costs of a plant against its expected value based on inflation; hence deriving a learning effect either as a function of time or the number of plants built. Hence, it can be seen that to establish learning effects for a sufficiently broad range of processes to conclude a useable general factor for inclusion in the estimating methods would be a mammoth task. Apart from the time involved in undertaking such a study, however, the required information needed to draw valid conclusions is not available in the literature. This prevents such a study from taking place and could only be effectively carried out with substantial industrial collaboration.

4.1.2.5 Market Effects

It is also known that plant costs can be greatly influenced by the overall business climate at the time of construction[84]. Depending on the supply and demand situation present in the fabrication and contracting industries at the time money is to be invested, plant costs will be either high or low relative to some norm depending on the level of competition for work which exists. Hence, it is advantageous for companies to plan their investments to meet an optimal supply and demand situation and make use of the expected market factor.

Although this factor has been identified by cost engineers its effect has proved difficult to assess due to the fragmentation and uneven performances of the fabrication and contracting industries, the confidentiality surrounding process plant purchases, and the consequent lack of data. The only quantitative study published in the literature to date has been by Bluck et al[85] who presented a study based on an analysis of Davy Powergas data for 1971 to 1974. An attempt was made to forecast a market factor using a relevant economic indicator and evolving a time series analysis technique to predict market factors (short term) based on previous data.

The latest available values of net new orders for the fabrication industry were obtained and from this data, using the time series transformation techniques, estimates of future orders on hand were made and thus future shop capacity deduced to predict probable market loadings. The paper is noteworthy since it is an original attempt by the authors to tackle a significant and complex problem relevant to the area of cost estimating. However, criticisms have been raised to the methods validity by Butler[84], who rightly commented that the success or failure of the method depended solely on the ability to accurately relate orders in hand to net new orders. Fluctuation of new orders with time are considerably more marked than those of orders in hand because of the smoothing out process which goes on as new orders in hand become translated into work in progress, and there must be some doubt as to whether a time series analysis can justifiably be used to relate one indicator to the other. Butler also casts doubts on whether or not the orders in hand indicator is an accurate representation of workload in vendor shops and suggests a possible alternative data base of percentage capacity utilization, an indication of which is published monthly in the Financial Times for the mechanical engineering industry.

These criticisms apart, however, the Bluck paper still only tackles one part of the market factor problem, the effect of the situation in the contracting industries not being considered. Hence, the problem of forecasting market factors for use in cost estimates is still in its infancy and much work is needed to be done before significant advances are made. No attempt will be made to establish market factors in this research for use in estimating methods and cost data analysis and standardization. Again, the lack of sufficient and adequate data is the cause.

4.1.3 Cost Data Standardization Procedure

Based on the study of cost standardization the following procedure was used to bring each cost data point to the standard base which, for the purposes of this thesis, will be defined as -

> the battery limits capital cost of a process plant (US \$) at June 1976 (time base), in the US Gulf Coast area (location base), with a service factor of 0.9 (e.g. 330 stream days/annum)

This base was defined primarily to reduce errors involved in standardizing the cost data. Since the majority of cost data collected referred to US locations the cost data standardization required for the project could be kept to a minimum and the associated errors similarly minimised. Also, the US location has the advantage of being a stable cost area. If base location costs vary significantly and erratically then other location costs have to be constantly checked for comparative effects.

1. Cost data Screening - to assess the reliability of the data and any known bias it is likely to have, such as source, e.g. contractor, client, and to accurately define its year of construction, location and other relevant parameters.

   Any cost data that did not contain sufficient information to enable standardization procedures and assumptions to be applied was rejected.

2. Standardization of published cost to the base definition equivalent of battery limits (140) using the assumptions described in Section 4.1.2.1.

3. Updating costs to the June 1976 time base using the EPE cost index applicable to the location.

4. Conversion of costs at location to the standard base US location using June 1976 exchange rates.

N.B. An additional experiment using location factors was added at an advanced stage of this work. This is reported later.

4.2 Mathematical Modelling

The research at this stage takes on a statistical aspect whereby the relationship between the dependent variable (cost) and independent variables (defined in Chapter 3) has to be derived. How this is done depends on how the form of the relationship between the variables is considered (either an additive or multiplicative relationship could be postulated) and depending on this the choice of suitable numerical methods capable of solving the formulated models and predicting the relationship between variables.

Previous studies[88] have revealed that multiplicative models seem best suited to representing the relationships involved whereby cost is expressed in the following way:-

$$\text{Cost} = f(x_1 . x_2, x_3 \ldots\ldots x_n)$$

or $$\text{Cost} = k . x_1^{\,a}\ x_2^{\,b}\ x_3^{\,c} \ldots\ldots x_n^{\,n}$$

This approach has become a standard method in the field of rapid capital cost estimating. Plotting such a relationship on algorithmic paper will give a series of straight lines. Such formulae are generally applicable and lend themselves to quick calculation on use. Their great merit is that although they are purely empirical they are very useful as a means of summarizing observations and very frequently quite complex phenomena can be concentrated into a useful model.

The relationship postulated by the multiplicative model, when fitted to a set of experimental data (in this case standardized cost data) is characterized by a prediction equation called a regression equation. The regression equation is produced using a statistical technique of Multiple Linear Regression Analysis. The theory of this technique is freely available in the statistical literature[89],[90] so it will not be discussed here. It is not necessary to understand its derivation, but rather to understand and interpret the statistical results and inferences that the analysis will provide on a postulated relationship.

For the experimental modelling in this research it was decided to use the Open University multiple regression analysis computer program, MULREG. The program was modified to increase its storage capacity (a requirement necessitated by the large cost data base) and to read and print log/integer values where necessary. This latter modification was required since MULREG assumes linearity between the dependent and independent variables. In this project, however, log-linearity exists and hence the program is adapted accordingly by merely taking logs of the postulated model to give a linear equation. The modified program was renamed MULSRT (and is stored and available for future use in the Aston Department of Chemical Engineering group program library).

The program outputs the following statistical information which is used to assess and test the accuracy and reliability of postulated cost models:-

(i) <u>Multiple Correlation Coefficient (R^2)</u> - which measures how much variation in the dependent variable is explained by the postulated model. i.e. a measure of the strength of linearity between variables and the forecasting ability of the model. R^2 may range from 0 to 1. A value of 1 indicates complete linearity, i.e. a complete and error-free fit between dependent and independent variables.

(ii) <u>Standard Error</u> - which represents the error of the regression model in fitting the historical data and is a major determinant in the width of the forecast confidence limits.

(iii) <u>T-Stat Ratio</u> - which measures the significance and reliability of a particular independent variable in predicting the dependent variable. If T-s TAT value > 2 x Standard Error then the variable is 'significant'[(89)]. This output is useful in eliminating insignificant variables from cost models.

(iv) <u>Durbin-Watson Statistic</u> - which measures the degree of linearity between the dependent and independent variables. Depending on the number of independent variables and the number of data points or observations, values are derived which indicate whether a linear relationship can be reasonably assumed to exist[(91)]. Hence, by noting the Durbin-Watson statistic the linearity assumption can be monitored.

Using this information then the formulated models were analyzed and assessed and the results used as feedback in order to produce finalized models capable of meeting the research aims. The results are described and discussed in the following chapter.

5 RESULTS

The experimental work performed in essentially three parts which are discussed in depth in this chapter.

(a) Preliminary model formulation

This consisted of essentially trial and error modelling performed on a small cost data base and served the purpose of highlighting problems associated with cost data analysis and standardisation, together with the expected limitations of the various parameter definitions. The results from this initial modelling phase were analysed and used as feedback so that a sound data base began to emerge.

(b) Final model production

Once the problems above had been identified and the procedures established for achieving a sound data base derived, the cost data base was extended to over 100 carefully considered standardised sets of cost data and the finalised cost models produced. The results of modelling cost as a function of the various process parameters listed in Chapters 3 and 4 are sequenced as follows:

1. Process Size
2. Materials of Construction
3. Process Operating Conditions
4. Functional Units
5. Location Effects

All the models produced are for volumetric phase type processes and as such the theory proposed for modelling cost as a function of phase has not been tested.

(c) Model Testing

Finally, when the research models are available they must be tested for their accuracy and reliability.

5.1 Preliminary Model Formulation

Preliminary modelling was aimed primarily at testing the validity of the proposed functional unit definitions and all of the early models were restricted to determining the relationship between capital costs (C) and the number of functional units (N) and process size expressed as capacity, as a convenient measure (Q). This restriction was applied because it had been concluded from the literature survey that the relationships between these variables were highly significant and had to be reasonably well determined if accurate cost models were to be produced. The principle was the establishment of a building block for later modelling.

Initially 30 data sets (costs escalated to mid-76 in US \$) were employed. The results were very disappointing. None of the early models of cost per functional unit (CPF) v process size (Q) achieved an $R^2 > 0.63$ and had standard errors of around 40%. As such the initial modelling results were unacceptable and far outside the accuracy limits set in the research aims (see section 2.3). The reasons for this were probed and considered as follows:-

(i) Was the functional unit definition proposed inadequate and so a prime cause of error? This assumption was tested by modelling CPF v Q for individual products and thus eliminating as far as possible material of construction and operating condition influences. The results of this approach seemed to disprove that serious problems existed with the definition (although minor discrepancies were noted and rectified) and reasonably good estimating accuracies were obtained for the individual products tested as follows:-

Nitric Acid - 6 data sets, $C = 1.412 \times 10^{-3}$ (N) $(Q_{mw})^{.631}$, $R^2 = .903$, SE = .25

Butadiene - 7 data sets, $C = 1.98 \times 10^{-3}$ (N) $(Q_{mw})^{.574}$, $R^2 = .96$, SE = .22

Cyclohexane - 8 data sets, $C = 6.347 \times 10^{-4}$ (N) $(Q_{mw})^{.67}$ $R^2 = .817$, SE = .31

Apart from seeming to validate the functional unit definition, this experiment also illustrated the need for a large and varied cost data base, since it could be seen that it was a relatively easy matter to achieve good correlations based on a few data points covering a limited number of products and processes.

(ii) Was process size being accurately measured and represented? It was questioned whether or not the capacity measurement was correct and consistent in that it had been decided to measure capacity on a 100% product purity basis (see 3.4.3.2.1), whereas some of the processes modelled (e.g. nitric acid, formaldehyde, and hydrochloric acid) produced output far in excess of these values, due to dilution.

However, experimentation on 13 data sets proved that the 100% capacity basis was far more accurate and significant in determining cost than the total (or diluted)output as can be seen from the following results:-

$$C = 6.25 \times 10^{-4}\ (N)\ (Q_{mw})^{.69}, R^2 = .923,\ SE = .198,$$
$$T_{STAT}Q_{TW} = 5.34$$
$$C = 5.068 \times 10^{-4} (N)\ (Q_{TW})^{.667}\ R^2 = .697,\ SE = .464,$$
$$T_{STAT}Q_{TW} = 5.34$$

(iii) Based on the results of the first two testing experiments it was concluded that the definition and measurement of the N and Q were sufficiently sound to expect reasonable modelling accuracies to be obtained. Having concluded this it was logical to assume then that the prime source of the errors being experienced was being caused by the remaining variable being used in the initial cost modelling experiments; the capital cost variable.

It was this realisation that led to intensive study of cost data standardisation and it was considered essential that if satisfactory estimate accuracies were to be obtained for a large number of processes then the capital cost data being used would have to be standardized to such a point where it was consistent and comparable. This conclusion was enforced by a further initial modelling experiment in which 14 cost data sets from the same source (i.e. Lummus, the engineering contractor) were selected and modelled with the following result:-

$$C = 1.98 \times 10^{-3} (N) (Q_{mw})^{.57} \quad R^2 = .95, \ SE = .14$$

This was considered to be a very significant experiment. The 14 data sets used covered a wide variety of processes (cyclohexane, formaldehyde, nitric acid, styrene, terephthalic acid, ethylene and vinyl chloride monomer) with widely different sizes, materials of construction and functional unit types and numbers and yet it could be seen that it was possible to achieve good accuracies provided a consistent cost base was being used.

Since it was evident that the Lummus data was consistently and significantly below comparable costs from other published sources and as such that the costs published in the literature were not being presented in a consistent manner (either intentionally or unintentionally) this need for cost standardisation was further in evidence. To summarise then, the initial modelling phase of experimentation was considered to be a valuable exercise and revealed the following points of significance:-

- the functional unit definition proposed for the research was sufficiently sound enough to achieve reasonable accuracies and could be used as the basis for further experimentation with a fair degree of confidence (even though it was likely that future experimentation on a larger data base would reveal flaws in the definition which would have to be identified and removed).
- reasonable estimate accuracies had been achieved by defining the relationship between cost per functional unit (CPF) and process size (Q) (measured as a function of process material flows) which were promising enough to prove the importance and significance of this relationship in forecasting costs and so indicate that further work should be done to accurately define the relationship and use it as a building block for the more refined models which would follow.

- a large data base was required if meaningful results were to be obtained. Initial experiments had shown how easy it was to produce good correlations over a small range of processes (14) and how misleading these could be if generally applied since models derived on the same basis for 30 data sets produced extremely poor results.
- cost data standardisation was essential and procedures for achieving this had to be produced and applied (see Chapter 4) if realistic forecasting models were to be obtained.

5.2 Finalised Cost Models

Based on the conclusions listed in the previous section, the final phase of cost modelling was implemented. The literature was thoroughly searched and 103 data sets were established and subsequently standardised using the procedures detailed in Chapter 4. As in the initial modelling phase the cost data used was escalated to the common time base of June 1976 and all models presented in this section predict costs in millions of US Dollars at this time.

The analysis and results of the cost estimating models produced from this data base, showing the relative influences on capital cost of the various process parameters discussed in Chapters 3 and 4, are presented as follows:-

1. Process Size
2. Materials of Construction
3. Process Operating Conditions
4. Functional Units
5. Location Effects

5.2.1 Process Size (Q)

From the discussion in Section 3.4.3. it was concluded that no best method existed for representing plant size in a cost model. Three alternatives were discussed (capacity, feed and throughput) all of which had advantages and disadvantages associated with their use depending mostly on the information available.

As such it was concluded that a series of models should be produced whereby cost could be predicted as a function of each of the three variables. This would create a flexible package which could provide estimates for any given situation and information constraints encountered in the early stages of process development. The objectives of experimentation on process size then were to determine the level of accuracy that could be expected from the three variables and the most suitable units of measurement (i.e. weight or volumetric) of plant size. These objectives could be met by producing simple models of cost per functional unit (CPF) as a function of the different size variables (Q).

The experiments were sequenced as follows:-

i) Capacity

As mentioned in the discussion on capacity in Chapter 3, multi-product processes pose a problem when using capacity to measure plant size, since a number of capacity values can be used to represent plant size; main stream, total or some weighted average value (as proposed in Section 3.4.3.2.1). An experiment was performed to assess which of these alternatives was the most suitable representation of multi-product process capacity. From a regression analysis of 22 multi-product process data sets the following results were obtained:-

a. Main product capacity (taken as largest product output)

$$C = 1.64 \times 10^{-3}(N)\ (Q_{mw})^{.60},\ R^2 = .949,\ SE = .255,\quad T_{STAT}(Q_{mw}) = 18.7 \tag{1}$$

b. Total capacity

$$C = 6.38 \times 10^{-4}(N)\ (Q_{TW})^{.66},\ R^2 = .938,\ SE - .281,\quad T_{STAT}(Q_{TW}) = 16.9 \tag{2}$$

c. Average weighted capacity

$$C = 1.11 \times 10^{-3}(N)(Qavw)^{.623},\ R^2 = .967,\ SE = .205,\quad T_{STAT}(Q_{avw}) = 23.5 \tag{3}$$

As expected the best results were obtained by using the weighted average capacity. It was thought that the results were conclusive enough to show that this was the best alternative for measuring multi-product process capacity and that the average weighted capacity concept was a success. It was also noted from this experiment that main product capacity gives better results than total capacity. This compliments the initial modelling experimental results which indicated that product dilution was not significant, since the dilution experiment essentially compared total output with main product output (expressed on a 100% purity basis).

Following the experiment on multi-product processes the next step was to model CpF v Capacity for single product processes, and compare the results of the two experiments. It was felt that this comparison was necessary in order to measure the significance of the previous results and to show whether or not similar accuracies could be obtained for multi-product processes as for single product processes. From a regression on 81 single product process data sets the following result was obtained-

$$C = 1.19 \times 10^{-3}(N)\,(Q_{mw})^{.623},\ R^2 = .946,\ SE = .258,$$

$$T_{STAT}\,(Q_{mw}) = 21.6 \qquad (4)$$

Because of the large difference in the number of data sets used in the two experiments it is difficult to compare experiment accuracies. However, it is interesting to note that the average weighted capacity used to represent multi-product process capacity has a similar T-STAT value and identical size exponent to the single product process capacity measurements. This indicates that the average weighted capacity measurement is consistent with and comparable to single product process capacity and as such similar estimate accuracies can be expected for both single and multi-product processes alike. Having concluded this the two data sets were combined and a regression performed on the total 103 data sets with the following result:-

$$C = 1.294 \times 10^{-3}(N)\,(Qavw)^{.615},\ R^2 = .956,\ SE = .157,$$
$$T_{STAT}\ (Qavw) = 42.46 \qquad (5)$$

where $Q_{av}\,w$ = average weighted capacity (metric tons/annum) (equivalent to main stream capacity for single product processes).

All modelling so far had consisted of using capacities measured on a weight basis. At this point the alternatives of using volumetric based measurement was investigated. A regression on the same 103 data sets as used above gave the following result:-

$$C = 1 \times 10^{-3}(N)\,(Q_{av\ v})^{.508},\ R^2 = .846,\ SE = .274,$$
$$T_{STAT}(Qav\ v) = 32.58 \qquad (6)$$

where Q av v = average weighted capacity (kilomols/annum).

It was clearly visible that volumetric based measurement was not as successful as weight based measurement of plant capacity and when used gave significantly poorer results. The variance in the two model accuracies was so great it was considered that volumetric measurement should be discounted as a viable alternative for measuring process flows and that all future modelling involving the process size variable (Q) should be based on weight measurements.

ii) Feed

A regression of 39 data sets was performed to assess the significance of process feed (see section 4.3.2.2) as a cost prediction variable. Initially, a base comparison model was derived showing cost per functional unit as a function of capacity;

$$C = 9.0 \times 10^{-4}(N)\ (Q_{av}\ w)^{.648},\ R^2 = .958,\ SE = .162,$$
$$T_{STAT}\ (Q_{av}\ w) = 20.7 \qquad (7)$$

and then process feed was modelled and the results compared-

$$C = 2.84 \times 10^{-4}(N)\ (Q_{FW})^{.686},\ R^2 = .955,\ SE = .168,$$
$$T_{STAT}\ (Q_{FW}) = 20.5 \qquad (8)$$

Although it was seen that capacity gave the better correlation (though only slightly) a very good model was also obtained by using process feed.

The accuracy achieved was sufficient enough to think that the model could be used to estimate costs where feed values are known instead of capacity, e.g. waste disposal processes.

iii) Throughput

A regression of 42 data sets was performed to assess the significance of process throughput, which was expressed as a function of capacity and reactor conversion efficiency (see section 4.3.2.3). Initially a base comparison model showing cost per functional unit as a function of capacity was produced-

$$C = 7.76 \times 10^{-4}(N)\ (Qav\ w)^{.675},\ R^2 = .947,\ SE = .243,$$
$$T_{STAT}(Q\ av\ w) = 19.9 \qquad (9)$$

and then process throughput was modelled with the following result-

$$C = 6.41 \times 10^{-4}(N)\ (Q\ av\ w)^{.67}\ (Rc)^{-1.11},\ R^2 = .9482,$$
$$SE = .238,\ T_{Stat}(Qav\ w) = 19.7,\ T_{STAT}(Rc) = -2.78 \qquad (10)$$

The introduction of the reaction efficiency variable into the model had very little effect and was virtually insignificant in affecting model accuracy. As such it was concluded that the experiment for process throughput was not a success. It was considered that although throughput was probably the best means of measuring plant size the above means used to express it were inadequate and an alternative expression was needed. To summarise this section on cost as a function of process sizes, then the following conclusions are noted:-

(i) The effect of standardising the cost data using the procedures described in Chapter 4 has been effective with greatly improved model accuracies having resulted from doing so.

(ii) Good correlations have been obtained for the relationship CpF v Q (expressed as plant capacity, feed and throughput) which will serve as a sound building block for more sophisticated modelling (future modelling fine tuning). It is noted however that the throughput definition is inadequate and the alternative form of expressing this variable is required.

(iii) Weight based measurement units give better results than volumetric based units and as such will be used for future modelling.

(iv) The assumption of linearity for the CpF v Q relationship, which is the basis for all of the models produced in this research, is considered to be valid and as such the models produced can be assumed to be valid predictions of cost within the stated limits of process size (2000 metric tpa $<$ capacity (average) $>$ 475 000 metric tpa).

(v) The averaged weighted capacity concept was successful.

Hence, it was possible to conclude that good estimating accuracies could be obtained using a simple model (CpF v Q), for a large number of different processes providing good cost data standardisation was applied.

5.2.2. Process Material of Construction

Using the procedure described in Section 3.4.4. a process installed cost material of construction factor (MOCFAC) was calculated for each of the 103 data sets being used at this stage of modelling.

Based on actual material cost ratios (fm) the individual functional unit installed cost ratios (Fm) were calculated for each material of construction encountered. These are shown in Table 3.4.4.3. Using these ratio values the overall process ratio (MOCFAC) was determined. The experiment to assess the influence of the overall process material of construction ratio (MOCFAC) on cost estimating accuracy was structured as follows:-

i) Derivation of base case model for comparison purposes

The base case model was in fact already available from the experimental results obtained in the previous section and is as follows:-

$$C = 1.294 \times 10^{-3}(N)\ (Qav\ w)^{.615},\ R^2 = .956,\ SE = .157,$$

$$T_{STAT}(Qav\ w) = 42.46 \qquad (5)$$

ii) Introduction of the material of construction variable (MOCFAC)

$$C = 9.41 \times 10^{-4}(N)\ (Qav\ w)^{.637}(MOCFAC)^{1.03}$$

$$R^2 = .976,\ SE = .117,\ T_{STAT}(Qav\ w) = 63.7,$$

$$T_{STAT}(MOCFAC) = 9.13 \qquad (11)$$

It can be seen that the introduction of the MOC variable was highly successful and that the variable is significant.

5.2.3 Process Operating Conditions

Although it was postulated in Chapter 3 that process temperatures and pressures were unlikely to have a significant influence on the installed capital cost of a process, the assumption had to be proved.

Hence, experiments were performed to measure the effect of temperature and pressure on capital cost. Only 47 data sets were used since temperature and pressure data was difficult to find. No detailed temperature and pressure profile information was available for any of the processes encountered in the literature and so modelling was restricted to using maximum process values. The experiments were structured as follows:-

i) <u>Derivation of base case models for comparison purposes</u>

$$C = 1.028 \times 10^{-3}(N)(Qav\ w)^{.623} \quad R^2 = .9465 \quad SE = .159 \tag{12}$$

and

$$C = 5.1 \times 10^{-4}(N)\ (Qav\ v)^{.567} \quad R^2 = .862 \quad SE = .252 \tag{13}$$

ii) <u>Introduction of temperature variable</u>

Two alternatives studied, a) $T_{max}(^oC)$ and b) $T_{max}\ (^oK)$

$$a)\ C = 7.4 \times 10^{-4}(N)\ (Qav\ w)^{.621}(F_T)^{.092},\ R^2 = .95,\ SE = .1524,\ T_{STAT}(F_T) = 2.58 \tag{14}$$

$$b)\ C = 4.39 \times 10^{-4}(N)\ (Qav\ w)^{.628}(F_T)^{.162},\ R^2 = .95,\ SE = .1515,\ T_{STAT}(F_T) = 2.68 \tag{15}$$

As expected the inclusion of the temperature variable had little impact. The T_{STAT} values obtained proved the statistical insignificance of the variable. The effect of the units of measurement was minimal with little difference being noted between the effects of using ^oC or ^oK (which was used to assess if gas law theory was affecting material volumes and thus cost), with ^oK giving slightly better accuracy.

iii) <u>Introduction of pressure variable</u> - (maximum process pressure measured in atmospheres)

$$C = 8.386 \times 10^{-4}(N)\ (Qav\ w)^{.663}\ (F_p)^{-.033}$$

$$R^2 = .946,\ SE = .158,\ TSTAT(F_p) = -1.74 \qquad (16)$$

Again, as expected the inclusion of the pressure variable into the model shows little impact on prediction accuracy when compared to the base case.

iv) <u>Introduction of both temperature and pressure variables</u>

$$C = 4.087 \times 10^{-4}(N)\ (Qav\ w)^{.643}\ (F_T)^{.144}\ (F_p)^{-.024}$$

$$R^2 = .9505,\ SE = .158,\ T_{STAT}(F_T) = 2.64,$$

$$T_{STAT}(F_p) = -1.72 \qquad (17)$$

and

$$C = 2.17 \times 10^{-4}(N)\ (Qav\ v)^{.542}\ (F_T)^{.183}\ (F_p)^{-.014}$$

$$R^2 = .865,\ SE = .25,\ T_{STAT}(F_T) = 2.66,$$

$$T_{STAT}(F_p) = -1.69 \qquad (18)$$

The above models prove further that the effect of including process operating condition parameters in the estimating models has little impact on base case accuracies and the above results serve to validate the assumption that process operating conditions are not significant parameters affecting installed capital cost and should not therefore be included in cost estimating models.

A further experiment involving process operating conditions was to compare the significance of the materials of construction parameter with the temperature and pressure parameters on estimating accuracy.

v) <u>Introduction of material of construction parameter (MOCFAC)</u>

$$C = 8.443 \times 10^{-4}(N)\ (Qav\ w)^{.646}\ (MOCFAC)\ 1.08,\ R^2 = .975,$$
$$SE = .107,\ T_{STAT}\ (MOCFAC) = 7.71. \qquad (19)$$

It can be seen that the material of construction parameter has a far greater influence on accuracy than process operating conditions, with significant improvements being achieved over the base case models by its inclusion. A further experiment also appeared to validate the theory that the material of construction parameter was accounting for some temperature and pressure effects as the model below indicates.

$$C = 6.041 \times 10^{-4}(N)\ (Qav\ w)^{.639}(MOCFAC)^{1.02}\ (F_T)^{.066}$$
$$(F_p)^{-.016},\ R^2 = .976,\ SE = .113,\ T_{STAT}\ (MOCFAC)$$
$$= 5.23,\ T_{STAT}(F_T) = .1421,\ T_{STAT}(F_p) = .118 \qquad (20)$$

Again it can be seen that the material of construction is more dominant that operating conditions in predicting costs, but the significance of the variable is shown to be reduced when temperature and pressure variables are modelled with it. This would be expected. Also, it was noted that the above model had a lower accuracy than the previous one which included a material of construction parameter only (without process operating conditions). This served to show that the inclusion of process operating conditions in the above form (i.e. maximum process values) can give misleading and incorrect results and as such should be done so with a great deal of caution.

To summarise then, from the experiments on process operating conditions it was possible to conclude that:-

a) Detailed process operating condition data is not available in the literature for use in rapid cost modelling. As such, it is not possible to derive realistic measurements of temperature and pressure profiles for processes and so to adequately represent them in a cost model. The only viable method was considered to be to use maximum temperature and pressure values. This was recognised as being essentially incorrect and in many cases misleading but no alternatives were available.

b) Process operating conditions expressed in the above form (T_{max}, P_{max}) are not significant influences on installed capital cost, their inclusion giving almost unnoticeable improvements in estimating accuracy (or even decreasing estimate accuracy in some cases - see final model).

c) Materials of construction have far more effect and the inclusion of this parameter was shown to take into account some of the effects on cost exerted by operating conditions.

To conclude, it was considered that process operating conditions should not be included in rapid pre-design cost estimating models because it was considered that the effect of their inclusion would be minimal (or even misleading). Whether this is due to their having no influence or whether it is due to the fact that they cannot be adequately represented in a cost model is uncertain.

5.2.4 Functional Units

All the experiments performed so far consisted of modelling cost per functional unit as a function of process size and materials of construction as follows:-

$C_pF = \frac{C}{N} = f\ (Q, \text{MOCFAC})$ i.e. N is an independent variable.

In order to assess the functional unit definition for consistency and accuracy however, it was thought advisable to make N a dependent variable and the following relationship modelled:-

$C = f(N, Q, \text{MOCFAC})$ i.e. N is a dependent variable.

The relationship was modelled using all of the 103 data points with the following result:-

$$C = 9.37 \times 10^{-4} (N)^{1.01} (Q\ av\ w)^{.636} (MOCFAC)^{1.036},$$
$$R^2 = .997, SE = .119 \qquad (21)$$

It could be seen that the result was extremely encouraging. The index value of 1.01 on N clearly indicated that the functional unit definition being used was cost consistent. As such it was felt that the definition that had been derived for the thesis was a success and was an improvement on the previous offerings that had been put forward. It is interesting to note that the index on N is >1. Though very small for the model above it is more marked in later models (see following section), rising to 1.06. This would seem to indicate that costs increase significantly as the process complexity increases, possibly due to increased engineering effort and process control requirements.

5.2.5 Location Effects

As stated in Chapter 4 (section 4.1.2.3) at the time of official research completion it was considered that location effects were insignificant due to the constant flux in exchange rates and so no work was done in this area.

However, in later discussions, Bridgwater (40) disputed this and, whilst conceding that this may have been the case at mid-1976, showed evidence that location effects did still exist in mid-1978. In the interests of estimating accuracy it was considered worthwhile to apply location factors (see Table 5.2.5.1) to the cost data. The data was escalated to mid-1978 time base using the EPE cost index and in addition, standardised to a location base (selected as the US Gulf Coast because most of the data used referred to this location and so errors involved in the location factor were minimised). The results are shown and, where applicable, compared against pre-location factor models.

		R^2	SE	T_{stat} Qavw	T_{stat} N	T_{stat} MOCFAC
$C = 4.7 \times 10^{-3} (Qavw)^{.675} (LOCFAC)$	-(22)	.543	.739	10.95		
$C = 1.006 \times (N)^{1.15} (LOCFAC)$	-(23)	.494	.778		9.92	
$C = 1.24 \times 10^{-3} (N) (Qavw)^{.67} (LOCFAC)$	-(24)	.965	.154	46.2		
c.f. $C = 1 \times 10^{3} (N) (Qavw)^{.615}$	- (5)	.956	.154	42.5		
$C = 9.25 \times 10^{-4} (N) (Qavw)^{.663} (MOCFAC)^{.96} (LOCFAC)$	-(25)	.978	.114	64.7		9.18
c.f. $C = 9.41 \times 10^{-4} (N) (Qavw)^{.637} (MOCFAC)^{1.03}$	-(11)	.976	.117	63.7		9.13
$C = 8.67 \times 10^{-4} (N)^{1.06} (Qavw)^{.66} (MOCFAC)^{.89} (LOCFAC)$	-(26)	.985	.111	56.3	52.1	6.24
c.f. $C = 9.37 \times 10^{-4} (N)^{1.01} (Qavw)^{.636} (MOCFAC)^{1.036}$	-(21)	.978	.113	56.0	51.8	6.18

TABLE 5.2.5.1

LOCATION FACTORS FOR U.K. AND U.S.A. FOR CHEMICAL PLANT OF SIMILAR FUNCTION (78)

		UK = 1.0	USA = 1.0
AUSTRALIA		1.4	1.3
AUSTRIA		1.1	1.0
BELGIUM		1.1	1.0
CANADA		1.25	1.15
CENTRAL AFRICA		2.0	2.0
CENTRAL AMERICA		1.1	1
CHINA	Imported element	1.2	1.1
	Indigenous element	0.6	0.55
DENMARK		1.1	1.0
EIRE		0.9	0.8
FINLAND		1.3	1.2
FRANCE		1.05	0.95
GERMAN (WEST)		1.1	1.0
GREECE		1.0	0.9
HOLLAND		1.1	1.0
INDIA	Imported element	2.0	1.8
	Indigenous element	0.7	0.65
ITALY		1.0	0.9
JAPAN		1.0	0.9
MALAYSIA		0.9	0.8
MIDDLE EAST		1.2	1.1
NEWFOUNDLAND		1.3	1.2
NEW ZEALAND		1.4	1.3
NORTH AMERICA	Imported element	1.2	1.1
	Indigenous element	0.8	0.75
NORWAY		1.2	1.1
PORTUGAL		0.8	0.75
SOUTH AFRICA		1.25	1.15
SOUTH AMERICA (NORTH)		1.5	1.35
SOUTH AMERICA (SOUTH)		2.5	2.25
SPAIN	Imported element	1.3	1.2
	Indigenous element	0.8	0.75
SWEDEN		1.2	1.1
SWITZERLAND		1.2	1.1
TURKEY		1.1	1.0
U.K.		1.0	0.9
U.S.A.		1.1	1.0
YUGOSLAVIA		1.0	0.9

NOTES:

a. Increase factor by 10% for each 1000 miles or part 1000 miles new location is away from major manufacturing and/or import centre.
b. For multisourcing of materials and/or labour use appropriate factors proportionately.
c. Investment incentives are ignored.

When comparing the models before and after compensation for location effects it is seen that in all cases where comparisons exist the major indicator of model reliability, Standard Error, showed a marginal improvement. Hence, it is possible to conclude that location effects on costs do exist and that the factors used are reasonable indicators of the extent of those effects.

5.3 Model Testing

The models listed in the previous section are the most sophisticated that will be produced in this thesis. The next phase in the research required that they be tested. The testing procedures used are described here.

5.3.1 Linearity Testing

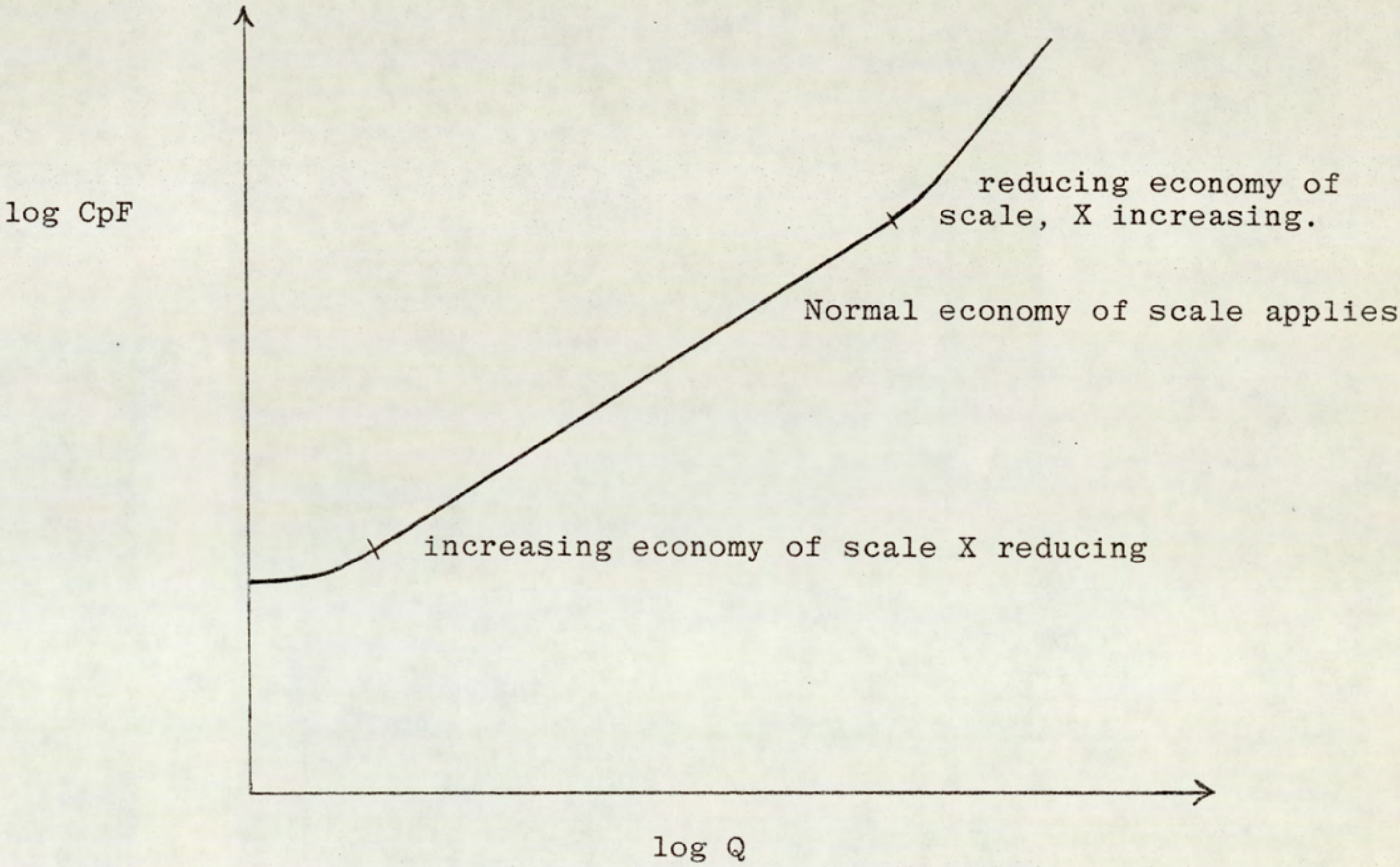
The basis of all the models produced in this research is that the cost per functional unit (CpF) v process size (Q) relationships defined so far via the regression analysis modelling, are assumed to be logarithmic, such that if plotted on a log-log graph a straight line relationship could be obtained. The relationship is usually expressed as:-

$$CpF = K. (Q)^{X}$$

where K = constant

X = scale factor

This assumption however, may not be valid and it has recently been postulated that a curve relationship exists as follows (20):-



Although no evidence has been published in the literature to support the curve relationship theory, logical reasons do exist to indicate that it does apply and represents a more accurate description of the cost-size relationship than the generally assumed straight line. These reasons have already been identified and discussed in the analysis of the exponential method of capacity adjusted covered in the literature survey (see section 2.2.1.3) and will not be detailed again here.

Having accepted that the relationship may not be truly logarithmic then the extent of the non-linearity had to be determined. If the curve is such that significant variation from the postulated straight line exists then the base assumption of modelling is incorrect and the models produced would not be viable cost predictions (particularly for low and high plant capacities). If the variance is insignificant however, then the assumption of linearity may still be used with a high degree of confidence. Hence, an experiment was performed to check the linearity assumption of the CpF v Q relationship. A regression analysis on 103 data sets was carried out. The cost per functional unit was calculated for each process. Each cost was standardised to the U.S. Gulf Coast location, mid-1978 time base, and a carbon steel material of construction base. This cost was plotted against the average weighted capacity. The result was as follows:-

$$C = 1.23 \times 10^{-3} (N) (Qav\ w)^{.62}$$

Durbin-Watson statistic = 1.85

It is the Durbin-Watson statistic that is significant here. It is a direct indication of the degree of linearity. A brief summary of the theory of it's use will be given. For a truly linear relationship a Durbin-Watson value of 2.0 would be expected. However, depending on the number of data points used in the correlation, and thus the degree of confidence in the relationship, acceptable divergant limits from the 2.0 value exist (either + or -) for the assumption of linearity to be acceptable. In this case, the limits of acceptance for the sample of 103 data sets being used are 1.76 to 2.24 (91). Hence, it can be seen that the obtained value of 1.85 is within the accepted limits and so it is possible to assume that the relationship is sufficiently linear for the models to be valid within size parameters used in the research (5540 $<$ Q av w $>$ 864000 tonnes/annum).

5.3.2 Accuracy Testing

The accuracy of the most sophisticated model produced:-

$$C = 8.67 \times 10^{-4}(N)^{1.06} \ (Qav\ w)^{.66} \ (MOCFAC)^{.89} \ (LOCFAC)$$

was found by examination of the residuals, to be +25% to -20% within 95% confidence limits (standard error ± 14%) for the 103 data sets used in the research. This was within the parameters set by the research aims and considering the large and diverse nature of the cost data used, was considered to be extremely good. However, it was realised that for these claims to be accepted the model would have to be tested on very high quality cost data in which the cost and process parameters were all known values. Two processes of known reliable cost were available to test the model. These were:-

. The BASF process for Butadiene production (which was actually used in the research)

. Acetic Anhydride by acetone pyrolysis.

The results were as follows:-

(1) BASF process (used in derivation of model)

Cost Data   Battery limits cost = \$3.6 million

Location = USA

Year of construction = 1971

Actual base cost (USA/1978) = 3.6 x 182/104

= \$6.3 million.

Process Data   Materials of construction = Carbon steel

(MOCFAC = 1.0)

Av. Capacity = 49,890 tonnes/annum

Number of functional units = 5 (absorption, stripping, washing, scrubbing, propyne distillation)

$$\text{Estimated cost, } C_{BASE} = 8.67 \times 10^{-4}(N)^{1.06}\ (Qav\ w)^{0.66}\ (MOCFAC)^{.89}\ (LOCFAC)$$

$= 8.67 \times 10^{-4}$ (5.5). (1261). (1). (1)

= $6.013 million

= 95% of actual cost

(2) Acetic Anhydride Process (not used in derivation of model)

Cost Data   Battery limits cost = £3.65 million

Location = U.K.

Year of construction = 1978

Actual base cost (US/1978) = £3.65 million

x 1.83 ($)

÷ 0.9 (LOCFAC)

= $7.42 million

Process Data Materials of construction = Stainless steel

(MOCFAC = 1.285)

Average capacity = 20,000 tonnes/annum

Number of functional units = 5 (reaction, quench, absorption, acetone distillation, acetic anhydride distillation)

$$\text{Estimated cost, } C_{BASE} = 8.67 \times 10^{-4}\ (N)^{1.06}\ (Qav\ w)^{0.66}\ (MOCFAC)^{.89}\ (LOCFAC)$$

$$= 8.67 \times 10^{-4}\ (5)^{1.06}\ (20000)^{0.66}\ (1.285)^{.89}\ (1.0)$$

$$= 8.67 \times 10^{-4}\ (5.5)\ (690)\ (1.25)\ (1.0)$$

= $4.11 million (£2.25 million, USA basis)

= 55% of actual cost

It may be seen that these two results taken in isolation are inconclusive. The butadiene process estimate is very good whilst the one for the acetic anhydride process is very poor and, whilst the model gives low estimates for both processes, it is not possible to produce a definite conclusion on its performance. Bridgwater however, has tested other step counting techniques on the acetic anhydride process (20) and thus it is possible to compare this correlation against other proposals in the field. The results given by the other methods ranged from £2.147 million up to £3.17 million with the mean value being £2.65. Thus, the research model compares very unfavourably with other techniques in this instance and the results given by it would seem to be unacceptably low. Although no further evidence can be published to support this statement the model has been further tested on confidential industrial data and the same low results achieved. Results obtained indicate that costs are being estimated at between 55-65% of actual values.

The possible reasons for the low estimating tendency are examined in the next chapter.

6 DISCUSSION

Identification of the reasons for the poor estimating accuracy achieved is difficult because the problem(s) may be in one or more of several areas: the modelling theory, the cost data standardisation procedures, and/or recent radical changes in capital cost structures not verified with either of the previous areas.

6.1 Theory

The theory proposed to derive the cost models was thought to be sound. To summarise:

Functional Units - the definition proposed is the best so far and together with the detailed list given in the thesis, overcomes many of the problems associated with previous attempts which were, in many cases, open to interpretation and misinterpretation. The success and cost-consistency of the definition were clearly demonstrated when N was made a dependent variable in the cost models with the result that an index value of very close to unity was obtained, thus validating the functional unit approach.

Process Size - again another improvement was obtained on previous efforts. The CpF v Q relationship was shown to be a very powerful cost predictor with good accuracies being obtained with the source data. Also, the concept of the average weighted capacity was successful, and to a large extent has overcome the problem of multi-product processes.

Materials of Construction - it was clear that the introduction of a process material of construction factor greatly improved estimate accuracy. It was derived using simple theory and, though crude, is effective. However, it is noted that discrepancies exist between the research derived factors and those of Stallworthy and the two need to be reconciled.

Process Operating Conditions - these were generally shown to be insignificant as cost predictors and so were largely ignored. This is considered valuable in itself as it enables research effort to be applied to more productive areas.

Model Constant (K) - this is a possible weak point in the research. Although the linearity testing showed that the relationship $CpF = K.\ Q^n$ was acceptable for modelling purposes, the relationship needs to be examined in detail and the variance of K with Q clearly identified.

The above points (with the exception of that of the model constant) serve to support the belief that the theory was basically sound and, whilst accepting that further research on all aspects of the theory is desirable, it is not believed to be the cause of the low estimating accuracy. This conclusion is also supported by the fact that it would have been impossible to have achieved the good accuracies obtained over the large data base used in the research, if any part of the theory was significantly suspect. Therefore, the indications are that the problems lies in one of the other areas of cost data standardisation or recent unidentified factors.

6.2 Cost Data Standardisation

Project Cost Definition - It is acknowledged that some errors will have arisen in relating different cost definitions to the battery limits base level of 140, due to the number of assumptions involved. Most of the data used was quoted as 'battery limits costs' and therefore relating different project cost definitions is not believed to have caused errors. While it is possible that the 140 level is too low, there is sufficient evidence that this is a reasonable base to preclude it from further consideration. Another possibility is that the published data used to derive the models was artificially low for commercial reasons, but this seems unlikely with such a wide data base and could not be proved.

Location Factors - These were clearly shown to improve estimate accuracies. Owing to their small effect on few of the data they cannot be considered to be the cause of the low cost data levels, and so cannot be considered as a prime cause of the low test result.

Learning Effects - These were not investigated in the research although they are known to exist. Their inclusion however would only have added to the problem as they would have generally acted to reduce cost levels and due to their undeterminate nature may have reduced the accuracy.

Market Effects - These were also not investigated in the research, although it is known that they can be powerful influences on costs. Their inclusion would again have had the effect of reducing the levels since the industry world wide has experienced low activity throughout the 1970's which has made for keen competition and tight costs.

Cost Indices - These are one of the few remaining possible problem areas. There is already evidence that long standing cost indexes have a poor current basis and do not adequately reflect capital cost changes with time.

For reasons previously described in the section on cost indices, the EPE index was used to update costs. Although it is not the most sophisticated index available it does not show any significant variation with comparable indices and as such was considered satisfactory for use in the research. Therefore the choice of index was not considered to be a contributory factor to the suspected low cost level of the data base.

However, an index problem does exist because of two basic defects in their derivation and use. All cost indices are compiled such that components of overall plant cost are weighted according to their supposed significance, the total of the weightings being unity. The component costs ratios are statistically derived from historical cost data. The theory is that rises in the component costs will be monitored and the resulting cost index value derived will be a direct indication of the escalation of overall plant cost. Hence, it is clear that a basic assumption exists in using any cost index that the component cost ratios will remain constant throughout the escalation period and will not vary with time. This is not so however. The ratios do change, for example:

. a combination of high labour rate escalation and a general decline in construction productivity have led to labour cost components increasing significantly. Most indices however are failing the monitor the full extent of this change as they are only taking the increase in labour rates into account.

- many costs such as scaffolding have increased sharply as a result of safety legislation thus increasing labour overhead cost ratios.
- engineering costs are becoming more significant as labour costs have increased, and process complexity has increased. In general, these changes are not reflected in the indices published in the literature.

The second area concerns the assumption that the hardware composition of a process plant remains constant through time. It takes no account of process technology improvements or the changes in design philosophies which constantly occur to meet the changing requirements of the industry. These changes have been all too apparent over the past decade - for example:

- energy conservation requirements have led to increased heat transfer equipment and insulation in plants.
- safety and environmental legislation has to be considered necessitating additional hardware and processes.
- computerized control has been widely developed and accepted as a means of improving operating efficiency with the results that instrumentation costs are now markedly higher than in older plants.

It appears then that process plants in general have, by necessity, become far more sophisticated over the past few years and that they will continue to grow in complexity as new demands arise. However, no published cost index is capable of measuring this growth.

It is now well known that traditional cost indices have not accurately reflected plant cost increases over the past decade. In 1977 Taylor (45) suggested that real cost increases since the early 1970's have been 30% greater than estimated by a cost index, which has been confirmed by other researchers. If this is so, then it serves to explain in some way the significant errors induced by using the correlations derived in the research. Most of the cost data used was pre-1973, which was indexed to the study base time of mid-1978. If current cost indices are not correctly escalating capital costs since the oil crisis of 1972/73, as suggested by Taylor and others, then use of correlations based on pre-1973 data, when updated to the present, will give unreasonably low estimation. This view is supported by the test results. The butadiene process estimate is fairly accurate, as would be expected, since the base data used was from 1971. Conversely, the acetic anhydride process estimate is very low, the base data being from 1978. Correction of the correlations to give realistic current costs is achieved by increasing the models by an empirical factor 1.6. This has been derived after considering the various published comments, as yet unpublished research, on cost indices, and comparison of current estimates from the correlations with a variety of authoritative "real" costs. While the credibility of this approach might be considered marginal, the best justification for this approach to capital cost estimation is that it works! The revised correlations is included in the conclusions.

This allowance also fully accounts for technological and legislative constraints already discussed and briefly reviewed in the following section.

To conclude, it is clear that cost indexing has serious problems. In general, they are inadequate to reflect technological and energy changes and effort is required to derive new ones which accurately reflect cost trends of process plants throughout time.

6.3 Technology and Legislation Changes

The third possibility which is known to exist and is closely related to the inadequacy of cost indices, is the effect of a changing world of more severe constraints on costs. Two well known examples are environmental legislation and health and safety requirements. These increase costs by requiring additional expenditure and/or processing requirements, which are not necessarily reflected in the number of functional units or steps. This effect is not explained but is implicitly included in the multiplicative factor of 1.6 deduced above.

7 CONCLUSIONS

In broad terms the research has, in most cases, achieved its objectives (as described in Section 2.3 - Research Aims) which were to derive an estimating technique which was:

. quick and easy to use.

. capable of utilizing a wide range of process information so that it could fulfil its primary aim of assisting in screening studies.

. flexible enough to cover a wide range of processes.

. accurate to within acceptable limits.

The success in meeting these objectives is discussed below and the conclusions of the thesis presented.

(i) Ease of Production and Use

This objective was fully met. Although, by definition, step counting methods are specifically designed to enable process engineers to produce cheap and quick estimates, the relatively simple correlations produced are advantageous compared to other procedures.

(ii) Process Information Utilization

That this objective was met is perhaps the most satisfying aspect of the research. All step counting techniques are primarily designed to assist in screening study activities where usually only the minimum amount of process information exists. Depending on the type of the process under review and its stage of development the type and amount of information will vary. However, a number of models have been derived which, when combined, provide a comprehensive series capable of utilizing all the information that is likely to be available during screening studies. i.e. number of process steps, process size (measured as capacity, feed, average capacity or throughput), materials of construction and process operating conditions, together with non-process data such as plant location. Hence the technique has no obvious information constraints within its intended field of application.

(iii) Flexibility

Originally it was intended that the research should produce models which were applicable to any type of process. Essentially, this meant that models would be derived for application to all process phase types (either fluid, liquid or solid). Due to time constraints and the scarcity of cost data for liquid and solid phase processes, this objective was regretfully abandoned, and only gaseous phase plants were covered (mainly petrochemical and organic chemical processes).

(iv) Accuracy

This was undoubtedly the most disappointing aspect of the research. Although excellent accuracies were achieved for the 103 sets of research data, the test results for the Acetic Anhydride process and other data, and the subsequent comparison of the technique against similar methods in this field, indicated that all the models gave extremely low results. It is thought that they predict costs at between 55-65% of actual values. The most significant factor is believed to be that simple cost indexing is inadequate to account for technological development and real energy cost increases.

It is recommended that all the models produced in the research study should be multiplied by 1.6 to make them realistic cost predictors. The final series of correlations is thus:

$$C = 7.52 \times 10^{-3} \, (Qavw)^{.675} \, (LOCFAC)$$

$$C = 1.61 \times (N)^{1.15} \, (LOCFAC)$$

$$C = 1.98 \times 10^{-3} \, (N) \, (Qavw)^{.67} \, (LOCFAC)$$

$$C = 1.48 \times 10^{-4} \, (N) \, (Qavw)^{.663} \, (MOCFAC)^{.96} \, (LOCFAC)$$

$$C = 1.39 \times 10^{-3} \, (N)^{1.06} \, (Qavw)^{.66} \, (MOCFAC)^{.89} \, (LOCFAC)$$

where C = cost in million US \$ at US Gulf Coast, mid-78.

8 RECOMMENDATIONS FOR FUTURE WORK

The conclusions presented in the last chapter show that while the stated research objectives were largely achieved, there is room for improvement. It is possible to identify certain problem areas which deserve major effort in future research and which could lead to significant improvements in this field of estimating.

The recommendations for future work are as follows:

Theory

Although generally considered a successful aspect of the thesis, continual refinement of the theory used to derive the models is considered essential if progress is to continue. Two major areas of model theory stand out as requiring more attention.

i) The model constant, K. The extent of its variance with process size needs to be measured and correlated.

ii) Process phase effects. Future research needs to include liquid and solid phase processes so that a fully comprehensive series of models is produced which, when combined with those covering fluid phase processes, will be capable of estimating the cost of any type of process (via the sub-process approach described in Section 3.4.1). This should be given high priority. If such a group of models could be obtained which, like those produced here, utilise <u>all</u> aspects of minimum process information, a means for providing a <u>consistent</u> estimating basis for the early screening of different process alternatives would be available. This is a major requirement in this field of study at the moment.

<u>Cost Data Analysis and Standardisation</u>

The major causes of the low estimating tendencies of the research models are thought to arise from:

i) The relatively poor quality data base, and

ii) The lack of information and thus techniques, available for equating cost data from different locations and time to a common base. Of particular concern is the lack of an adequate cost index. Hence, it is apparent that if future research is to produce accurate estimating techniques, it is essential that a sound cost data base be used in their derivation.

Two options exist for producing such a base:

- Preferably clearly defined industrial source data of known quality should be obtained from various interested parties and a cost data library compiled. The library should be continually maintained so that only recent cost data is stored and cost levels are realistic.

  This would overcome the low estimating tendency and minimise the problems associated with cost data standardisation.

This option, however, is unlikely ever to materialise. Academic calls for industrial cost data have in part largely been rejected and it is difficult to envisage industrial sources supplying the necessary quantity and quality of data required. This is unfortunate. Academic research does have value and deserves to be encouraged.

. It is probable then that academic research will continue to be based on published cost data which tends to be ill defined and of suspect quality. Furthermore, because of the requirement of a large data base and because of the scarcity of 'USABLE' cost data in the literature, the objective of only using recent data is unrealistic and 'old' data will have to be utilised. Hence, cost data standardisation will continue to be of great significance and considerable research effort will be required in this area, particularly in the area of cost indexing.

With either of the above standardisation options, considerable time and effort will be involved in searching out, screening, standardising cost data and finally compiling a data base for use in a research programme. If this responsibility is left for future individual researchers, it is bound to have an adverse affect on the time available for the theoretical and experimental aspects of their work. This in turn must serve to hamper the rate of progress of the research as a whole in this area.

To overcome this problem the research centres in this field should develop a centralised cost data management facility. Resources need to be pooled and information transferred. For example, cost data references could be shared and a reference library compiled (as per Appendix IV).

This suggestion has been proposed several times, but nothing has yet materialised.

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APPENDIX 1 - NOMENCLATURE

The abbreviations used in the experimental work and results chapters are described as follows:-

Cyoc/loc	=	Capital cost of plant at location in year of construction.
YOC	=	Year of Construction.
LOC	=	Plant Location.
CBASE	=	Standardised Capital Cost (Battery Limits investment, U.S.Gulf Coast, mid-78) - million U.S.$.
N	=	Number of process funcational units.
CpF(BASE)	=	Cost per functional unit (standardised) - million U.S.$.
QMW	=	Process capacity (main product stream), kilo tonnes/annum.
QMV	=	Process capacity (main product stream), kilogram mols/annum X 10^3.
QTW	=	Total process capacity, kilo tonnes/annum.
QTV	=	Total process capacity, kilogram mols/annum X 10^3.
QFW	=	Process feed (excluding utilities), kilo tonnes/annum.
QFV	=	Process feed (excluding utilities), kilogram mols/annum X 10^3.
QAV	=	Average process capacity (for multi-product processes), kilo tonnes/annum.
Rc	=	Reaction (if present) conversion per pass.
T	=	Maximum process temperature (^{o}C).
P	=	Maximum process pressure (atm).
MOCFAC	=	Process material of construction factor.
LOCFAC	=	Process location factor (at mid-1978).

APPENDIX 2 - CAPITAL COST DEFINITIONS

The capital cost definitions encountered in the literature are described and defined as follows:-

Fixed Capital Investment

Fixed capital investment is generally taken to comprise the following capital costs:- (9)

1) The capital necessary for major process equipment items with all the ancilliaries needed for complete process operation such as piping, insulation, electrical equipment, steel and civil works.
- generally referred to as on-site Capital Costs.

2) The capital needed for the process buildings, administrative and other offices, warehouses, transportation, shipping and receiving facilities, utility and waste disposal facilities and other permanent parts of the plant.
- generally referred to as off-site Capital Costs.

The costs incurred are for the purchase and installation of the above items, plus indirect expenses such as field office and supervision, home office overheads, miscellaneous construction costs, engineering expenses, contractor fees and contingencies.

Total Investment/Grass Roots Investment

Total plant investment for a plant consists of the fixed capital investment, working capital, start-up charges and land costs.

Working capital and land investment are defined as follows:-

Working Capital Investment

Working capital, the cash in hand required to begin plant operations, is defined by Bauman as the funds in addition to the fixed capital and land investment which a company must provide a project with to get it started and meet subsequent obligations as they come due. It includes:-

1) Raw materials and other supplies carried in stock. Usually taken as one months supply valued at delivered prices.

2) Finished products in stock. These have an average value equal to the total manufacturing cost for one months production.

3) Semi-finished products in the process of being manufactured.

4) Cash kept on hand for monthly payment of operating expenses, such as salaries and raw materials purchase.

4) Receivable credits; that is unpaid accounts against products already delivered. An average allowable 30 day payment period is usual to customers and so these credits are normally taken to amount to the production cost for one month of operation.

6) Accounts receivable.

7) Taxes payable. It is generally taken to be 10 to 20% of total investment or alternatively, 15 to 25% of turnover. However, if unusually large stocks of raw materials or products have to be held for appreciable periods of time, as the companies producing products of seasonal demand, then the working capital may account for over 50% of the total investment.

Land Investment

Self explanatory. Usually amounts to between 1 and 2% of total investment.

Battery Limits Capital Investment

Battery limits investment, generally referred to as on-site investment, is defined as:- (34), (65)

> the boundaries enclosing a plant or process unit so as to include those facilities directly involved in the conversion of raw material to finished product. It applies to all buildings, equipment, piping, instruments etc., that are specifically involved in the process or manufacturing operation. It does not include provision of storage, utilities, administrative and auxilliary buildings.

The cost of installing the battery limits of the plant is the battery limits investment.

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	ACETAL-DEHYDE	ACETAL-DEHYDE	ACETONE	ACETIC ESTER	ACETIC ESTER	ACRYLIC ACID	ACRYLIC ACID
PROCESS	CHISSO	UHDE	UHDE	IMI	IMI	TOYO SODA	NIPPON SHOKUBAI
Cyoc/Loc	0.8m$	2.86m$	2.5m$	3.1m$	3.9m$	1708mYen	2419mYen
yoc	1966	1969	1969	1974	1974	1968	1970
Loc	USA	USA	USA	USA	USA	Japan	Japan
C(BASE)	1.91	5.87	5.14	4.41	5.64	17.55	20.18
N	3	4	7	5	5	14	13
CpF(BASE)	0.64	1.47	0.74	0.88	1.13	1.25	1.55
Qmw	21.77	68.0	24.0	18.15	27.21	77.08	68.0
Qmv	494.8	1546.0	414.6	177.8	266.8	1070.5	944.4
Qtw	21.77	68.0	24.0	18.15	27.21	79.3	70.0
Qtv	494.8	1546.0	414.6	177.8	266.8	1070.5	927.0
Qfw	-	174.6	57.1	-	-	833.4	626.6
Qfv	-	685.8	1770.2	-	-	32300.0	24285.0
Qav	21.77	68.0	24.0	18.15	27.21	53.7	53.9
Rc	1.0	0.95	1.0	1.0	1.0	0.85	0.85
T	70	100	100	140	140	350	350
P	2.5	-	15	9.5	9.5	1.0	1.0
MOCFAC	1.0	1.0	1.0	1.285	1.285	1.183	1.197
LOCFAC	1.0	1.0	1.0	1.0	1.0	0.9	0.9

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	ACRYLIC ACID	METHYL ACRYLATE	ACRYLO NITRILE	AMINES	ANILINE	BENZENE	BENZENE
PROCESS	NIPPON SHOKUBAI	NIPPON SHOKUBAI	MONT-EDISON	LEONARD	HALCON	HDA	HDA
Cyoc/Loc	2112mYen	307mYen	9000mLire	1.6m$	7.5m$	0.8m$	0.98m$
yoc	1970	1970	1972	1973	1976	1965	1965
Loc	Japan	Japan	Italy	USA	USA	USA	USA
C(BASE)	17.63	2.54	20.5	2.13	9.1	1.84	2.1
N	8	5	10	5	5	2	2
CpF(BASE)	2.2	0.51	2.05	0.43	1.82	0.92	1.05
Qmw	68.3	22.1	120.0	4.5	90.7	29.5	42.0
Qmv	903.8	232.6	2264.2	-	975.3	378.7	538.9
Qtw	75.2	22.1	180.0	9.1	90.7	29.5	42.0
Qtv	921.6	232.6	2946.0	-	975.3	378.7	538.9
Qfw	610.9	26.0	1219.0	-	-	-	-
Qfv	23678.0	500.0	42140.0	-	-	-	-
Qav	75.2	22.1	123.5	7.8	90.7	29.5	42.0
Rc	0.85	1.0	0.66	-	-	0.98	0.98
T	350	163	-	-	-	677	677
P	1.0	1.0	-	-	-	42	42
MOCFAC	1.285	1.057	1.057	1.0	1.0	1.024	1.024
LOCFAC	0.9	0.9	0.9	1.0	1.0	1.0	1.0

APPENDIX 3 EXPERIMENTAL DATA PRESENTATION

PRODUCT	BENZENE	BENZENE	BENZENE	BUTA-DIENE	BUTA-DIENE	BUTA-DIENE	BUTA-DIENE
PROCESS	HDA	HDA	HDA	UNION CARBIDE	UNION CARBIDE	BASF	BASF
Cyoc/Loc	1.4m$	2.8m$	5.0m$	1.87m$	1.87m$	1.87m$	3.6m$
yoc	1965	1965	1965	1964	1969	1969	1971
Loc	USA	USA	USA	USA	USA	USA	USA
C(BASE)	3.2	6.43	11.48	4.1	3.65	3.33	6.3
N	2	2	2	7	7	6	6
CpF(BASE)	1.6	3.22	5.74	0.59	0.52	0.55	1.06
Qmw	84.1	210.1	456.0	16.3	16.3	16.3	49.9
Qmv	1077.8	2694.5	5845.2	301.0	301.0	301.0	923.9
Qtw	84.1	210.2	456.0	16.3	16.3	16.3	49.9
Qtv	1077.8	2694.5	5845.2	301.0	301.0	301.0	923.9
Qfw	-	-	-	48.3	48.3	55.6	170.6
Qfv	-	-	-	853.9	853.9	979.0	3004.6
Qav	84.1	210.2	456.0	16.3	16.3	16.3	49.9
Rc	0.98	0.98	0.98	-	-	-	-
T	677	677	677	150	150	140	140
P	42	42	42	5	5	4	4
MOCFAC	1.024	1.024	1.024	1.0	1.0	1.0	1.0
LOCFAC	1.0	1.0	1.0	1.0	1.0	1.0	1.0

APPENDIX 3 EXPERIMENTAL DATA PRESENTATION

PRODUCT	BUTA-DIENE	BUTA-DIENE	CHLORINE	CHLORINE	CHLORINE	CHLORO-ETHYLENE	CYCLO-HEXANE
PROCESS	GEON	GEON	KELLOGG	KELLOGG	KELLOGG	TOAGOSEI	LUMMUS
Cyoc/Loc	5284m Yen	651m Yen	14.1m$	20.0m$	4.4m$	1003mYen	0.6m$
yoc	1965	1965	1972	1972	1974	1968	1967
Loc	Japan	Japan	USA	USA	USA	Japan	USA
C(BASE)	7.05	8.93	23.2	32.8	7.9	10.2	1.82
N	8	8	9	9	9	13	1
CpF(BASE)	0.88	1.12	2.58	3.64	0.88	0.79	1.82
Qmw	30.0	45.3	149.7	169.4	22.7	28.6	100.0
Qmv	555.5	839.9	2107.8	3836.6	320.4	171.3	1190.5
Qtw	30.0	45.3	149.7	269.4	22.7	60.54	100.0
Qtv	555.5	839.9	2107.8	3836.6	320.4	947.5	1190.5
Qfw	96.6	146.0	187.0	341.5	29.1	66.3	100.5
Qfv	1703.5	2575.3	6304.2	8091.9	1000.7	1954.0	5013.9
Qav	30.0	45.3	149.0	16.3	16.3	16.3	49.9
Rc	-	-	0.99	0.99	0.99	-	1.0
T	100	100	288	288	288	-	80
P	4.5	4.5	4	4	4	-	20.4
MOCFAC	1.0	1.0	1.147	1.147	1.147	1.122	1.0
LOCFAC	0.9	0.9	1.0	1.0	1.0	0.9	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	CYCLO-HEXANE	CYCLO-HEXANE	CYCLO-HEXANE	CYCLO-HEXANE	CYCLO-HEXANE	CYCLO-HEXANE	CYCLO-HEXANE
PROCESS	LUMMUS	HA-84	HA-84	HA-84	HA-84	IFP	IFP
Cyoc/Loc	0.73m$	0.6m$	1.19m$	1.42m$	1.83m$	1.35m$	2.26m$
yoc	1969	1966	1966	1966	1966	1974	1974
Loc	USA	USA	USA	USA	USA	USA	USA
C(BASE)	1.97	1.3	2.0	2.58	3.07	2.16	3.28
N	1	2	2	2	2	1	1
CpF(BASE)	1.97	0.65	1.0	1.29	1.54	2.16	3.28
Qmw	1051.1	23.1	46.0	69.0	92.0	100.0	215.0
Qmv	1251.2	273.6	547.5	821.3	1095.0	1209.5	2600.5
Qtw	105.1	23.0	46.0	69.0	92.0	100.0	215.0
Qtv	1251.2	273.6	547.5	821.3	1095.0	1209.5	2600.5
Qfw	105.7	23.9	47.8	71.7	95.6	107.2	230.5
Qfv	5269.6	1529.9	3059.8	4577.7	1137.7	7073.0	15206.5
Qav	105.1	23.0	46.0	69.0	92.0	100.0	215.0
Rc	-	-	-	-	-	-	-
T	80	200	200	200	200	493	493
P	20.4	35	35	35	35	40	40
MOCFAC	1.0	1.0	1.0	1.0	1.0	1.0	1.0
LOCFAC	1.0	1.0	1.0	1.0	1.0	1.0	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	CYCLO-HEXANONE	CYCLO-PENTANE	ETHY-LENE	ETHY-LENE	ETHY-LENE	ETHY-LENE	ETHY-LENE
PROCESS	IFP	IFP	LUMMUS	LUMMUS	LUMMUS	LUMMUS	LUMMUS
Cyoc/Loc	2.25m$	2.76mFF	41.5m$	48m$	50m$	51m$	55m$
yoc	1969	1966	1966	1966	1966	1974	1974
Loc	USA	France	USA	USA	USA	USA	USA
C(BASE)	4.8	1.43	93.6	108.3	112.8	115.0	124.0
N	8	4	14	14	14	14	14
CpF(BASE)	0.6	0.36	6.69	7.74	8.05	8.21	8.86
Qmw	20.1	7.7	453.6	453.6	453.6	453.6	453.6
Qmv	2040.8	110.0	16199.0	16199.0	16199.0	16199.0	16199.0
Qtw	20.0	7.7	479.5	699.8	716.4	928.9	907.1
Qtv	2040.8	110.0	-	-	-	-	-
Qfw	-	-	563.4	1079.9	1059.7	1453.8	1680.0
Qfv	-	-	-	-	-	-	-
Qav	20.0	7.7	365.5	575.5	716.4	770.0	863.7
Rc	0.99	0.97	-	-	-	-	-
T	160	-	-	-	-	-	-
P	10	-	-	-	-	-	-
MOCFAC	1.0	1.0	1.0	1.0	1.0	1.0	1.0
LOCFAC	1.0	0.95	1.0	1.0	1.0	1.0	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	ETHYLENE OXIDE	ETHYLENE OXIDE	ETHYLENE OXIDE	FORMAL-DEHYDE	FORMAL-DEHYDE	FORMAL-DEHYDE	HCl ACID
PROCESS	SD	SD	SD	LUMMUS	CdF	CdF	POULENC
Cyoc/Loc	2.4m$	4.9m$	8.3m$	1.36m$	5.1mFF	5.9mFF	6.5mFF
yoc	1965	1965	1965	1969	1973	1973	1976
Loc	USA	USA	USA	USA	France	France	France
C(BASE)	4.37	8.92	15.1	2.35	2.61	3.02	1.81
N	7	7	7	4	3	3	5
CpF(BASE)	0.62	1.27	2.16	0.59	0.87	1.01	0.36
Qmw	27.2	68.0	158.8	16.8	25.0	33.0	5.5
Qmv	618.4	1546.1	3608.0	559.3	833.3	1100	152
Qtw	27.2	68.0	158.8	45.4	67.5	90.0	5.5
Qtv	618.4	1546.1	3608	-	-	-	152
Qfw	62.6	156.5	365.1	-	96.37	127.1	-
Qfv	1723.7	4309.8	10057	-	3226.3	4259	-
Qav	27.2	68.0	158.8	16.8	25.0	33.0	5.5
Rc	0.64	0.64	0.64	-	0.75	0.75	-
T	-	-	-	288	600	600	1200
P	-	-	-	-	2	2	-
MOCFAC	1.0	1.0	1.0	1.285	1.285	1.285	1.285
LOCFAC	1.0	1.0	1.0	1.0	0.95	0.95	0.95

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	HCL ACID	HCL ACID	HCL ACID	HCL ACID	HCL ACID	HCL ACID	HYDROGEN
PROCESS	POULENC	POULENC	POULENC	POULENC	POULENC	HALO-HYDRAL	GIRDLER
Cyoc/Loc	8.5mFF	10.2mFF	8.7mFF	11.48mFF	13.6mFF	0.8mFF	1.8m$
yoc	1976	1976	1976	1976	1976	1973	1970
Loc	France	France	France	France	France	UK	USA
C(BASE)	5	5	7	7	7	6	7
N	2.37	2.84	2.42	3.2	3.78	3.12	3.31
CpF(BASE)	0.47	0.57	0.35	0.46	0.54	0.52	0.47
Qmw	8.3	10.4	5.5	8.3	10.4	11.5	8.4
Qmv	227.8	284.8	152	227.8	284.8	315.1	4169
Qtw	8.30	10.4	5.5	8.3	10.4	11.5	8.4
Qtv	227.8	284.8	152	227.8	284.8	315.1	4169.0
Qfw	-	-	-	-	-	-	-
Qfv	-	-	-	-	-	-	-
Qav	8.3	10.4	5.5	8.3	10.4	11.5	8.4
Rc	-	-	-	-	-	-	-
T	1200	1200	1200	1200	1200	1200	815
P	-	-	-	-	-	-	-
MOCFAC	1.285	1.285	1.285	1.285	1.285	1.285	1.0
LOCFAC	0.95	0.95	0.95	0.95	0.95	0.9	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	ISO - BUTANE	ISOPROPYL ALCOHOL	ISOPROPYL ALCOHOL	ISOPROPYL ALCOHOL	LUBE OIL	LUBE OIL	MALEIC ANHYDRIDE
PROCESS	UOP	TEXACO	TEXACO	TOKUYAMA	IFP	IFP	SAVA
Cyoc/Loc	1.4m$	19.0mDM	7.6m$	700m Yen	2.31mFF	3.51mFF	1188m Lire
yoc	1970	1972	1972	1972	1971	1972	1971
Loc	USA	Germany	USA	Japan	France	France	Italy
C(BASE)	2.58	14.34	13.7	4.74	1.37	2.28	3.82
N	2	7	7	5	2	2	7
CpF(BASE)	1.29	2.04	1.86	0.95	0.69	1.14	0.55
Qmw	63.0	100.0	100.0	30.0	28.2	39.9	10.0
Qmv	1085.6	1666.7	1666.7	500.0	-	-	100.4
Qtw	63.0	100.0	100.0	30.0	28.2	39.9	10.0
Qtv	1085.6	1666.7	1666.7	500.0	-	-	100.4
Qfw	-	520.0	520.0	160.5	-	-	473.1
Qfv	-	26666	26666	8228.6	-	-	160.4
Qav	63.0	100.0	100.0	30.0	28.2	39.9	10.0
Rc	-	0.75	0.75	0.6	-	-	0.92
T	-	140	140	265	-	-	360
P	-	80	80	290	-	-	1.5
MOCFAC	1.0	1.04	1.04	1.05	1.0	1.0	1.245
LOCFAC	1.0	1.0	1.0	0.9	0.95	0.95	0.9

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	MALEIC ANHYDRIDE	MALEIC ANHYDRIDE	METHANOL	MEK	NITRIC ACID	NITRIC ACID	NITRIC ACID
PROCESS	SAVA	SAVA	LURGI	HOESCHT	WEATHERLY	HOKO	HOKO
Cyoc/Loc	2792m Lire	1577m Lire	29.7mDM	2.8m$	1.79m$	6.0mDM	8.9m DM
yoc	1971	1972	1970	1969	1966	1966	1966
Loc	USA	Germany	USA	Japan	France	France	Italy
C(BASE)	8.98	4.93	28.54	4.11	5.61	6.79	10.07
N	7	8	8	10	3	7	7
CpF(BASE)	1.28	0.62	3.57	0.41	1.87	0.97	1.44
Qmw	30.0	12.0	299.3	10.0	86.8	24.8	44.6
Qmv	301.3	122.5	8803	138.8	1377.7	398.8	717.9
Qtw	30.0	12.0	299.3	10.0	86.8	24.8	44.6
Qtv	301.3	122.5	8803	138.8	1377.7	398.8	717.9
Qfw	1419.2	576.8	-	-	-	-	-
Qfv	48170	19526	-	-	-	-	-
Qav	30.0	12.0	299.3	10.0	86.8	24.8	44.6
Rc	0.92	0.92	-	-	0.95	0.95	0.95
T	360	360	900	-	900	900	900
P	1.5	1.5	15	-	9	8.5	8.5
MOCFAC	1.245	1.215	1.0	1.03	1.285	1.285	1.285
LOCFAC	0.9	0.9	0.9	1.0	1.0	1.0	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	NITRIC ACID	NITRIC ACID	ALPHA-OLEFINS	PARAFFINS	PARAFFINS	PARAFFINS	PARAFFINS
PROCESS	BAMAG	CONIA	MITSUI	ISOSIV	ISOSIV	MODIFIED ISOSIV	BP
Cyoc/Loc	12.14m DM	6.07m DM	6.3m$	3.1m$	3.45m$	3.5m$	1.1m£
yoc	1971	1971	1973	1969	1973	1970	1973
Loc	Germany	Germany	USA	USA	USA	USA	UK
C(BASE)	14.13	7.07	10.1	6.05	5.52	6.44	4.3
N	3	5	11	4	4	5	4
CpF(BASE)	4.71	1.41	0.92	1.51	1.38	1.29	1.08
12w	305.3	40.3	11.9	68.0	68.0	45.4	49.9
Qmv	484.5	639.0	-	-	-	-	-
Qtw	305.3	40.3	34.1	68.0	68.0	45.4	49.9
Qtv	484.5	639.0	-	-	-	-	-
Qfw	-	- -	-	-	-	-	-
Qfv	-	-	-	-	-	-	-
Qav	305.3	40.3	23.7	68.0	68.0	45.4	49.9
Rc	0.95	0.95	-	-	-	-	-
T	850	850	100	-	-	-	-
P	1.5	1.5	15	-	9	8.5	8.5
MOCFAC	1.285	1.285	1.0	1.0	1.0	1.0	1.0
LOCFAC	1.0	1.0	1.0	1.0	1.0	1.0	0.9

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	PARAFFINS	PHENOL + ACETONE	CUMENE	PHENOL + ACETONE	PHTHALIC ANHYDRIDE	PHTHALIC ANHYDRIDE	PHTHALIC ANHYDRIDE
PROCESS	BP	UOP	UOP	UOP	VON HEYDEN	BASF	BSW
Cyoc/Loc	1.97m£	32.5m$	5.7m$	26.8m$	4.2m$	3.7m$	4.8m$
yoc	1976	1975	1975	1975	1968	1967	1966
Loc	UK	USA	USA	USA	USA	USA	USA
C(BASE)	4.04	40.96	7.49	33.77	8.68	8.09	10.78
N	4	18	4	14	8	8	9
CpF(BASE)	1.01	2.28	1.87	2.41	1.09	1.01	1.12
Qmw	49.9	100.0	139.0	100.0	27.2	27.2	27.2
Qmv	-	-	-	-	183.8	183.8	183.8
Qtw	49.9	161.5	139.0	100.0	27.2	27.2	27.2
Qtv	-	-	-	-	183.8	183.8	183.8
Qfw	-	-	-	-	1008.3	876.4	-
Qfv	-	-	-	-	34027	-	-
Qav	49.9	136.0	139.0	100.0	27.2	27.2	27.2
Rc	-	-	-	-	0.66	0.76	-
T	-	-	-	-	450	450	450
P	-	-	-	-	-	-	-
MOCFAC	1.0	1.032	1.0	1.04	1.07	1.07	1.06
LOCFAC	0.9	1.0	1.0	1.0	1.0	1.0	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	PHTHALIC ANHYDRIDE	PHTHALIC ANHYDRIDE	PHTHALIC ANHYDRIDE	PHTHALIC ANHYDRIDE	STYRENE	STYRENE	TPA
PROCESS	PROGIL	PROGIL	PROGIL	PROGIL	LUMMUS	LUMMUS	LUMMUS
Cyoc/Loc	35mFF	12mFF	15mFF	30mFF	21.7m$	9.54m£	17.0m$
yoc	1973	1973	1973	1973	1973	1965	1973
Loc	France	France	France	France	USA	UK	USA
C(BASE)	15.61	5.12	6.4	12.81	48.6	23.3	38.1
N	7	7	7	7	12	12	12
CpF(BASE)	2.23	0.73	0.91	1.83	4.05	1.94	3.18
Qmw	90.0	20.0	30.0	90.0	300.0	91.4	150.0
Qmv	608.1	135.1	202.7	608.1	-	-	903.6
Qtw	90.0	20.0	30.0	90.0	319.0	97.2	150.0
Qtv	608.1	135.1	202.7	608.1	-	-	903.6
Qfw	1311	437.0	655.5	1311	-	-	-
Qfv	-	-	-	-	-	-	-
Qav	90.0	20.0	30.0	90.0	271.3	84.2	90.3
Rc	0.75	0.75	0.75	0.75	-	-	-
T	370	370	370	370	-	-	-
P	11	11	11	11	-	-	-
MOCFAC	1.08	1.08	1.08	1.08	1.0	1.0	1.14
LOCFAC	0.95	0.95	0.95	0.95	1.0	0.9	1.0

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	TPA	UREA	VCM	VCM	XYLENE	XYLENE	XYLENE
PROCESS	IFP	STAMI-CARBON	LUMMUS	LUMMUS	AROMAX	ESSO	ICI
Cyoc/Loc	4.0m$	18.8m$	15.7m$	25.0m$	2281mYen	1.5m$	0.84m£
yoc	1970	1975	1973	1973	1973	1969	1979
Loc	USA	USA	USA	USA	Japan	USA	UK
C(BASE)	7.36	23.5	35.2	55.1	14.47	2.29	5.78
N	5	6	9	9	8	4	5
CpF(BASE)	1.47	3.92	3.9	6.12	1.8	0.73	1.16
Qmw	44.7	335.3	226.8	340.2	100.0	22.7	61.0
Qmv	269.3	-	3628.5	5442.8	943.8	213.9	575.1
Qtw	44.7	335.3	226.8	100.0	22.7	61.0	27.2
Qtv	269.3	-	-	-	943.4	213.9	575.1
Qfw	-	-	226.8	400.1	-	-	-
Qfv	-	-	6392	9581	-	-	-
Qav	44.7	335.3	226.8	340.2	100.0	22.7	61.0
Rc	-	-	-	-	-	-	-
T	-	185	410	410	200	-	450
P	-	140	7	7	19	-	1
MOCFAC	1.39	1.24	1.1	1.1	1.0	1.0	1.0
LOCFAC	1.0	1.0	1.0	1.0	0.9	1.0	0.9

APPENDIX 3

EXPERIMENTAL DATA PRESENTATION

PRODUCT	XYLENE	XYLENE	WAX
PROCESS	KRUPP	MARUZEN	SFP/BP
Cyoc/Loc	3.67mDM	3117mYen	1.48mFF
yoc	1969	1973	1967
Loc	Germany	Japan	France
C(BASE)	3.82	24	1.59
N	4	10	3
CpF(BASE)	0.96	2.4	0.53
Qmw	50.8	100.0	16.5
Qmv	479.3	943.4	-
Qtw	50.8	100.0	16.5
Qtv	479.3	943.4	-
Qfw	-	-	-
Qfv	-	-	-
Qav	50.8	100.0	16.5
Rc	-	-	-
T	-	-	
P	-	-	
MOCFAC	1.0	1.0	1.0
LOCFAC	0.9	0.9	0.95

APPENDIX 4. - COST DATA REFERENCES

PRODUCT	PROCESS	REFERENCE
ACETALDEHYDE	CHISSO	CPE March 1968 p.75
ACETALDEHYDE	HOESCHT UHDE	HP November 1969 p.137
ACETONE	HOESCHT UHDE	HP November 1969 p.140
ACETIC ESTERS	IMI	HP April 1975 p.185
ACRYLIC ACID	TOYO SODA	HP May 1969 p.152
AND METHYL	NIPPON	CE October 30th 1972 p.84
METHACRYLATE	SHOKUBAI	HP November 1972 p.85
ACRYLONITRILE	MONTEDISON	CE March 20th 1972 p.80
ACRYLONITRILE	MONTEDISON	HP November 1972 p.144
AMINES	LEONARD	HP November 1973 p.151
ANILINE	HALCON	HP November 1976 p.145
BENZENE	HDA	HP May 1966 p.140
BENZENE	HDA	HP November 1969 p.153
BUTADIENE	UNION CARBIDE	CE July 31st 1967 p.70
BUTADIENE	UNION CARBIDE	HP May 1967 p.166
BUTADIENE	UNION CARBIDE	BCE October 1967 p.1509
BUTADIENE	UNION CARBIDE	CPE March 1970 p.75
BUTADIENE	BASF	CEP March 1970 p.75
BUTADIENE	BASF	CE September 1968 p.135
BUTADIENE	BASF	HP November 1968 p.65
BUTADIENE	BASF	Priv.Comm.Dr.A.V.Bridgewater
BUTADIENE	GEON	CE May 9th 1966 p.134
BUTADIENE	GEON	CPE March 1970 p.69
BUTADIENE	GEON	HP November 1966 p.151
BUTADIENE	GEON	Priv.Comm.Dr.A.V.Bridgewater
CHLORINE	KEL-CHOR 1	CEP April 1973 p.51
CHLORINE	KEL-CHOR 1	HP November 1974 p.151
CHLORINE	KEL CHLOR 11	HP November 1974 p.151
CHLOROETHYLENE	TOAGOSEI	IEC May 1970 p.31
CHLOROETHYLENE	TOAGOSEI	CE May 4th 1970 p.74
CHLOROETHYLENE	TOAGOSEI	CPE June 1966 p.268
CYCLOHEXANE	AROSAT	CEP June 1967 p.73
CYCLOHEXANE	AROSAT	HP November 1969 p.169
CYCLOHEXANE	AROSAT	HP May 1967 p.169

APPENDIX 4 (continued)

PRODUCT	PROCESS	REFERENCE
CYCLOHEXANE	HA-84	HP May 1967 p.169
CYCLOHEXANE	HA-84	CPE June 1963 p.63
CYCLOHEXANE	HA-84	CEP June 1967 p.73
CYCLOHEXANE	IFP	HP November 1973 p.117
CYCLOHEXANE	IFP	ECN April 19th 1974 p.14
CYCLOHEXANONE	IFP	CEP June 1969 p.71
CYCLOPENTANE	IFP	HP August 1973 p.105
ETHYLENE	LUMMUS	HP August 1973 p.105
ETHYLENE OXIDE	SCIENTIFIC DES.	HP October 1970 p.105
ETHYLENE OXIDE	SCIENTIFIC DES.	Chemical Week, March 5th 1966
FORMALDEHYDE	LUMMUS	CE May 18th 1970 p.118
FORMALDEHYDE	C D F	HP September 1973 p.179
HYDROCHLORIC ACID	RHONE POULENC	HP August 1976 p.117
HCL GAS	RHONE POULENC	HP August 1976 p.117
HCL GAS	HALOHYDRAL	P.E. October 1974 p.6
HYDROGEN	REFORMING	HP September 1970 p.270
ISOBUTANE	UOP	HP September 1970 p.194
ISOPROPANOL	DEUTSCHE-TEXACO	HP November 1972 p.113
ISOPROPANOL	DEUTSCHE-TEXACO	HP November 1973 p.141
ISOPROPANOL	TOKUYAMA	HP November 1973 p.143
LUBE OILS	IFP	CE February 21st 1972 p.54
LUBE OILS	IFP	HP April 1974 p.129
MALEIC ANHYDRIDE	SAVA	HP September 1971 p.167
MALEIC ANHYDRIDE	SAVA	CE November 17th 1972 p.64
METHANOL	LOW PRESS.LURGI	HP September 1970 p.281
METHYL ETHYL KETONE	HOESCHT-UHDE	HP November 1969 p.169
NITRIC ACID	WEATHERLEY	CE May 23rd 1966 p.116
NITRIC ACID	HOKO	HP November 1966 p.183
NITRIC ACID	BAMAG-WEAK ACID	CEP April 1972 p.68
NITRIC ACID	BONIA	CEP April 1972 p.69

APPENDIX 4 (continued)

DATA REFERENCES.

PRODUCT	PROCESS	REFERENCE
- OLEFINS	MITSUI	HP November 1973 p.102
n - PARAFFINS	ISOSIV	HP November 1969 p.213
n - PARAFFINS	ISOSIV	HP November 1973 p.154
n - PARAFFINS	ISOSIV	HP December 1970 p.77
n - PARAFFINS	BP	HP September 1974 p.204
n - PARAFFINS	HP	HP September 1970 p.274
PHENOL (ACETONE)	UOP	HP March 1976 p.91
PHTHALIC ANHYDRIDE	VON HEYDEN	HP September 1971 p.162
PHTHALIC ANHYDRIDE	VON HEYDEN	HP November 1968 p.162
PHTHALIC ANHYDRIDE	VON HEYDEN	BCE September 1969
PHTHALIC ANHYDRIDE	BASF	HP September 1971 p.162
PHTHALIC ANHYDRIDE	BASF	HP November 1968 p.162
PHTHALIC ANHUDRIDE	BASF	ECN September 22 1967 p.40
PHTHALIC ANHYDRIDE	BASF	HP November 1973 p.59
PHTHALIC ANHYDRIDE	BASF	CE June 1969 p. 80
PHTHALIC ANHUDRIDE	RHONE PROGIL'71	CE March 4 1974 p.82
PHTHALIC ANHYDRIDE	RHONE PROGIL'71	HP February 1975 p.111
STYRENE	ETHYLBENE	CPE December 1967 p.37
STYRENE	PRYOLYSIS	HP Vol 44 (12) 1965 p.137
STYRENE	LUMMUS	HP November 1973 p.180
STYRENE	LUMMUS	CEP December 1967 p.37
TEREPHTHALIC ACID	LUMMUS	HP September 1973 p.209
TEREPHTHALIC ACID	IFP	BCE January 1971 p.11
UREA	STAMICARBON	CPE May 1969 p.81
UREA	STAMICARBON	ECN September 26th 1975 p.40
VINYL CHLORIDE MONOMER	TRANSCAT	CEP October 1973 p.89
VINYL CHLORIDE MONOMER	TRANSCAT	HP November 1973 p.192
VINYL CHLORIDE MONOMER	TRANSCAT	CE June 24th 1974 p.114
p - XYLENE	AROMAX	CE September 17th 1973 p.106
p - XYLENE	ESSO	HP November 1969 p.250
p - XYLENE	ICI	HP August 1969 p.109
p - XYLENE	KRUPP	BCE March 1970 p.301
p - XYLENE	MARUZEN	HP November 1973 p.195
WAX	SFP/BP	HP May 1968 p.177