A CRITICAL STUDY OF
PLASTICS SHEET EXTRUSION PROCESSES

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PREFACE

This report is based on a programme of studies undertaken full time by the candidate over the period October 1960 to February 1964 for admission to Membership of The College of Technologists (M.C.T.). The programme of studies was devised jointly by the Brunel College of Advanced Technology and B.X. Plastics Ltd. and it was officially accepted by The National Council For Technological Awards as being a suitable programme of work for the M.C.T. degree on the 20th February 1961.

The original programme of work on "A Critical Study of Plastics Sheet Extrusion Processes" is given in appendix 1. However, as work progressed it became apparent that the programme of studies was too broad for the candidate to cover in detail, in the time available. Therefore, bearing in mind the facilities available to the student, the college and industrial supervisors directed the candidate to concentrate on the technical factors, rather than the economic and administrative factors, involved in the extrusion of toughened polystyrene (TPS), polyvinyl chloride (PVC) and cellulose acetate (CA) sheet materials on a commercial scale. For the purposes of this report 'sheet' is defined as being equal to or greater than 0.010 inches thick, thinner materials are termed 'film' or 'foil' and as such are not considered to come within the detailed scope of this work.

All practical work was conducted at B.X.Plastics Ltd. because of the limited facilities available to study the extrusion process on a commercial scale at The Brunel College of Advanced Technology; however, close contact was maintained with the college tutor at all stages of this work. Various activities were undertaken by the candidate to obtain background knowledge and information necessary to the programme of studies and these are considered to make up the college aspect of this work. These activities include, a detailed study of over four hundred published papers and books on the science of extrusion; attendance at courses on 'The Physics of High Polymers', 'Inspection and Quality Control' and 'Work Study Appreciation'; attendance at numerous exhibitions, lectures and symposia which were connected with the field of studies; numerous visits to the Process Evaluation Group and the Operational Research Dept. at the Distillers Co. Ltd's Research Station, the Patents Office, and the Board of Trade Library; also visits to or correspondence with numerous companies manufacturing or using extruders. The candidate particularly wishes to acknowledge the hospitality and co-operation given by Battenfeld Ltd. in West Germany, Cascelloid Ltd. and the Distillers Co.Ltd. at various stages of this work. The author also obtained much valuable information on the extrusion of thermo-plastic materials from discussions and correspondence with many leading authorities on this subject, in Britain and overseas.

This report is presented in two major parts, namely 'The Background to Plastics Sheet Materials' and 'The Extrusion Process', each part being subdivided into sections. A summary of anticipated future developments is also included at the end of the report. Abbreviated terms for plastics materials are used wherever possible in this report, taken from a "List of Common Names & Abbreviations for Plastics" B.S. 3502: 1962.

The candidate wishes to thank the Directors of BXP Ltd. for sponsoring this course of studies and allowing publication of this work. The writer also expresses his appreciation to the appointed industrial supervisors, initially Dr.F.C.Lloyd and subsequently and for the major part of this work Dr.J.A.Hetherington, and the college supervisor Mr.G.J.L.Griffin for their assistance and guidance throughout this course of studies. Special thanks are also extended to Mr.S.J.H.Cooke for his help with the writing and continual encouragement and to Mr.J.Jack, Mr.N.G.Perkins and Dr.T.Watson, all of BXP Ltd., for facilities advanced, interest shown and helpful discussions at various stages of this work. The author is also indebted to Mr.P.Chandler (formerly of the Statistics Sect.), Mr.D.R.Reid (Physics Research Dept.), Mr.J.Welham (Engineering Drawing Office) and all other colleagues at BXP Ltd. who have in any way assisted in the completion of this work.

PART 1

THE BACKGROUND TO PLASTICS SHEET MATERIALS

1. INTRODUCTION

The sections presented in Part 1 of the work are primarily a background study of the tonnage growth and market applications of three major plastics sheet materials, namely toughened polystyrene (TPS), polyvinyl chloride (PVC) and cellulose acetate (CA). These sections do not involve practical work conducted by the author, but they provide the essential background information which enables the reader to grasp and understand the commercial importance which the sheet extrusion process has assumed in recent years. This background data shows that there is an ever increasing demand for extruded plastics sheet materials with improved quality characteristics and it therefore indicates why a better understanding of the technical factors involved in the extrusion process is of considerable commercial importance to the plastics industry. A critical study of plastics sheet extrusion processes and the science of extrusion is given in Part 2 of this report, together with an account of all relevant practical work conducted by the candidate.

The major historical developments and the tonnage growth of the above three plastics materials are given in these sections together with anticipated future trends. Past, present and anticipated future applications for these three materials are also discussed and the major extruded sheet applications are emphasized.

There is a natural tendency by polymer producing companies to withhold technical details of production methods which have taken years of research and considerable capital expenditure to develop. Because of this it is impossible to give full details of the manufacture of the above plastics raw materials and the technical developments which have led to new and more economic methods of production.

It is also very difficult to obtain accurate information on the tonnage growth and marketing trends of these materials, particularly in extruded sheet form. Official government statistics are the most reliable source for such data but it will be appreciated that these were introduced and designed in order to give a general picture of the plastics industry as a whole, in connection with government reviews of the national economy, and they do not always provide specialised information concerning small sections of the industry.

Many difficulties arise in the collection and presentation of these data which involve compromises between the ideal and the practicable. Too detailed a breakdown into precise categories would entail much labour in collecting and recording information, quite apart from the serious difficulties of definition and interpretation. Still further problems are the complex and changing patterns of trade, the constant introduction of new products, the need to keep confidential the figures for any given firm, and the competence and conscientiousness of people handling the data. It is not even easy to ensure that information is collected from every firm dealing in a given product or that no duplication occurs. However, although the data given in these sections is subject to these uncertainties it is thought that the trends shown do present a fairly true picture of what is happening in the industry.

Market requirements, control testing and competitive sheet materials are also discussed in these sections and wherever possible price trends are indicated.

2. THE BACKGROUND TO EXTRUDED TOUGHENED POLYSTYRENE SHEET

2.1. Introduction

Extruded TPS sheet materials were developed in the U.S.A. in the early 1950's and they were commercially established before either extruded PVC or extruded CA sheet materials were marketed on a large scale. For this reason, more space is devoted in this report to extruded TPS sheet materials than to either of the other two. To understand the development and growth of extruded TPS sheet it is first necessary to consider the commercial development of untoughened polystyrene (PS) materials.

The major developments and the tonnage growth of PS and TPS materials in both the U.K. and the U.S.A. are given in this section, together with an estimate of the combined tonnage growth of these two plastics materials in many of the principal producing countries of the world. Wherever possible selling price trends in both the U.K. and the U.S.A. are given, and past, present and anticipated future applications and market trends for TPS extruded sheet materials in both countries are stated. Comparisons between the U.K. and the U.S.A. are made because the latter country is the largest producer of TPS materials in the world and the U.K. has in the past tended to follow the market trends and manufacturing methods established in the U.S.A. In particular the refrigerator markets in the U.K. and the U.S.A. are compared because the production of refrigerator door and tank linings constitutes the largest single outlet for extruded TPS sheet materials in both countries. Copolymers such as acrylonitrile - butadiene - styrene (ABS) and styrene - acrylonitrile (SAN) also receive a brief mention.

In this section official government statistics issued by the U.K. Board of Trade, H.M. Customs and Excise, and the U.S.A. Tariff Commission etc. are quoted wherever possible because these are the most reliable sources of information available for production, sales, consumption, import and export data. However, as stated in the introduction to Part 1 of this report, these sources of information suffer from certain limitations and often they are not sufficiently detailed to avoid duplication of data or to give satisfactory data on small specific sections of the industry. There are numerous examples of these limitations. For instance, prior to 1962 some duplication of PS data occurred in the official Board of Trade returns, also no reliable data showing the proportion of PS to TPS grades of plastics materials produced in the U.K. was available until 1962, although the U.S.A. Tariff Commission had classified these two materials separately a number of years previously. Likewise, official data on the quantity of TPS sheet materials produced in the U.K. was not made available until the Board of Trade introduced a new sheet category for toughened and other polystyrene materials in 1962.

In cases where no official data exists some of the figures given in the tables of this section have been estimated with the help of information from trade journals and other sources. It is emphasized that extra care must be taken when using such data which is subject to further uncertainties including journalistic optimism. For instance production data etc. published in the January edition of 'British Plastics' and 'Modern Plastics' for the previous year is based on official government returns for the first nine months of that year, figures for the last three months are less certain, and have to be estimated.

However, in spite of the limitations of the statistics, the data given in this section is believed to represent a fairly true picture of the production and sales trends occurring in the industry.

2.2. The History Of Polystyrene Production

Storax was the first known source of styrene monomer. Storax is a balsam derived from the tree species of Liquidambar orientalis. It has been known since the later Greek period and has had many practical uses over the centuries. Styrene from this source has been identified in resins obtained from embalmed Egyptian mummies dating back 3,000 years.

The earliest published work referring to styrene monomer appeared in 1786 when William Nicholson (1) described how a clear oil was distilled from storax by a chemist named Newman.

This work was repeated by Bonastre (2) who published a paper in 1831 giving details of the production of a colourless oily distillate from American balsam. Out of curiosity, the Dow Chemical Co. repeated the work of Newman and Bonastre a few years ago and proved the presence of styrene in the products.

Gerhardt and Cahours (3) prepared styrene from cinnamic acid. They were the first to record the correct empirical formula of C_8H_8 for styrene in 1841.

A Berlin apothecary, Simon (4), was the first to record the polymerisation of styrene by heat in 1839. He produced a glassy solid but put this down to an oxidation effect, not knowing the importance this product would later assume. This, however, was a beginning and by 1845 Blyth and Hofmann (5) had recorded the precise details of the polymerisation process by heat showing it to be a strongly exothermic reaction. They further showed it was not an oxidation reaction because despite changes in its physical properties styrene appeared to maintain its original composition.

By 1866 Berthelot (6,7) had established that benzene and ethylene react to form a compound which on dehydrogenation gives styrene. He named the product styrolene and derived the following equation

$$C_6H_6 + C_2H_4 = C_8H_8 + H_2$$

He then assumed the formula for styrolene to be $C_6H_4\ C_2H_4$ which although not correct was getting very near the truth.

Styrene gave rise to the first serious attempts to understand the polymerisation process. Kronstein (8) and Stobb (9,10) both attempted to follow the reaction by studying refractive index and viscosity changes. However, little was known of this process until the classical researches of Staudinger dating from 1926.

In its early years polystyrene (PS) was investigated by many leading chemists, namely Berthelot, Kopp, vant Hoff, Glénard and Herzog. The many names which styrene and PS assumed at the hands of these early investigators gave rise to considerable confusion, e.g. storax, styrol, meta-styrol, styrolene, cinnamol, cinnamene and draconyle. Although many of these were later recognised to be the same compound, PS remained a chemical curiosity until after the turn of the 20th century.

It was 1911 before the first signs of PS becoming a commercial venture were seen. In this year F.E. Matthews (11) was granted a British patent in which he described both the catalytic and thermal polymerisation of styrene. The products were stated to be suitable for use in the production of articles then made from hard rubber, celluloid, glass and wood. However at this time no economic methods for the manufacture, storage and transportation of styrene monomer were known.

Moureu and Defraise solved the problem of preventing the polymerisation of styrene monomer during storage and transportation in 1922. They studied autocatalytic reactions and their inhibitions. Numerous materials, mainly aromatic amines and phenols which were capable of preventing the polymerisation of styrene, were discovered by them. An economic method of manufacture of the styrene monomer was tackled by I.G. Farbenindustrie (12,13,14) of Germany and the Naugatuck Chemical Co. of the U.S.A. (15,16,17,18,19,20,21,22,23). A number of patents were granted to the above two companies between 1925-1933 which dealt with both the preparation and polymerisation of styrene monomer.

I.G. Farbenindustrie had produced PS commercially by 1930 under the trade names of "RESOGLAZ" and "TROLITUL". The Naugatuck Chemical Co. marketed PS under the trade name "VITRON" in 1933. The preparation of the styrene monomer was relatively expensive entailing the dehydrochlorination of monochlorinated ethylbenzene. "Vitron" tried to break into the denture market but failed. The product was generally unsuccessful and by 1937 the rights to the Naugatuck patents on PS were transferred to the Carbide and Carbon Chemicals Corporation. However, Ostromislensky had followed up the early work carried out by Berthelot and he is credited with a great deal of work on the preparation of styrene from ethylbenzene by the dehydrogenation process.

By 1930 the Dow Chemical Co. possessed large quantities of benzene and ethylene and were looking for a good commercial outlet for these chemicals, and by 1934 they had established a commercial method for the production of styrene. It was based on the work of Dreisback (24,25,26) who found styrene was produced in good yield by passing vapours of ethylbenzene into steam that had been superheated sufficiently to supply the heat of pyrolysis. Stills were designed which operated at low pressures and enabled inhibitors to be fed into the top thus preventing the monomer polymerising. The Du Pont Co. are also known to have been active in this field about this time.

The properties of styrene polymer can be modified by interpolymerisation with other unsaturated substances not necessarily polymerisable by themselves. Wagner-Jauregg (27) was the first to make use of this technique. He heated styrene and maleic anhydride together and produced a homogeneous mass which differed from styrene polymer by being insoluble in benzene but soluble in alcohol and aqueous alkali. This was the beginning of research which later led to the discovery of styrene acrylonitrile, styrene butadiene acrylonitrile (ABS) and other copolymers.

By 1935 Dow used Berthelots pyrolysis method successfully for the production of 98% pure styrene monomer and they produced polystyrene in 1000 lb. batches under the trade name "STYRON". Large scale production began in 1937. Monsanto and Bakelite were supplying PS based on Dow monomer by 1937. The products were mainly used for specialised high frequency electrical applications. The Koppers Coke Co., and the Catalin Corporation of America very soon realised the commercial importance of the venture and were quick to follow on Dow's heels, all being in production by 1940.

In the U.K. the Distillers Company Ltd. began PS production on a pilot plant scale in 1937 and by 1939 production on a larger scale had commenced. Monomer was made from various raw materials in the U.K. but production costs were high and manufacture ceased in 1943-44. It is interesting to note that A. Boake Roberts & Co. Ltd. (28) were manufacturing styrene-monomer on a small scale by 1934 by dehydrating perfumery grade beta-phenyl ethyl alcohol. Two years later a round-about route from ethyl benzene was devised and by 1939 a more refined method of styrene monomer production from the same source began. The styrene produced was the purest available at that time and was supplied to Standard Telephones. Output during the war was 1-2 tons/week which went to the production of optical and high grade electrical components. A small quantity of very high molecular weight PS was made for the production of thin cast film for condenser dielectrics. The production of both monomer and polymer ceased shortly after the war because production costs were high.

As we have seen the early growth of the PS industry was not confined to the U.S.A. Important developments were also taking place in Germany and a historical introduction would not be complete without a mention of these developments. By 1943 Germany were ahead of the U.S.A. in PS production though they by no means produced such large quantities of synthetic rubber based on styrene and butadiene. In that year Germany produced 7,000 tons (29) of PS compared to the U.S.A. total of 3,000 tons (30), the I.G. Farbenindustrie having been responsible for this achievement.

A full report (31) on the state of the German PS industry during the war is given by a group of specialists and observers who were sent into Germany to assess the situation after the war. After 1943 severe bombing considerably restricted raw material supplies although the monomer and polymer producing plants were hardly damaged. Production ceased in Germany in 1945 and was not allowed to recommence until 1950. The U.S.A. were thus able to take the lead and by 1950 had established PS as a major thermoplastic material. In that year the U.S.A. and Canada together accounted for 86% of the total world PS consumption.

2.3. Growth of PS and TPS Production in Major Countries

There are four commercial methods used for the production of PS from styrene monomer, namely bulk (mass), solution, emulsion and suspension polymerisation. Details of polymerisation processes and comparisons of the physical properties of the polymers manufactured by each of the above four methods are given in various books and technical papers (31, 32) and therefore will not be discussed here.

Styrene monomer is produced by two main industrial methods, the first involves a Freidel-Craft alkylation reaction with benzene and ethylene followed by dehydrogenation, the second entails the oxidation of ethylbenzene to acetophenone followed by reduction to alphaphenylethyl alcohol which is then dehydrated to give styrene. Since the demand for acetophenone for the perfumery industry has been growing the higher styrene monomer production costs by the latter process can be largely offset, also there is no difficult fractionation in the purification process. Outlines for these commercial methods for producing styrene monomer, together with details of raw material and monomer plant costs have been given in European Chemical News (33,34).

The growth of polystyrene consumption between 1952-1962 by the major producing countries is given in fig. 1 , the World figure does not include the U.S.S.R., Eastern Europe or China. Estimated future consumption trends for 1962-1970 are also given.

The present excess world capacity of polystyrene production plants is thought to be approx. 30%, but the world consumption trend shows this could be absorbed by 1965. However, there is little indication that this gap of capacity and consumption is closing, because new polymer plants are springing up at a rapid rate, encouraged by the ready availability of styrene monomer. The collapse of the market price of PS materials over the past six years does not seem to have deterred companies from making large capital investments in new polymer producing plants. Consequently the market for PS materials will remain very competitive and the price could fall still lower, although there seems little hope of a significant reduction in the price of styrene monomer before 1966. Profit margins will thus become even smaller. There is little doubt that the consumption of polystyrene materials in all world markets still has an expanding future. There is also little doubt that present trends will have the following two major effects.

1. Force the smaller general purpose PS manufacturers to retire from this very competitive struggle.
2. Give an incentive to the production of the more profitable specialised PS resins.

WORLD CONSUMPTION TREND FOR POLYSTYRENE

(Excluding the U.S.S.R., Eastern Europe and China).

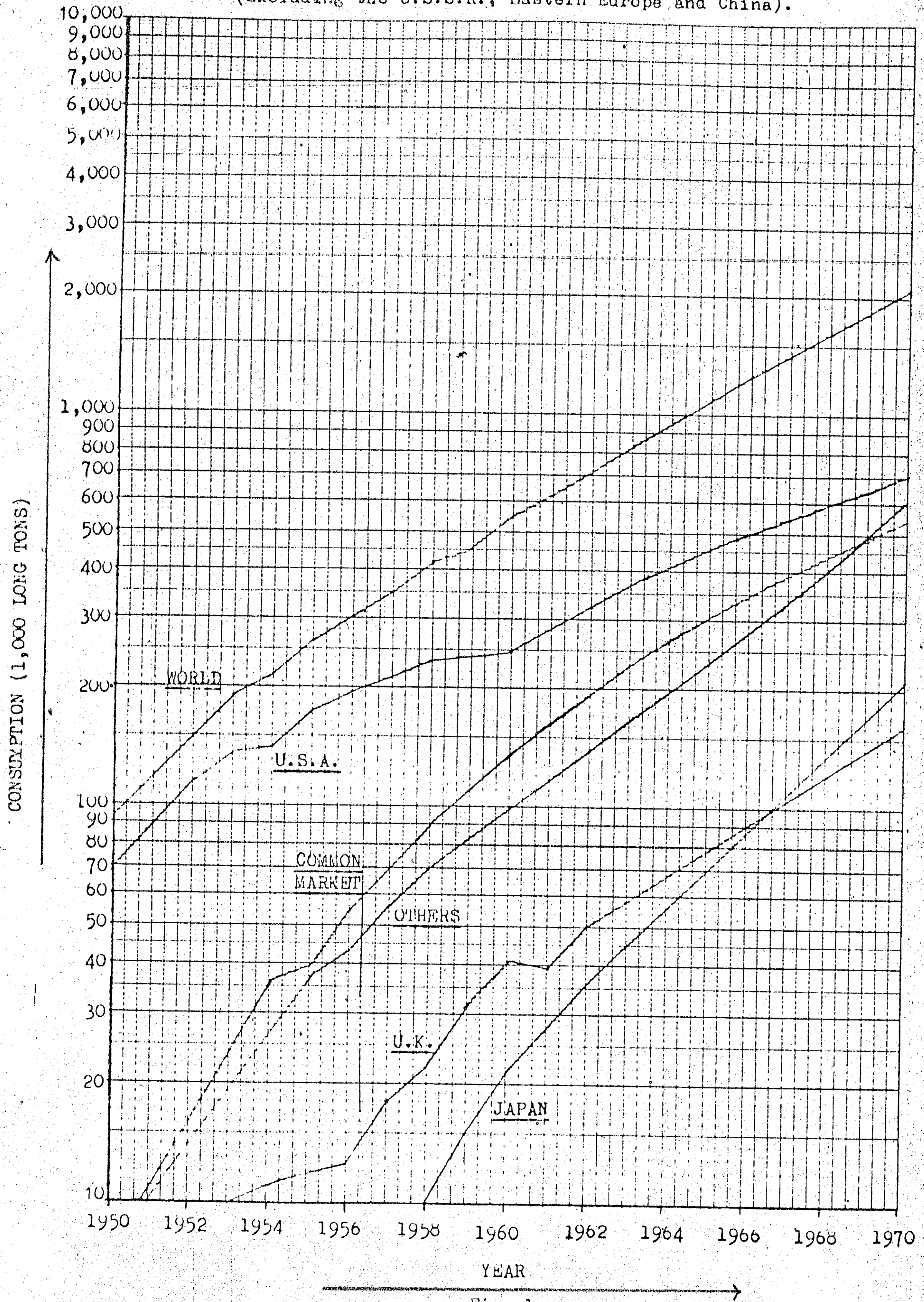

Fig. 1

In 1963 a positive effort to stabilize the price of PS and even raise it is being made in the U.K. and U.S.A. However there is little chance of these prices holding firm unless other European countries raise their prices. In Belgium, Italy and Germany the selling price of G.P. and H.I. grades were still as much as 20% lower than U.S. and U.K. prices by mid 1963.

2.3.1. Production of PS and TPS resins in the U.S.A. and the U.K.

The figures supplied in detail by the U.S. Tariff Commission are for sales, but it is generally thought that these figures with very few exceptions show the production trend of PS over the years. The U.K. Board of Trade figures classify the production of PS materials under the headings "moulding and extrusion compounds" and "others" up to 1961.

2.3.1.1. Production in the U.S.A. (1945-1962)
(all weights are given in 1,000 long tons)

Year	Total Production Capacity	Total Production of PS Materials	Sales of PS for Home Consumption			Exports	Copolymer Sales	Price Cents/lb.	
			Total	General Purpose	High Impact			General Purpose	High Impact
1945	-	10.0	9.8	-	-	-	-	27	-
46	-	33.0	29.9	-	-	-	-	26	-
47	-	47.0	42.0	-	-	-	-	25	-
48	80.0	73.0	56.2	-	-	-	-	27	-
49	90.0	100.0	80.4	68.7	11.7	-	-	26	-
50	112.0	157.0	107.0	91.0	16.0	5.0	-	27	38
51	134.0	160.0	112.0	86.6	25.4	-	-	33	41
52	156.0	167.0	116.0	80.6*	35.4*	-	-	33	41
53	174.0	199.0	134.0	88.9*	45.1*	-	-	32	34
54	201.0	206.0	139.0	81.0	58.0	12.5	-	31	33
55	223.0	256.0	172.0	93.9	78.1	19.6	-	30	32
56	245.0	269.0	194.0	112.0*	81.5*	25.9	14.3	28	32
57	312.0	289.0	209.0	122.0*	87.9*	28.1	17.9	25	29
58	357.0	323.0	234.0	124.0*	110.0*	24.1	21.0	25	29
59	402.0	404.0	240.0	127.0	113.0	40.2	24.6	22	29
60	491.0	438.0	246.0	133.0	112.0	55.8	26.8	18	28
61	536.0	481.0	283.0	140.0	142.0	60.3	32.7	18	28
62	580.0	542.0	324.0	156.0*	167.0*	44.7	40.2	17	27

Table 1.

N.B. 1. * = estimated by author based on trade literature (no official U.S. Tariff Commission figures could be found).

2. Column 2 represents the production capacity of moulding powder and extrusion grades only.

3. Column 3 includes surface coating resins which accounts for some of the figures being higher than those of column 2.
4. Column 4 does not include copolymer sales.
5. Columns 9 & 10 give the market price at the end of each year for 1 ton lots.

2.3.1.2 Production in the U.K. 1950-1962
(all weights are given in 1,000 long tons)

Year	Total Production Capacity	Total Production	Production*		Exports	Imports	Price pence/lb.	
			General Purpose	High Impact			General Purpose	High Impact
1950	10.0	3.0*	3.0	-	-	2.3		
51	15.0	11.0*	11.0	-	-	1.6		
52	18.0	14.0*	14.0	-	-	0.6*		
53	20.0	18.0*	17.1	0.9	10.0	0.5*		
54	22.0	20.9*	19.2	1.7	8.0	0.4	33	33
55	25.0	20.6	18.0	2.6	9.5	0.8	32	32
56	25.0	23.0	18.4	4.6	10.6	0.2	31	32
57	30.0	29.0	19.7	9.3	11.5	0.8	31	32
58	30.0	29.8	19.4	10.4	10.4	1.5	31	32
59	35.0	35.0	21.1	13.9	12.7	4.5	26	29
60	57.0	37.1	18.5	18.6	11.5	10.5	23	29
61	61.0	45.4	22.7	22.7	15.2	3.4	19	24
62	69.0	57.8	28.2	29.6	21.0	4.0	19	24

Table 2.

N.B. 1. * = estimated by author based on trade literature (no official Board of Trade figures exist for columns 4 and 5 up to and including 1961).
2. Column 2 is an estimate of total production capacity.
3. Column 3 is based on "moulding and extrusion compounds" only.
4. Columns 4 and 5 are the author's estimates based on trade journals of the proportion of G.P. to H.I. grades of PS (these are not official Board of Trade figures)
5. Columns 6 and 7 represent exports and imports of all PS materials including those classified under the heading "others".
6. Columns 8 and 9 give the market price at the end of each year for 1 ton lots.

Sources of information for tables 1 & 2 include

1. U.S. Tariff Commission Reports, Synthetic Organic Chemicals U.S. Production & Sales for years 1941 - 1951.
2. January editions of Modern Plastics for years 1945 - 1963.
3. Chemical Economics Handbook II 1950.
4. U.S. Tariff Commission Reports. American Embassy Library, London.
5. U.K. Board of Trade Statistics, Board of Trade Library, London.
6. January editions of British Plastics. 1948 - 1963.
7. World Survey of Plastics 1954-57, United States Department of Commerce. Business & Defence Services Administration.
8. Approx. thirty other government papers, books and company estimates, too numerous to mention in detail.

2.3.1.3. U.S.A. production & trends

(a) Pre 1950

The industrial production of polystyrene had been established by Dow in 1937 and production had reached 300 tons/year by 1939. However large scale production of the polymer did not gain momentum until after the second world war. Shortage of imported natural rubber during the war had forced the U.S.A. government authorities to divert nearly all the available styrene monomer into synthetic rubber production. This made styrene monomer a strictly allocated material in the years 1942-45, and restricted the manufacture of PS to essential war-time requirements.

Styrene monomer capacity expanded rapidly in the war years from 6,700 tons in 1941 to approx. 200,000 tons by 1945. Because of the large government expenditure into research and technical development cheaper methods for the production of styrene monomer were devised. This coupled with the considerable economies of large scale production brought a rapid decrease in the price of styrene monomer. G.R.S. synthetic rubber is a styrene-butadiene copolymer consisting of approx. 25% styrene. It is important to realise how G.R.S. production varied in the years 1941 to 1950 in order to understand the rapid post war growth of PS. Table 3 shows how the natural rubber consumption and G.R.S. production altered over these years.

Year	Natural Rubber Consumption (Long Tons)	G.R.S. Production (Long Tons)
1939	592,000	-
41	775,000	230
45	105,400	719,400
46	277,600	613,400
47	562,700	407,800
48	627,500	393,900
49	574,500	295,200
50	711,000	324,000

Table 3

At the end of the war natural rubber became more freely available and the production of G.R.S. dropped by more than 50% between 1945-49. Ample supplies of cheap styrene monomer became available and all the requirements for a rapid PS growth rate were on hand. Moulders had already had limited experience of handling PS and this together with the natural demand for consumer goods which had been starved during the war and the accumulation of consumer purchasing power gave rise to rapid expansion between 1945 and 1950.

The larger manufacturing companies began to buy back and take over the running of plants which had been under the full control of government authorities. As the demand for PS increased so the price of monomer rose from 7 c/lb. in 1945/6 to 15 c/lb. in 1948/50 and 21 c/lb. by 1952. In spite of these price increases PS remained the cheapest injection material with good moulding properties.

By 1947 PS was being manufactured by four main producers, Dow, Monsanto, Bakelite and Koppers although in 1947 the Monsanto monomer and polymer plants at Texas City were totally destroyed. By 1948 PS had taken over the dominant position, previously held by CA, in the injection moulding field.

It is apparent from table 1 that the rate of growth of total capacity for production of PS moulding and extrusion materials was gradually being overtaken by the demand in the early years after the war. In 1950 there was a sudden boom and demand exceeded supply by approx. 20,000 tons. The main reason for this rapid increase in the sales of PS materials in one year was panic buying because of the Korean invasion. Many sources blame this shortage of PS on a sudden boom in G.R.S. synthetic rubber production programme, but U.S. Tariff Commission figures for that year show that only approx. 7,500 tons of extra styrene monomer were diverted to this use compared with 1949. It is also apparent from table that production capacity was at full stretch in 1950. Thus the production needs of G.R.S. rubber was probably used as an excuse by PS manufacturers to combat the demands made by consumers. The fact that panic buying did take place in 1950 partly accounts for the recession for all plastics materials which came at the end of 1951 and the beginning of 1952. During this time it is thought that consumers were using up their stocks bought in 1950 and the beginning of 1951. It is probably also true that the armaments programme caused shortages of metals for the manufacture of refrigerators, cars, washing machines etc. and this was also partly responsible for the recession. However as the market surveys sponsored by the U.S. Federal Reserve Board indicate, such shortages were satisfied by the end of 1951.

It is not thought that the rate of growth of the PS industry was hampered by any patent situation.

(b) Post 1950

The overproduction coupled with the general recession which hit the U.S. plastics industry in 1950-51 was responsible for the levelling off in sales which took place up to 1954, although it was generally realised that this levelling off would only be temporary. Consequently many companies invested large sums of money to increase existing capacity or build new polymer plants. The sales of PS moulding powder and extrusion compound only rose by 32,000 tons between 1950 and 1954, whereas the production capacity increased by almost 90,000 tons in that same period. 1955 was undoubtedly another breakthrough year for PS, and total sales for general purpose and high impact PS rose by approx. 25%.

The production capacity in 1950 was 112,000 tons, but by 1955 this had doubled to 223,000 tons. By 1960 the capacity had again risen by more than 100% to 491,000 tons. There were twelve major producers of PS in the U.S. by 1959. It is realised that this trend cannot continue indefinitely but at present there is no sign of it slowing down. Production capacity is still increasing at the astonishing rate of approx. 100,000 tons every two years. By the end of 1962 there were approx. twenty PS manufacturers and there existed an excess production capacity of approx. 45%. Although general purpose grades of PS are presently selling at almost negligible profits large petroleum companies such as Shell who have stood on the side lines for so long are anticipating entering this field. It can be presumed that they are building plants for the more profitable specialized grades.

(c) High Impact Grades

High impact grades of PS have become increasingly popular since 1950. In that year they took only 15% of the total market but they now account for over 50%. This proportion of G.P. to H.I. grades has remained fairly constant for the past four years in the U.S.A. and it is puzzling why H.I. grades have not gained more ground over this time. One reason for this could be that the lower grade impact materials have been grouped with the G.P. category in the yearly returns sent by the manufacturers to the U.S. Tariff Commission. On the other hand the cost of G.P. grades is now so low that for certain applications it is very difficult to dislodge. It is generally thought that H.I. grades will level out at approx. 60% of the total moulding powder and extrusion grades by 1965.

(d) Exports

U.S.A. exports were making rapid progress until quite recently. In 1962 exports dropped by 27% compared to the previous year's total. The export market even at non profit making selling prices is breaking down rapidly, mainly because of the price competition between the U.S. and the Common Market countries. The excess world production capacity of PS polymers is no doubt responsible for this worsening situation. Indeed the position in the U.S. is now critical particularly for the smaller specialized companies who started producing between the years 1955-58. The situation is rapidly approaching that of the PVC resin manufacturers (see U.S. Production of PVC).

(e) Selling Price

It is very noticeable that H.I. materials have held their price very much better than G.P. grades since 1955. The question of selling price will be dealt with more fully in section 2.3.1.4.(g).

(f) Copolymer manufacture

Acrylonitrile Butadiene Styrene (ABS) is the most popular copolymer produced in the U.S.A. at the present time. Out of 40,200 tons of copolymer produced in 1962 over 30,000 tons was ABS.

The U.S. Rubber Co. were the first to produce ABS commercially in the late 1940's under the trade name "ROYALITE". Marbon Chemicals started production in the mid 1950's with "CYCOLAC" and B.F. Goodrich followed in 1961 with "ABSON". More recently Monsanto have started producing ABS under the name "LUSTRAN".

Styrene Acrylonitrile (SAN) has been produced since the early 1950's by Union Carbide and Dow, later joined by Monsanto.

2.3.1.4. U.K. production and trends

(a) Pre 1950

German made polystyrene under the trade name "TROLITUL" first became available in 1933. From 1939 to 1950 all polymer apart from a little produced on a pilot plant scale was imported from the U.S.A. From July 1947 import licences were conditional upon 90% of the polymer being used for essential or re-export purposes. In 1950 Monsanto and B.R.P. began large scale polymer production, and in the December of that year a 6,000 ton/year plant at Styrene Products Ltd. began partial operation.

(b) Post 1950

By the end of 1951 a fourth plant erected by O. & M. Kleeman Ltd. commenced operation. All monomer for these plants was imported mainly from Hüls in Germany, which recommenced production in 1950. It was 1953 before monomer production began in the U.K. when Forth Chemicals opened up an 8,000 ton/year plant, and by 1956 this plant had been expanded to 30,000 tons/year.

BXP entered the PS field in 1953 bringing the number of producers up to five, with a total production capacity at that time of approx. 20,000 tons. A breakdown of plant capacity up to the present day is given in table 4 below.

Plant capacity in long tons 1955-1962

Producing Company	Year		
	1955	1960	1962
Styrene Products	6,000		
Kleestron	5,000		
B.R.P. (Distrene - D.C.L.)	7,000	11,000	11,000
BX Plastics	3,000	9,000	11,000
Monsanto	5,000	14,000	17,000
Shell	-	18,000	23,000
Sterling	-	5,000	5,000
Kaylis	-	-	2,000
Total	26,000	57,000	69,000

Table 4.

The Styrene Products Ltd. plant was later taken over by Shell.
The Kleestron Ltd. plant was later shut down.
Mobil hope to open a 7,000 ton/year plant by the end of 1964.

The rate of production growth in the early 1950's was such that demand caught up with supply in 1954. Consequently exports dropped slightly in 1954 and imports almost doubled in 1955. Between 1954-58 production plants were operating at near capacity and the price of G.P. and H.I. grades remained steady. The Board of Trade estimates for total plant production capacity over these years are on the low side because technical development work on many existing plants speeded up the polymerisation process, thus more polymer could be produced than the initial estimates to the Board of Trade suggested. Again in 1959 and early 1960 supply could not keep pace with demand, particularly for H.I. grades, and imports increased sixfold between 1958 and 1960. These imports came mainly from Canada via B.R.P.

The large production capacity extensions carried out by Monsanto and Shell did not come on stream until the second half of 1960, but by then the boom period was over. Between 1958 and 1960 the price of PS dropped from 2/7 per lb. to 1/11 per lb. This was partly due to low priced imports, but the combination of greatly increased capacity and the slowing down of consumer buying in the second half of 1960 were mainly responsible for the fierce competition which took place. In 1961 plant capacity in the U.K. exceeded demand by 15,000 tons and consequently imports dropped from the 1960 figure of 10,500 tons to 3,400 tons.

The time for large scale expansion was in 1954 or 1957. The credit squeeze in 1957 may have been responsible for deterring large scale expansion but it certainly did not arrest the rapid progress of polystyrene materials. That year saw an amazing 45% rise in home consumption and polymer plants were working at capacity over the first half of the year. In spite of the low profit margin it would appear to be the policy of such companies as Shell and Monsanto to carry on rapid expansion in world polystyrene markets and it is mainly from these sources that further expansion in the U.K. can be anticipated towards the end of 1964.

(c) Comparison of rate of expansion in the U.K. & U.S.A.

Both the U.S.A. and U.K. expanded their PS plant capacity by 100% between 1950-54. The U.S.A. again doubled up on the plant capacity in the next five years 1955-59 although production requirements only took up 60% of the available plant capacity. Over this same period production requirements in the U.K. accounted for nearly 100% of the available plant capacity and yet only a 50% rate of expansion took place. Consequently when plant capacity increased by 100% between 1960-62 it was long overdue and came at a time when the market price of PS was falling rapidly.

(d) High Impact Grades

H.I. grades expanded from 12.5% to 50% of the total market between 1955 and 1962. By 1955 H.I. grades in the U.S. had already cornered 46% of the total market. This is a good example of how we often follow the trend established in the U.S. By 1957 U.K. producers were converting plant used for the manufacture of G.P. grades over to H.I. grades. The H.I. market expanded by approx. 27,000 tons between 1955 and 1962 whereas G.P. grades expanded by only 10,000 tons. There will be a tendency in the U.K. market for H.I. grades to level out at approx. 55-60% of the total market by 1965.

(e) Copolymer Production

Large quantities of ABS copolymer have only recently been introduced into the U.K. In 1961 there were only three companies manufacturing ABS in Europe. By 1962 this material was being produced by British Geon (B.R.P.), other plants getting in stream include Monsanto, Marbon Chemicals and Sterling.

(f) Expanded Polystyrene Production

Expanded PS has been produced in the U.K. since 1960 by both Shell and Monsanto. The consumption of this material in 1962 was 3,800 tons. This is only 20% of the West German production which now stands at approx. 20,000 tons. However West Germany have been marketing this product since 1956.

(g) Selling Price

Although U.K. polymer plants are operating at near capacity a further reduction of up to threepence per lb. can be expected for both G.P. and H.I. materials by the end of 1964. There are two main reasons for this

1. The large excess PS production capacity which exists in the world markets and which shows little sign of becoming saturated because of continued expansion.

2. The fact that the U.K. market is wide open to imports of plastic materials.

The excess world capacity situation has been shown previously. The U.K. import situation compared to the U.S.A. and Common Market countries is given below.

(h) Import Duty

The following import duties for the three major plastics materials under consideration are based on the Carriage, Insurance, Freight (C.I.F.) value which includes landed freight and insurance charges. Depending on the country importing plastics materials, extra tax and duty charges may be added to the C.I.F. figure before the import duty is calculated.

Country	Import Duty
U.K.	10%
U.S.A.	2.5 cents/lb. + 12.5%
Belgium	18%
Holland	16%
France	23%
Italy	31.5%
Germany	19%

Table 5.

2.4. Market Growth and Future Trends for TPS Sheet in the U.K. & U.S.A.

2.4.1. U.S.A. production

Rubber modified high impact TPS was introduced in the U.S.A. in approx. 1948. After much research and development into the extrusion of plastic sheet, thin gauge high impact TPS sheet was finally marketed in 1952. At that time sheet up to 40in. wide and 0.040in., 0.060in. or 0.080in. thick was being sold at 55 cents/lb. The major outlet for this sheet was the manufacture of refrigerator door liners by a vacuum forming technique.

The market growth of TPS sheet has been mainly dependent on

1. cost of sheet
2. physical properties of sheet
3. development of extrusion and vacuum forming techniques.

It must be stressed that the development of markets for TPS sheets owes as much to the development of vacuum forming machines as it does to sheet extrusion equipment. In retrospect it is remarkable to observe that neither sheet extrusion nor vacuum forming equipment was developed to any commercial extent before 1950, although sheet forming machinery was patented as early as the 1890's.

By 1955 thicker gauge sheet (0.080 - 0.200in.) was being extruded by two or three companies in the U.S.A. This and the following year proved to be breakthrough years for high impact polystyrene sheet production. Over 90% of the TPS sheet extruded at that time went into the manufacture of domestic refrigerators. One company experimented with extra high impact TPS for luggage and travel cases.

Figures showing the early growth of TPS sheet in the U.S.A. are given in table 6, more recent figures are not available.

Year	TPS Sheet Sold (Long Tons)
1954	9,000
55	13,500
56	22,000
57	26,000
58	28,000

Table 6

2.4.2. U.K. production

In 1956 both Saro and BXP Ltd. imported sheet extrusion equipment from the U.S.A. and began production of TPS sheet, and by the end of 1957 five companies in the U.K. were extruding it for use in refrigerators. It was this rapid expansion in sheet production which was largely responsible for the 100% increase in the manufacture of TPS materials in 1957. Expansion was rapid in the U.K. between the years 1957-60 and over this period approx. 90% of all TPS sheet manufactured was absorbed by the refrigerator industry.

Although the total sales of TPS materials did not expand as anticipated in 1960, the TPS sheet market had another good year. Approx. 5,000 tons of TPS sheet were made, which at that time accounted for almost 30% of the total TPS moulding and extrusion material.

2.4.2.1. Price of TPS sheet

Selling prices for TPS sheet have been falling rapidly over the past few years in both the U.K. and the U.S.A. The price of sheet since its introduction in the U.K. to the present time is given in table 7 below.

Year	Selling Price (pence/lb.)		
	Home Market		Export
	Thin (0.010-0.040in.)	Thick (0.040-0.250in.)	
1956	66	52	-
57	55	48	47
58	63	53	45
59	66	51	36
60	66	48	34
61	51	39	28
62	40	33	27
63	36	28	25

Table 7

It is this downward trend in the market price of sheet which is encouraging the expansion of TPS sheet into many new applications. However the profit margin is now so low that at least one large sheet manufacturer has ceased production in the U.K. and it is very likely that others will also be forced out of the market in the near future. The profitability will no doubt improve at some time in the future but before it does so the price of sheet will probably fall by at least another 4d within the next two years because of fierce competition between sheet producers within the industry, unless there is a sudden boom in the refrigerator market.

2.4.3. Comparison of U.K. and U.S.A. markets

The U.S. Tariff Commission and the U.K. Board of Trade both give approx. figures for the percentage of PS channelled into various outlets. The figures given below are for applications which absorbed large quantities of extruded sheet, but it must be remembered that these figures represent the total PS consumption including moulding materials.

		Percentage Consumption in Each Year								
Country	Application	1954	55	56	57	58	59	60	61	62
U.S.	Refrigeration	20	22	21	20	17	16	14	13	12
U.K.	"	-	-	15	22	23	25	27	20	20
U.S.	Wall Tile	15	14	12	10	8	7	5	4	3
U.K.	"	-	-	-	-	-	-	1	1	1
U.S.	Packaging	13	15	16	17	18	19	20	20	21
U.K.	"	-	-	10	12	14	17	19	21	23
U.S.	Radio & TV	7	7	7	6	5	5	4	4	3
U.K.	"	-	-	10	12	12	13	13	13	8
U.S.	Lighting & Signs	-	-	-	-	-	-	3	3	4
U.K.	Sheet other than refrigeration	-	-	-	-	-	-	7	8	8
U.S.	Miscellaneous	45	42	44	47	52	53	54	56	57
U.K.	"	-	-	65	54	51	45	33	37	40

Table 8.

From the above figures certain trends are very noticeable. The total percentage of PS materials used for refrigeration, wall tile, and radio and TV purposes are seen to be declining, whereas the total percentage used for packaging and other sheet applications is on the increase. Polystyrene wall tiles have rapidly lost ground since 1955 in the U.S.A. to low priced competitive ceramic materials. Each year new applications for PS have been found and thus the total tonnage has been increasing. However with markets such as refrigerators, television and radios the tonnage of PS used varies directly with the yearly production and sale of these items. In the case of television sets, radios and refrigerators, sales in the U.K. and the U.S.A. have either been static or have fallen since 1960, and therefore the percentage of total PS used for these applications has fallen.

The refrigerator industry absorbed over 90% of all TPS sheet produced in the early years in both the U.K. and U.S.A. Today it is still the largest single market for high impact TPS sheet. Because of this the refrigerator market in both countries is presented in greater detail (see section 2.4.3.1.).

2.4.3.1. Refrigerator markets in the U.K. and U.S.A.

The total sales of refrigeration units since TPS sheet was marketed in these two countries is given below.

Country	Sales of Refrigerators in 1,000 Units Year									
	1953	1954	1955	1956	1957	1958	1959	1960	1961	1962
U.S.A.	3,650	3,593	4,015	3,700	3,350	3,117	3,785	3,478	3,480	3,500
U.K.		-	-	450	500	560	930	1,045	875	845

Table 9.

Sales of refrigerators in the U.S.A. have fluctuated slightly but have been fairly constant. On the other hand TPS sheet became available at a time of rapid expansion in the U.K. refrigerator market. The boom years were 1959 and 1960 but they were suddenly followed by a relative slump in 1961 and 1962. It is estimated that the present refrigerator production capacity in the U.K. is 1,750,000 units and therefore this industry is only just ticking over at 50% capacity.

The following table shows how the U.K. lags behind the U.S.A. and other European countries in the ownership of domestic refrigerators.

Country	Percentage of Homes with Refrigerators	
	1957	1961
U.S.A.	90	98
Austria	70	85*
Sweden	50	60
W. Germany	40*	50
Denmark	25	50*
France	12	35*
U.K.	10	25

Table 10

* estimated by author based on trade literature

U.K. producers have so far been unable to tap the 75% of homes which do not possess refrigerators in this country. This is perhaps amazing at first sight when one considers the price drop domestic refrigerators have taken in recent years. Many theories have been put forward to explain the low percentage of homes with refrigerators in the U.K. compared to the U.S.A. and other European countries. It has been claimed that the U.K. economy lags behind the U.S.A. and other European countries. It is also claimed that the American way of life and their hot dry summer climate account for the staggeringly high proportion of homes with refrigerators. A refrigerator is looked upon as a necessity in the U.S. and not a luxury. This may well be so, particularly in view of the fact that 30% of American homes are known to possess deep freeze units. Nearly one million deep freeze units were sold in the U.S. last year whereas in the U.K. less than that number of domestic refrigerators and only 3,000 deep freeze units were sold.

The sale of domestic refrigerators in the U.S. is enormous particularly when one considers that a high percentage are bought as replacement units for existing refrigerators. In the author's opinion the weather is largely to blame for the sorry state in the U.K. market. The summer months of the past three years (1960-1962) have been relatively mild with no extended heat wave, however, one long hot dry summer could again send the sales of domestic refrigerators rocketing skywards, particularly at today's low competitive prices.

(a) Size of refrigerators

In the U.S. market, porcelain-enamelled steel refrigerator door liners had been replaced by paper based phenolic laminated sheet before TPS sheet was introduced in 1952. However the latter had captured most of this market by 1956. Some of the U.S. refrigerator manufacturing companies also made the deep tank liner out of TPS sheet between 1955 and 1958. The average size domestic refrigerator in the U.S. is however 7-8 cu.ft., and the rigidity and strength of TPS after forming was not sufficient to withstand the weight distribution of loaded shelves or suspended freeze compartments in refrigerators of this size and this resulted in many breakages. The breakage rate was particularly high for refrigerator liners during transit to other countries. Consequently manufacturers have returned to porcelain or painted enamelled steel for the larger tank liners and the main application for PS sheet has been confined to the door liners.

In the U.K. the position is somewhat different, the most popular size for a domestic refrigerator is 3-5 cu.ft. The rigidity and strength of a well designed vacuum formed tank liner of this size is sufficient to support a suspended freezer compartment provided the latter is restricted in size, and the weight is evenly distributed by metallic cross members situated above the plastic liner. The strength of the formed edge runners is also sufficient to take the shelf weight of a fully loaded 5 cu.ft. refrigerator. Consequently approx. 75% of the domestic refrigerators manufactured in this country have PS door and tank linings.

(b) Price of refrigerators

The price of a domestic refrigerator in the U.K. has been dropping considerably over the last six years. The table below shows how the price of an average 4 cu.ft. refrigerator has fallen by £20 every 3 years.

Year	Selling Price Range £	Average Selling Price £
1956	80 - 100	90
1959	60 - 80	70
1962	40 - 60	50

Table 11.

The price of some 4 cu.ft. models in May 1963 was as low as £35.

The most popular thickness range for sheet in the refrigerator industry is 0.060 - 0.080in. for door liners and 0.180 - 0.200in. for the deeper tank liners. The refrigerator industry has also used the vacuum forming process for the production of drip trays, crisper trays, egg holding trays, shelves and breaker strips.

(c) Production of refrigerator liners

Plastic refrigerator liners are manufactured by either the sheet extrusion and vacuum forming process or the injection moulding process. The advantages and disadvantages of these two processes will not be discussed but the former process is gaining in popularity over the latter.

There has been a trend in the last few years for refrigerator producers both in the U.S. and the U.K. to equip themselves with vacuum forming machines for the production of liners. This trend will continue because it cuts out the profit made by the custom moulder and reduces the cost of transport and storage. There has also been a tendency in the U.S. for the larger appliance manufacturers to install their own sheet extrusion equipment. About 1960 it was thought that U.K. producers may follow suit, however the present low price of PS extruded sheet hardly makes such a large capital investment worthwhile and no more than one or two refrigerator manufacturers are still seriously contemplating such a move.

It has been the usual practice of vacuum formers to allow a large clamping area during the forming process, this excess material was later removed by either a stamping or sawing process. In order to encourage sales the PS sheet manufacturers have been buying back this edge trim scrap from the larger refrigerator manufacturers. More recently sheet extrusion and vacuum forming techniques have become so refined that there is no edge trim scrap. After forming, the liner is ready to be fitted directly into the refrigerator casing. This requires tolerances on size to within ± 1/32 inch in both the longitudinal and transverse directions, the sheet must be perfectly rectangular in shape otherwise it will not fit correctly into the vacuum forming machine. This latter trend towards trimless formings will become even more popular in the future.

(d) Surface finish

Coating one side of the TPS sheet with a PS film to give the refrigerator liner a high gloss and make it more eye appealing and easier to clean was begun in the U.S.A. in 1954 and in the U.K. in 1957. However foil coated sheet has for some time now been giving way to heat polished and heat glazed materials and this trend will continue mainly because the latter two processes are cheaper. It is conceivable that an extrusion coating technique combining a base PS sheet with a thin film of clear PS may eventually be adapted to compete against heat polished sheet. Methods of producing a gloss on the surface of TPS sheet are discussed in detail in Part II section 11.2.2.

On the question of colour, experience has shown that it is the cold colours which sell. Refrigerators with pink interiors have been marketed but have not sold well and were later withdrawn. The present trend of white, off white, pale blue and pale green interiors will continue.

Printing of refrigerator external surfaces has been tried in the U.S. but did not take and the idea has now been shelved. There is no real scope for printing inside a refrigerator due to the very nature of the complexity of modern door liners which contain several small compartments, shelves, and egg racks etc.

(e) Insulation of refrigerators

The newly developed polyurethane foaming in situ process could greatly increase the use of polystyrene sheet in the refrigerator market. In this process polyurethane foam is injected at approx. 70°C into the space between the liner and outer components of a refrigerator. Not only does this give better insulation properties than the present glass matting but it also imparts greater strength and rigidity to the inner liner. This greater strength will enable the use of TPS sheet in the larger U.S. domestic refrigerators and also in cold storage display cabinets and industrial deep freeze units. This process is already commercially used in the U.S. and four companies in the U.K. are known to be conducting development work along these lines. Only 4% of the total U.S. market has so far been captured. This is partly because many refrigerator manufacturers are reluctant to scrap their equipment for porcelain enamelled liners and partly because they are under the mistaken impression that the more expensive ABS copolymer sheet will have to be used with polyurethane foam to resist freon attack and stress crazing. The latter difficulty has now been overcome by recent development work with both TPS sheet and numerous blowing agents.

Efforts to capture 60% of the interior fridge liner market by 1970 are now being made in the U.S. If this is achieved interest in TPS sheet will be reborn.

2.4.4. Other market outlets for TPS sheet

Because of the seasonal nature of the refrigerator market most TPS sheet manufacturers are trying to break into the packaging market with thin gauge sheet so that production plants can run at capacity all the year round. This objective has already been achieved in the U.S.A., but the big breakthrough has yet to be made in the U.K. It is however notable that since 1960 the Board of Trade has introduced a new category into their yearly statistics, namely "Sheet other than that used for refrigerators".

All other major applications for polystyrene sheet apart from refrigerator liners can be split roughly into three categories

Application	Thickness range of sheet
Advertising & Display	0.010in. - 0.030in.
Lampshade	0.020in. - 0.060in.
Packaging	0.006in. - 0.060in.

Table 12

The market for lampshade material has for some time now been static, it is not thought that this application is likely to expand further. The markets for advertising, display and packaging are expanding rapidly, particularly the latter application. By 1957 over 20% of all PS sheet manufactured in the U.S. went into the packaging industry and by 1962 the figure was nearer 50%.

Some of the applications for TPS sheet include radio housings, the surround for the TV picture tube, and a dust cover over the back of the tube. Other applications include three dimensional signs for advertising purposes, photographic print washers, baffle plates for air conditioners and toilet bowl liners for aircraft.

The use of vacuum formed 3D objects for advertising and display purposes usually requires "offset" or "silk screening" printing techniques. This is often done in a distorted form on flat sheet so that when formed the print assumes the correct dimensions, thus eliminating costly hand painting or spraying methods.

Punching and stamping designs into flat surfaces has been carried out in the U.S. for decorative purposes but this has never developed into a major market.

There has been an increasing demand for PS in the lighting industry and new grades which possess better resistance to UV light have been developed.

A probable breakdown of the main outlets in the U.K. for TPS sheet in 1963 is given below

Refrigeration Industry	55%
Packaging	20%
Advertising & Display	20%
Lampshades	5%

Of these applications, packaging is expected to make the greatest progress in the next few years.

The latest report by the Board of Trade shows that 4,400 tons of TPS sheet were produced in the U.K. in 1962.

2.4.4.1. Packaging applications

TPS sheet lends itself to the constantly changing packaging styles and designs, and possesses many physical properties and decorative advantages over paper, glass and metal. The possibilities for expansion into this field of packaging are almost inexhaustible. TPS sheet will no doubt expand into the fields of foodstuffs, cosmetics, pharmaceuticals, horticulture and cleaning materials. It could be used for packaging cream, butter, cheese, preserves of all types, jellied foods, various meat and fish spreads, cleaning waxes, face creams and ointments.

All types of powders such as custard and baking powder, various spices, tea, coffee, chocolate drink concentrates, sugar, salt, detergents, talcum powder, face powder, medical powders, scouring powders, seed, and pest controllers could be packaged in containers fashioned from TPS sheet.

It has already been used for ready-to-serve dishes, trays, and for packaging biscuits, confectionery, soap, skin plasters, car mending kits, tobacco, film and sound tape reels. TPS sheet has also found use in the production of torch and car battery cases, cigarette cases and packs, disposable drinking containers, containers and thin walled boxes of all types and chocolate packaging. Foodstuffs, pharmaceuticals and pipe tobaccos could be vacuum packed if necessary.

Such specialised uses have led to the development and introduction of specific grades which give improved heat resistance, light stability, increased impact strength, butterfat resistance, better low temperature and anti-static properties.

There was great excitement early in 1963 over a new cigarette pack developed by Phillip Morris in the U.S.A. The pack is vacuum-formed from oriented H.I. PS sheet, using polymer developed by Union Carbide, and Phillip Morris claim to have already invested $.10,000,000 in this project on development and machinery. The pack features sharply defined embossings and printing, it is presently sold in a standard cellophane overwrap but it is thought that this could later be dispensed with. The finished product has an even thickness of approx. 0.015in. The total potential in the U.S. is said to be nearly 180,000 tons and in the U.K. approx. 39,000 tons, if all cigarettes were packed in PS instead of paper and board. This one sheet application therefore could potentially double the total tonnage of TPS materials made in both the U.S. and the U.K. in 1962. While a complete switch to TPS is unlikely, the figures serve to emphasise the impact which a single application could have on the sales of a material, indeed just as the refrigerator business did at an earlier stage.

2.4.5. Market outlets for copolymer sheet materials

ABS has for some time now tried to break into the TPS sheet market, although the rigidity, tensile strength and gloss of ABS are better than TPS the former suffers from the serious disadvantage of being considerably more expensive. For instance the price of ABS in the U.S. was 36 cents/lb. compared to 27 cents/lb. for TPS materials in 1962, however it must be remembered that the required thickness of ABS sheet is often less than that of TPS sheet for a particular application. Typical extruded sheet applications for ABS include sun visors, facia panels and boot lids of cars, portable typewriter cases and other housings and cases for a wide variety of industrial and consumer equipment, safety helmets, luggage cases, camera cases, trays and panelling in public transport.

Last year the price of SAN dropped by almost 15 cents/lb. to 25 cents/lb. However, SAN is a very brittle material and would have to be rubber modified before it could be used as a good sheet material. This virtually means that it would be converted to a type of ABS and the cost would go up proportionately.

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3. THE BACKGROUND TO EXTRUDED POLYVINYL CHLORIDE SHEET

3.1. Introduction

Although the PVC sheet extrusion process has been established in Italy and West Germany for a number of years it was not introduced into the U.K. and the U.S.A. on a commercial scale until approximately 1960.

The major historical developments and the tonnage growth of PVC materials both in the U.K. and the U.S.A. are given in this section, together with an estimate of the tonnage growth of PVC materials in the major producing countries of the world. Wherever possible selling price trends in both the U.K. and the U.S.A. are given, and past, present and anticipated future applications and market trends for PVC extruded sheet materials in both countries are stated. Comparisons are made between the U.K. and the U.S.A. because the latter country is the largest producer of PVC resins in the world, although it is thought that Japan will overtake the U.S.A. within the next few years. The use of rigid sheet for roofing and building components in the building industry is particularly emphasized because this is the largest single outlet for extruded rigid PVC sheet in most countries.

In outline this section follows a similar lay-out to that of section 2, except that greater emphasis is placed on the development and growth of extruded PVC sheet materials in Italy, West Germany and Japan. In the field of rigid PVC sheet extrusion the latter countries are far in advance of the U.K. and the U.S.A. both in terms of the annual tonnage of sheet produced and technical know-how. The sources of information used to obtain production, sales, consumption, import and export data are similar to those described in the introduction to section 2. Therefore, it must be again stated that because of the limitations of official government statistics, many of the figures given in the tables of this section are subject to uncertainty but the trends shown are thought to be a fairly true picture of what is happening in the industry.

3.2. History of Polyvinyl Chloride Production

The polyvinyl chloride (PVC) story began approximately 130 years ago, but a further 100 years passed before PVC was commercially developed.

The discovery of vinyl chloride is credited to Regnault (1) in 1835. He produced vinyl chloride from ethylene dichloride and alcoholic potash and a later paper (2) shows that Regnault was also the first to observe the polymerisation of vinyl chloride in 1838. In this same year he also prepared polyvinylidene chloride.

In 1860 A.W. Hofmann prepared polyvinyl bromide and by 1872 E. Baumann (3) had studied the polymerisation reactions of both vinyl chloride and vinyl bromide in great detail.

Forty years later Ostromislensky (4,5) showed that polyvinyl halides existed in several different forms. He obtained the earliest patent (6) for polymerising both vinyl chloride and vinyl bromide and was the first person to appreciate the potentialities of the polyvinyl halides. Ostromislensky later described experiments on the polymerisation of the vinyl halides by irradiation with a mercury vapour lamp.

In 1922 Plotnikov (7) showed that certain metallic salts assisted in the photopolymerisation of vinyl chloride, he gave the polymer the following formula $(CH_2 = CHCl)_{16}$.

At this stage the discovery and development of polyvinyl acetate must be mentioned because it assumed more importance than PVC in the early days. The final breakthrough in the commercial use of these polymers came with a vinyl chloride acetate copolymer. Monomeric vinyl acetate was first produced in an impure state by Miasnikoff in 1860 but neither its physical properties nor its tendency to polymerise were recorded at that time. Klatte was the first to polymerise vinyl acetate by using benzoyl peroxide as a catalyst. By 1915 Chem Fabrik Griess - Elektron (8,9,10,11,12) had manufactured and marketed polyvinyl acetate. The product however did not sell and was discontinued. Polyvinyl acetate was produced in Canada in 1917 but again failed to find favour. Investigations into methods of production and uses of vinyl acetate were begun in America about this time.

H. Plauson (13) seems to be the first American who was actively associated with the development of PVC.

About 1930 E.W. Reid (14) and L.C. Shriver (15) realised that by copolymerising vinyl chloride and vinyl acetate a much more manageable product was produced. Following this discovery vinyl chloride - acetate copolymer resins were developed by the

Carbide and Carbon Chemicals Corporation and sold under the trade name "VINYLITE". The copolymers were first manufactured by this company as early as 1927 but large scale production did not begin until 1933. The I.G. Farbenindustrie (16,17) in Germany also developed a process for the production of polyvinyl halides of high Mol. Wt. about this time.

Two other American companies were interested in the possibilities of polyvinyl acetate. Shawinigan Chemicals Ltd. began investigations in 1917 and by 1932 were producing vinyl acetate at a rate of 2 tons/day. The Du Pont Co. (18, 19) also marketed polyvinyl acetate and polyvinyl alcohol emulsions. The patent literature is a good guide to the companies responsible for most of the early development of PVC. This clearly shows that I.G. Farbenindustrie in Germany and the Carbide and Carbon Chemical Corp. (later named Union Carbide Chemical Corp.) did the majority of the early development work. Out of 52 Patents granted on this subject between 1930 and 1940, 44 went to the above two companies.

In the early 1930's W.L. Semon (20) worked on the problem of plasticisers for PVC for B.F. Goodrich and Co. The Union Carbide Chemicals Co. (21) and Du Pont (22,23) were also off to an early start in this field. They worked on esters of 2-ethylbutanol and other higher alcohols. This work on plasticisers together with the research and development which went into the production of PVC polymers and copolymers brought about the birth of PVC as a commercial commodity in the early 1930's. From then onwards until the year immediately preceding the second World War the uses of plasticised PVC grew very slowly. In the U.K. PVC compounding was carried out by Cellomold Ltd. by 1937. World War II was responsible for rapidly expanding the PVC industry. Plasticised PVC was an effective substitute for rubber particularly in the application of cable and wire insulation. Rubber machinery was modified to cope with this new plastic material.

The extent to which PVC grew in Germany during the war and the methods by which it was prepared are given in full detail in "German Plastics Practice" (24). Nearly all PVC polymerisation plants were still standing in Germany after the war. Schkopau was by far the largest producer of vinyl chloride monomer with a rated capacity of approx. 40,000 tons/year. Nearly all the monomer and polymer producing plants in Germany at this time differed from one another although they all came under the direction of one man and were sponsored by one company, however, all the polymer plants were based on the emulsion process. 1943 is taken as a peak year for the production of PVC in Germany during the war because after that heavy bombing severely restricted raw material supplies. In that year both Germany and the U.S.A. produced approx. 38,000 tons of polymer. It is thought that at that time Germany had the edge on the U.S.A. as far as the technological understanding of PVC was concerned. However, production in Germany did not recommence after the war until 1950 and by then the U.S.A. were a long way ahead of all other countries.

By 1940 both B.F. Goodrich Chemical Co. (25) and the Carbide & Carbon Chemical Co. were producing PVC in the U.S.A. Dow began producing polyvinylidene chloride resins in the same year. Rigid copolymer sheet was introduced in 1937 and flexible copolymer sheeting followed two years later. By 1939 "VINYON" textile fibre was also being produced. Shortages of plasticisers acted as a brake on the post-war growth of the American PVC industry, particularly in 1946 and 1947.

By 1939 only a pilot plant for PVC manufacture existed in the U.K., but the rubber shortage during the war stimulated production and two factories were built and operated for the Ministry of Supply by I.C.I. Production was 500 tons in 1941 and had risen to 745 tons by 1943. These factories were purchased by I.C.I. after the war. War-time production was augmented by lease-lend supplies of U.S. polymer and war-time uses of the resin were almost wholly in applications in which it could replace rubber, the main one being cable insulation. From 1946 to 1951 the joint rated capacity of these plants is stated as being approx. 12,000 tons. Plasticisers were available after the war but most of them were of poor quality. Consequently articles such as calendered sheeting were sold to the public which were plasticised with unsuitable and often volatile materials. The flexible sheet thus became soft and sticky or rigid depending on climatic conditions, and by 1948 there was widespread public dissatisfaction which affected the sales in the whole of the plastics industry. The boom before 1948 did, however, encourage the buying of new calendering plant specifically designed for handling plastics materials and not rubber. It is interesting to note that although only half as much sheet was made in 1948 compared to 1947, sheet production capacity had increased threefold.

In 1948 British Geon Ltd. began PVC production on a 5,500 ton scale. The capacity of this plant was increased to 8,000 tons by 1950 and 10,700 tons per year by 1951. By 1952 I.C.I. capacity for PVC was approx. 18,000 tons/year. By 1953 PVC was selling at one shilling a lb. cheaper in the U.K. compared with the U.S.A. although monomer costs were approximately the same.

The Germans and Japanese commenced post-war production in 1950. Between 1950 - 1960 there were approx. 200 patents granted on PVC production in Europe, Japan, America and Russia. 28 of these were Japanese. The U.S.A. are still ahead in most of the speciality fields of PVC but Common Market countries and Japan have now taken the lead with rigid PVC applications.

By the end of 1960 approximately 600 patents relating to the use of plasticisers with PVC had been granted in the U.S.A., Canada, U.S.S.R., Japan and European countries.

3.3. Growth of PVC Production in Major Countries

There are four commercial methods used for the production of PVC from vinyl chloride, namely bulk (mass), solution, emulsion and suspension polymerisation. Emulsion and suspension polymerisation, particularly the latter, are the most popular methods of producing PVC resins. Details of polymerisation processes and comparisons of the physical properties of the polymers manufactured by each of the above four methods are given in various books (24, 26) and therefore will not be discussed here.

The monomer vinyl chloride is produced by two main industrial methods, the first involves the interaction of acetylene and hydrogen chloride and the second entails the chlorination of ethylene followed by pyrolysis to vinyl chloride. Outlines of these commercial methods for producing vinyl chloride monomer together with details of raw materials and monomer plant costs have been given in European Chemical News (27,28).

Large scale monomer and polymer production has been developed mainly in the U.S.A., Germany, Japan, France, Italy, and the U.K.

The growth of PVC production between 1950-1962 for these five major producing countries is given in fig. 2, together with estimated future production trends for 1962-1970.

From this graph it is apparent that Japanese PVC production and consumption has grown at a much faster rate than any other country. This trend will continue because Japan intends to double her present PVC production capacity to 750,000 tons/year within the next few years.

The Common Market capacity is likely to treble before 1970 mainly because of huge extensions which are planned for Italy, based on cheap natural gas resources.

Total world PVC consumption has grown from 190,000 tons in 1950 to 1,250,000 tons in 1960. Some estimates place the capacity of all plants in 1960 at 1,450,000 tons, which signified a world surplus of 200,000 tons or approx. 15%.

By 1963 there were approx. 60-70 major producing plants in the world. Present excess world production capacity is estimated at approx. 25% which is only sufficient for 2-3 years increase in consumption at present rates. Unless rapid expansion of PVC pro-

PVC PRODUCTION IN FIVE MAJOR COUNTRIES

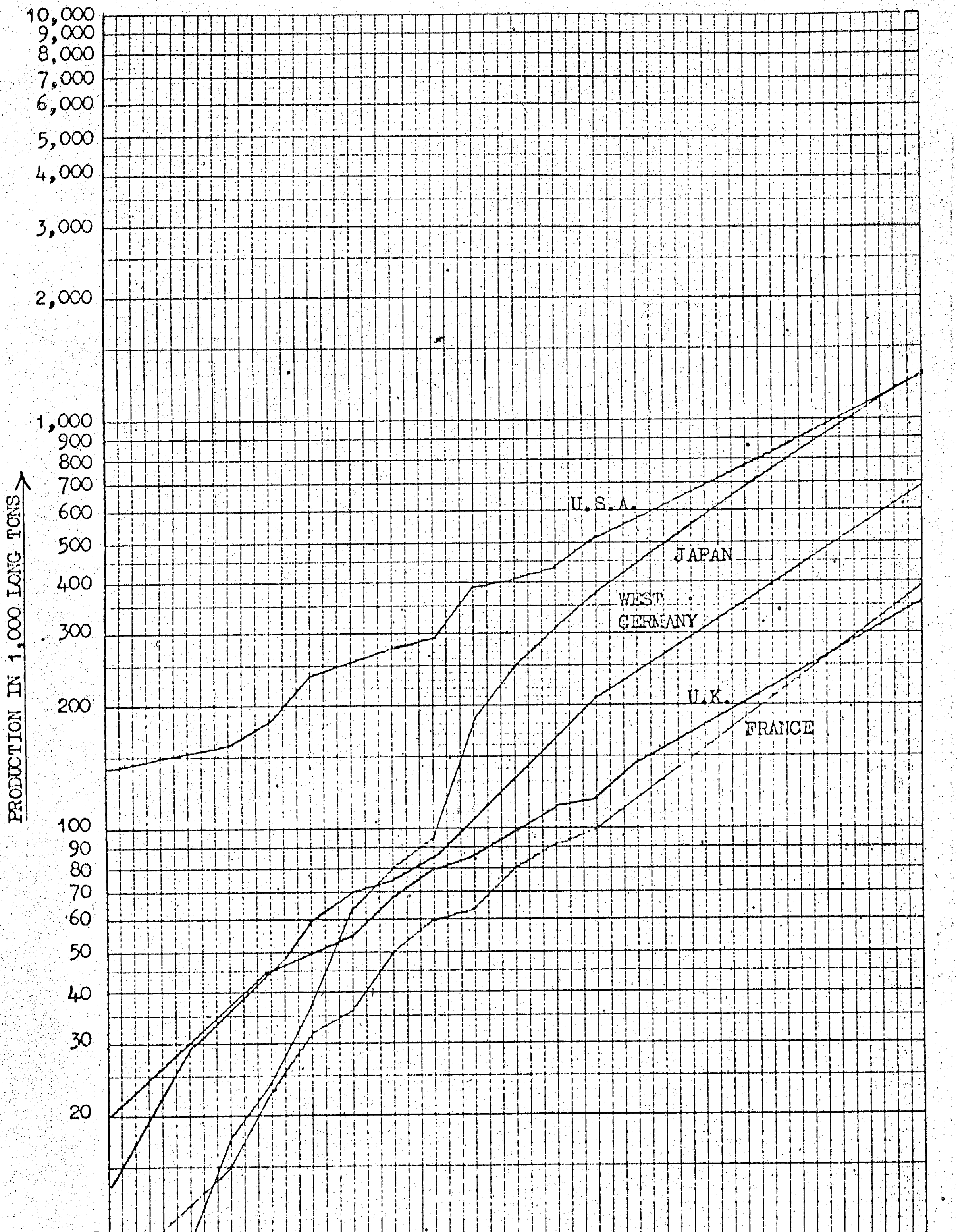

FIG. 2

duction plants is continued there may be a shortage of PVC resins, particularly some speciality resins, within the nexttwo years.

A detailed breakdown of the production of PVC in the U.K. and the U.S.A. is given in section 3.3.1.

3.3.1. Production of PVC and copolymer in the U.S.A. and the U.K.

3.3.1.1. Production in the U.S.A. 1945-1962

Year	PVC Resin Capacity 1,000 Long Tons	PVC Production 1,000 Long Tons	Imports 1,000 Long Tons	Exports 1,000 Long Tons	Price of G.P. Resin Cents /lb.
1945	62.5	54.5	-	-	35
1946	80	69	-	-	33
1947	90	79	-	-	33
1948	110	87	-	-	34
1949	125	89	-	-	35
1950	155	141	-	-	36
1951	200	156	-	-	38
1952	215	153	-	-	38
1953	240	160	-	-	38
1954	270	184	-	14.1	38
1955	290	233	-	15.6	37
1956	310	254	11.2	15.2	32
1957	335	284	6.7	13.4	32
1958	425	290	6.9	6.7	30
1959	450	390	6.3	8.9	24
1960	555	404	6.3	13.4	19
1961	625	431	8.0*	18.0*	18
1962	670	517	10.0*	20.0*	15

Table 13

* = Estimated by author based on trade literature (no official U.S. figures could be found).

3.3.1.2. Production in the U.K. 1952-1962

Year	PVC Resin Capacity 1,000 Long Tons	PVC Production 1,000 Long Tons	Imports 1,000 Long Tons	Exports 1,000 Long Tons	Home Consumption 1,000 Long Tons	Price of G.P. Resin pence/lb
1952	30					
1953	40					21.5
1954	50	45	8.0			20.5
1955	60	50	11.0	10.0	51	20.5
1956	67	55	6.0	10.0	51	20.0
1957	85	68	8.5	10.0	66	20.5
1958	90	80	14.0	10.5	83	18.5
1959	100	86	12.0	13.0	85	16.0
1960	115	98	28.5	16.5	110	16.0
1961	135	112	21.0	17.0	115	13.0
1962	170*	117*	30.0*	20.0*	124*	13.0
1963	205*	145*	38.0*	30.0*	153*	13.0

Table 14

N.B. All prices quoted are for the December in each year.

Sources of information for tables 13 & 14 include

1. U.S. Tariff Commission Reports. American Embassy Library, London.
2. January editions of Modern Plastics for years 1945-1962.
3. U.K. Board of Trade Statistics, Board of Trade Library, London.
4. January editions of British Plastics for years 1952-1962.
5. Approx. twenty other government papers, books and company estimates, too numerous to mention in detail.

3.3.1.3. U.S.A. production and trends

PVC was not available for civilian use until after the war. By the end of 1949 the production capacity for PVC in the U.S. was 125,000 tons. Bakelite and B.F. Goodrich produced 80% of the resin sold in that year. The other 20% was split between the following, Monsanto Chemical Co., Goodyear Tyre and Rubber Co., Firestone Tyre and Rubber Co., and the U.S. Rubber Co.

Panic buying at the start of the Korean war was responsible for the rapid rise in production in 1950. By 1955 there were nine companies manufacturing PVC in the U.S. with a total capacity of 290,000 tons/year. The price of PVC at that time had been stable for the past five years. World production was also expanding and there is known to have been approx. twelve polymer producing plants in other countries by 1955. In this same year approx. fourteen companies in the U.S. were rumoured to be interested in setting up

PVC plants. Many of these companies were PVC converters and intended to consume a major proportion of their own resin. They encouraged bankers to back them and showed how in view of the stable price of PVC they would make a handsome profit Many were successful in raising the necessary capital and began to design and build new polymer plants.

1956 saw an increase in imports to 11,200 tons, these imports came mainly from Japan and Italy and they were so low in price that even with the high import tariff which operates in the U.S. they were able to compete against home produced polymers. The price of PVC resin was reduced by 5 cents/lb. and consequently imports fell in the following year by 4,500 tons. 1957 also saw a new threat to the price of PVC, in that year eight new producers came on stream and by 1958 plant capacity in the U.S. had risen by 90,000 tons to 425,000 tons/year. These new producing companies manufactured more resin than they required and the excess resin was sold on the open market. In 1958 the PVC production capacity exceeded demand by almost 50% and thus the market became highly competitive and the selling price dropped to 30 cents/lb. Because the price was already 7 cents/lb. lower than they had anticipated in 1955 the newer producers decided to expand their plants to take advantage of large volume economies. This of course made competition even fiercer and in the following year, 1959, the price dropped by a further 6 cents/lb. to 24 cents/lb. This was the biggest price drop ever to take place in the history of PVC general purpose resins in the U.S.A.

There has been no sign that the low profit margin and large excess plant capacity have deterred producers from investing large sums of money in expansion programmes and the building of entirely new PVC polymer plants, consequently the rapid expansion which has taken place since 1959 shows no sign of slowing down. By the end of 1962 the total production capacity for PVC resins was 670,000 tons and actual production rose to over 510,000 tons. The cost/lb. of polymer has dropped a further 9 cents/lb. since 1959 and at the end of 1962 stood at 15 cents/lb.

Most producers now claim that profits are so low that they can no longer maintain research facilities. However they also admit that without such facilities they will have to withdraw from the PVC resin market. The manufacturers depending on general purpose resin for calendering and extruded grades are the worst hit. The larger manufacturers with a wide range of products are least worried by over-capacity problems because if necessary they can stand a loss for a short time. All are trying desperately to develop newer and more profitable speciality resins.

3.3.1.4. U.K. production and trends

The PVC position in the U.K. differs appreciably from the U.S.A. The U.K. was a few years behind the U.S.A. and German developments and in the early years relied on imports of PVC from the U.S.A. The production capacity in the U.K. was only 12,000 tons by 1951, this was raised to 60,000 tons by 1955. In that year the Bakelite plant suffered serious damage and did not resume full production until 1957. Production of PVC in the U.K. has been confined mainly to two companies namely I.C.I. and British Geon as the table below shows. Imports were high in 1955 because of the new Coal Board ruling on mine belting, but by 1956 this speciality resin was being produced in the U.K. and imports dropped.

Company	Production Capacity in the U.K. 1,000 Long Tons						
	1955	1956	1957	1960	1961	1962	1963
British Geon	27	27	30	40	50	50	70
I.C.I.	30	40	50	70	80	115	125
Bakelite	5		5	5	5	5	10
Total	62	67	85	115	135	170	205

Table 15

The production capacity of PVC resins in the U.K. just managed to keep ahead of the quantities sold in each year up to 1960, but in that year it was overtaken by consumer demands. Consequently imports of resin doubled between 1959 and 1960. The following two years 1961 and 1962 saw a rapid expansion of production capacity, by the end of 1962 there was a 50% excess capacity based on the quantity of PVC sold, in spite of this imports rose to their highest ever value of 30,000 tons. The reasons for this rise were the importation of speciality resins and the competitive nature of imported resin from Italy, Japan, Norway, U.S.A., Germany, France and the Netherlands. As explained in section we in the U.K. are wide open to imports of plastics materials.

Further expansion plans will come into operation in the U.K. over the next few years in spite of the fact that a large excess production capacity already exists. In this respect we are following the U.S. trend.

Exports have been rising steadily since 1958 and now stand at 20,000 tons which is double the 1958 figure. However the exportation of PVC resins is becoming increasingly difficult because new polymer plants are opening up all over the world and competition is fierce.

It is very noticeable both in the U.K. and the U.S.A. that a sudden rise in sales follows within two years of a price reduction. With each price drop PVC becomes more competitive and often breaks into markets previously held by longer established materials. Occasionally an entirely new application is found which is ideally suited to low priced PVC.

The selling price was nearing rock bottom in 1962 and cannot be expected to fall appreciably lower. Indeed early in 1963 there was a move to stabilize PVC prices in the U.K. by raising the selling price of general purpose resin to 15 pence/lb. If this price holds, PVC will be unable to enter new markets at lower price levels. However, further expansion will come in the many applications in which PVC has gained a foothold but has not yet exploited to the full, particularly in the packaging and building industries.

Therefore, although the profits are low, sales can be expected to carry on rising rapidly until at least 1970. There is however a danger that new polymers and plastics coming on to the market will attack established PVC markets in areas where they are competitive.

3.4. Market Growth and Future Trends for PVC Sheet in the U.K. and the U.S.A.

The market requirements for PVC sheet vary considerably from country to country. Tables 16 and 17 show an approximate breakdown of the quantities of PVC resin used in various sheet applications in the U.K. and the U.S.A. between 1954 - 1962.

3.4.1. U.K. production

All weights are given in 1,000 long tons.

Application		YEAR							
		1955	1956	1957	1958	1959	1960	1961	1962
Unsupported film & sheet (calendered).	Wt.	3.5*	7.7*	14.6	24.8	24	23.4	28	23.4
	%	7	14	21.5	31	28	24	25	20
Supported film (on fabric)	Wt.	8*	7.7*	11.9	8	7.7	16.7	16.8	8.2
	%	16	14	17.5	10	9	17	15	7
Conveyor belting	Wt.	10.5	9.9	8.8	11.2	6.9	5.9	10.1	5.9
	%	21	18	13	14	8	6	9	5
Flooring	Wt.	2.75	3.0	4.1	3.6	2.8	6.4	6.7	14
	%	5.5	5.5	6	4.5	3.2	6.5	6	12
Rigid sheet	Wt.	1.75	2.2	3.1	2.2	3.1	3.6	3.4	4.8
	%	3.5	4	4.5	2.75	3.6	3.7	3.0	4.0
Rigid & Flex. Extrusions	Wt.	5.5	6.6	9.2	7.4	10.8	10.6	17.8	16.3
	%	11	12	13.5	9.25	12.6	10.8	16	14
Miscellaneous	Wt.	18	17.9	16.3	22.8	30.7	31.4	29.2	44.4
	%	36	32.5	24	28.5	35.6	32	26	38
Total	Wt.	50	55	68	80	86	98	112	117

Table 16

* = Estimated by author based on trade literature.

3.4.2. U.S.A. production

All weights are given in 1,000 long tons

Application		YEAR								
		1954	1955	1956	1957	1958	1959	1960	1961	1962
Film under 0.010 in.	Wt.	30.8	37.1	34.8	34.8	33.0	40.2	43.8	40.2	38.0
	%	16.7	15.9	13.7	12.3	11.4	10.3	10.8	9.3	7.4
Sheet over 0.010 in.	Wt.	24.6	22.8	23.7	35.7	33.0	42.4	50.0	65.6	73.7
	%	13.4	9.8	9.3	12.6	11.4	10.9	12.3	15.2	14.3
Moulding & Extrusion cmpd.	Wt.	66.1	81.7	91.5	98.2	100.0	136.7	134.0	138.5	179.9
	%	35.9	35.0	36.1	34.6	34.5	35.0	33.1	32.1	34.8
Flooring	Wt.	15.2	25.0	29.5	36.6	51.8	69.2	71.4	82.6	102.7
	%	8.2	10.7	11.6	12.9	17.9	17.7	17.9	19.2	19.9
Miscellaneous	Wt.	47.5	66.8	74.3	78.2	71.9	101.7	104.4	104.4	122.7
	%	25.8	28.6	29.3	27.6	24.8	26.1	25.9	24.2	23.8
Total	Wt.	184.2	233.4	253.8	283.5	289.7	390.2	403.6	431.3	517.0

Table 17

3.4.3. Comparison of U.K. and U.S.A. markets.

The above two tables show that the thin sheet market in the U.K. and U.S.A. is fairly static but the thick sheet and flooring markets are expanding rapidly. Conveyor belting in the U.K. has been loosing ground since 1955.

By 1950 there were between 50 - 60 PVC sheet calendering units in operation in the U.S.A. The main uses of PVC in the early years in both the U.K. and U.S.A. were rainwear, upholstery in the car trade, handbags, book binding, luggage covering, wall covering, shower curtains etc. In the U.K. as much as 70% of all calendered film went for rainwear in 1953. About this time rigid PVC sheet made by the calendering and pressing technique was developing considerable interest. Early applications for this sheet were surrounds for T.V. sets, refrigerator components, lighting fittings, aircraft window surrounds, trays for food stores and linings for chemical tanks and industrial chemical equipment.

The quantity used for conveyor belting in the U.K., particularly mine belting, dropped considerably in 1957 because thinner coatings were being applied to the core of the belt.

By 1958 there were approx. 20 calendering units in the U.K. Many of the newer calendering units installed since 1958 incorporate adjustable cross axis and roll bending refinements which enable the production of better quality PVC film and sheet.

Flexible extruded PVC sheet was manufactured and used in the U.S.A. as gun covers in the last world war. After the war interest in this process faded out in favour of the faster calendered sheet process. Little progress was made in the extrusion of rigid vinyls in the early 1950's. Techniques and knowledge were building up slowly during this time, more slowly than many people anticipated. Extruded rigid PVC finally began to break into the U.S. market about 1960 but it was 1962 before any real progress was made in the U.K. Many people have predicted a rapid expansion for extruded rigid PVC sheet in the late 1960's. Some prophesy a 500% expansion within the next six years similar to that which hit the PVC flooring in 1953-59 after a late start.

It is thought that the greatest immediate growth potential in the U.S.A. and the U.K. for extruded rigid PVC sheet is the production of roofing etc, for the building industry. This could well be true because the U.S.A. and U.K. are very backward in this field compared with W.Germany, Japan and Italy, as the following figures presented by D.G.Owens (29) in 1961 indicate.

<u>Comparison of Usage by Application in 1960</u>
<u>(lb/head of population).</u>

Country	Application								
	Calendering	Cables	Coated Fabrics	Flexible Extrusions	Records	Flooring	Belting	Rigid Application	Total
W.Germany	2.3	1.7	0.4	0.7	-	0.4	-	0.9	6.3
Japan	1.3	0.6	0.4	0.4	-	0.1	-	2.5	5.5
U.K.	1.2	0.8	0.5	0.5	0.2	0.3	0.3	0.2	4.7
U.S.A.	1.2	0.5	0.4	0.4	0.3	0.9	-	0.1	4.6
France	1.2	0.6	0.3	0.3	0.1	0.4	-	0.3	4.0
Italy	0.6	0.5	0.1	0.7	0.1	-	-	0.9	2.8

Table 18

Because the building industry is the largest single outlet for rigid PVC extruded sheet, this application will be presented in greater detail (see section 3.4.3.1.).

3.4.3.1. Rigid PVC sheet for the building industry

The quantity of PVC consumed by the building trade in Holland, Italy and Japan was 37%, 28% and 22% respectively in 1960, in the same year only 4.5% of the total PVC consumed in the U.K. went into the building industry. However there are now signs that the U.K. and the U.S.A. are determined to make up this lost ground. In the latter country rigid PVC pipes, conduits, tubing, corrugated sheet, window frames and other items accounted for approx. 22,300 tons in 1962, by comparison approx. 140,000 tons of similar materials were consumed in Japan in this same year.

(a) U.K. and U.S.A. producers of extruded sheet

Extruded rigid PVC sheet in the U.K. and U.S.A. has so far been mainly translucent and opaque and it has been sold in a corrugated form. There are now five major producers of rigid PVC corrugated sheet in the U.S.A. The Navaco Co. was the first company in production with a process developed jointly by the National Rubber Machinery Co. and B.F.Goodrich. The former supplied the extruder and die and the latter developed the PVC compound. Within the last two years Allied Chemicals, Monsanto and Johns - Manville have entered this field.

In the U.K. B.X.Plastics Ltd., commenced production in 1961 with Italian extrusion equipment and polymer. Marley and Allied-Structural-Plastics Ltd. have since followed suit.

(b) Needs of the building industry

It has been estimated that an output of 7,150 tons (approx. 33 million square feet) per year will be required if PVC corrugated sheeting is to capture 20% of the U.S. market by 1965. Eight extruders operating at 400 lbs/hr. on a 20 hr./day, five days a week basis,for 50 weeks in the year, would satisfy this demand. A large number of such units in a few countries may therefore cause severe over production, this in turn would result in a rapid fall of the selling price.

(c) Advantages of rigid PVC sheet

The advantages of this sheet are light transmission, resistance to fire spreading and low maintenance costs. The present product will withstand extremely low temperatures over a long period but will not withstand extreme heat.

The history of this product in the U.K. and U.S.A. is too short as yet to establish the resistance of PVC corrugated sheet to strong sunlight although reports based on tests made over the past ten years in Italy and Germany are good.

PVC corrugated sheet can either supplement or compliment the established steel, aluminium and asbestos products and other building materials. It was introduced to offer the consumer a strong light weight material of pleasing design at a relatively low cost. It is used mainly in the construction of non-supporting walls, roofs, patios and various industrial enclosures.

(d) Size and profile shapes of sheet

The sheeting in the U.S.A. is offered in thicknesses varying from 0.047 - 0.060 inches but over 0.055 in. is usually specified unless extra framing is used. In Europe the sheet is marketed in 24 - 54 in. widths in thicknesses varying between 0.050 - 0.070 inches.

The sheet is manufactured in a large variety of colours and can be either transparent, translucent or opaque. It is mainly marketed in corrugated form but could be flat or embossed depending on the design of manufacturing plant used. Corrugated sheet is usually made in one of two basic shapes, sinusoidal or trapezoidal. The pitch of corrugation can be one of many different sizes to match up to builders requirements. The two main types being offered in the U.S.A. are both sinusoidal having a 2.67 in. pitch x 9/16 in. depth and a 1.25 in. pitch x ¼ in. depth. In the U.K. the two most popular sizes match up to 3 in. and 6 in. asbestos sheets.

Plant is available in the main producing countries for the manufacture of either longitudinal or transversely corrugated sheet. The former must be transported in sections of sheet form and this can raise transport problems when only small quantities of large sheet are required. On the other hand the transversely corrugated sheet can be transported in roll form. Other roofing accessories e.g. roof cappings and gutterings are also made from extruded PVC sheet, by heat forming techniques, to match up with the roofing sheet and thus complete a roof structure.

(e) Selling price

The price for 0.060 in. thick PVC corrugated sheet in the U.S. is approx. 55 cents per lb. and in the U.K. 3/6d per lb.

3.4.4. Other market outlets for extruded PVC sheet

(a) Rigid sheet

Extruded PVC sheets in thicknesses varying from 0.010 - 0.040 in. could be used for such applications as illuminated letters and advertising signs, other signs, transparent containers for packaging and display covers, wall covering, flooring, toys menu covers, office accessories, switch plates, mirror backs, battery cases, collar stiffeners, transparent covers for ice boxes, industrial tapes, developing dishes, etc.

Hundreds of tons of German film are being imported into the U.S.A. each year for use as magnetic tapes. Heavier gauge sheet from Germany is also imported in large quantities for the production of artificial Christmas trees.

Use of rigid PVC sheet in windows and doors is growing after years of effort to eliminate contraction and expansion problems, using improved resins and processing techniques. Production of high clarity thin PVC sheet in the U.S.A. is being held up slightly because the F.D.A. have not yet approved its use as a packaging material for foodstuffs.

Unplasticised thinner gauge 0.010 - 0.040 in. extruded PVC sheet is still a small factor in the U.K. and U.S.A. although it is widely used in Italy, Germany and Japan. The latter country in particular is now marketing excellent quality high clarity thin extruded PVC sheet. There is a large potential market for this type of sheet particularly in food packaging e.g. butter, margarine, lard, baking ingredients, spices, paste spreads of all kinds, powders of all kinds, jams, salads, fish, mayonnaise, sauces, housewares and disposable camping containers, liquid drinks (cold or hot) etc.

Large quantities of 0.010 in. extruded PVC sheet are now being made into heavy duty bags for the packaging of fertilisers etc. in the U.K. and U.S.A.

(b) Flexible sheet

Extruded plasticised sheet was an important growth area in the U.S.A. in 1962. Much development work on this process is being carried out both in the U.S.A. and Germany. It is now being used extensively for upholstery in furniture, cars and boats. Extruded flexible 0.060 in. thick PVC sheet is being used for the production of car floor mats, flooring, inflatables and handbags.

Present manufacturing techniques for the extrusion of PVC sheet is from the more expensive compounded material, rather than powder blends.

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4. THE BACKGROUND TO EXTRUDED CELLULOSE ACETATE SHEET

4.1. Introduction

Cellulose acetate sheet materials have been extruded, without the aid of solvents, on a commercial scale in the U.S.A. since the early 1950's but this process was not introduced into Europe until 1962. Prior to this date, most of the CA sheet manufactured in Europe was made by the long established CN block pressing and slicing technique, and because of this both CA and CN materials will be discussed in this section.

The major historical developments of CA sheet materials and the tonnage growth of CA and CN materials in the U.K. and the U.S.A. are given, and wherever possible selling price trends are stated. Past, present and anticipated future applications and market trends for CA sheet materials in both countries are also given and the importance of extruded CA sheet in the spectacle frame market is particularly emphasized.

CA is of secondary importance in todays plastics industry because of the growth in relative importance of some of the more recently developed plastics materials such as TPS and PVC. The publicity given to CA has declined in recent years although the annual tonnage of CA materials produced in the U.K. and the U.S.A. has been gradually increasing. Information on world production trends for CA materials is not readily available and, unlike the previous two sections, such data is not given here. However, comparisons between the U.K. and the U.S.A. are made because the latter country is a major producer of extruded CA sheet materials and similar methods of sheet production have recently been adopted in the U.K.

Official government statistics and other sources of information used in this section suffer from similar limitations to those described in the introduction to section 2. For instance in the U.K. Board of Trade statistics the category for CN sheet, rod and tube materials has been grouped with CA data since 1957. However the data presented in the tables of this section is thought to give a fairly true picture of the production and sales trends occurring in the CA industry.

4.2. History of Cellulose Acetate Production

A full account of the early stage by stage development details of the production of cellulose acetate (CA) is given in a work produced by E.C. Worden (1). Here it is intended to give a brief account of only the most important developments which led to the present day manufacture and uses for acetate sheet.

The very early work on the experimental production of CA is not recorded in detail. The earliest reference to its manufacture appeared in a paper by P. Schutzenberger (2) in 1865. This gives a description of heating cotton and acetic anhydride in a sealed tube at 180°C, CA was precipitated by the addition of water. Schutzenberger's description of the physical and chemical properties of the purified dried product later proved to be remarkably accurate as is often the case with these early chemists.

The next milestone in the preparation of CA came in 1879 when Franchimont (3) used sulphuric acid as a catalyst. The first British patent relating to the production of CA was granted in 1894 to Cross and Bevan (4).. This described a method of production using zinc acetate and acetyl chloride. They claimed that the CA prepared by this method could be used for the production of films, and as a replacement for the then popular collodion varnish. However, it was the method of preparation described by Franchimont which proved to be the most successful. Around 1900 both Lederer (5) and Eichengrün (6) were each granted British patents which involved the use of sulphuric acid as a catalyst.

There is little doubt that by the end of the last century at least two German companies namely the Bayer Co., and Knoll & Co., were interested in CA production but did not think the product was worth marketing. This is understandable because all the patents granted up to that time described the preparation of CA at high temperatures which we now know give a degraded product of poor strength. The cellulose nitrate (CN) industry was already established and thus the essential raw material, cotton linters, was commercially available, but the time was not yet ripe for the commercial production of CA. This point was most unfortunately proved by Fürst Henckel Von Donnersmark who bought up the patent rights and developed a process at Strettin. The CA produced was marketed under the trade name of "CELLESTRON". The doubts of the Bayer and Knoll Co.'s seem to have had foundation because "CELLESTRON" had little practical value and although no production or sales data is available it is known that this venture soon failed.

By this time a ready market existed for anyone who could produce a clearer non-inflammable product to compete against, and expand the market created by CN. Both the Americans and Germans worked hard to this end with some success. Between 1903 and 1905 Miles (7,8,9) in America prepared CA with improved solubility characteristics. This process involved a partial hydrolysis of the cotton linters using dilute sulphuric acid to produce an acetone soluble product. Prior to this CA had been insoluble in all common solvents of a non-toxic nature. Meanwhile similar work using weak acids was being carried out by Eichengrün, Becker and Guntrun for the Bayer Co. By 1905 the latter company felt that the quality of product was sufficiently acceptable to market under the trade name "CELLIT". This material succeeded and the Bayer Co. soon expanded their plant, the main outlets being acetate lacquers and cinematograph films.

Bayer, together with three other companies, namely,

Soc: Usines de Rhones	at Lyons
Cellonite Co.	at Basel
Cellulose Products Co.	at Boston, Mass.

are thought to be the only companies producing CA by the outbreak of World War I.

With the 1914-18 war came the increased need for dopes for aeroplane fabrics and the U.K. speeded up its research into this field. The British Government secured the help of two Swiss chemists Drs. D. & H. Dreyfus, who had already manufactured this material on a small scale at Basel between 1910 - 1913, to form the British Cellulose and Chemical Manufacturing Co. Ltd., in 1916.

During and just after the first world war some CA was converted into sheet by the established CN method of block pressing and slicing. A full account of the latter method of sheet production is given by P.S. Adamson (10). Unfortunately, camphor was not compatible with CA and although many plasticisers were tried, none possessed retention properties which compared to the camphor-nitrocellulose partnership. The CA sheets produced weathered and warped badly and did not possess the mechanical rigidity of CN sheet.

When hostilities ceased in 1918, CA was available in large quantities and the industry naturally tried to find other outlets. The earliest efforts were directed to produce a cellulose derivative which would possess the physical properties of CN, and yet overcome the yellowing effect produced on exposure to sunlight, and its inflammability. No real success was achieved and so the British Cellulose and Chemical Manufacturing Co. Ltd. turned its attention to the commercial manufacture of solvent spun artificial silk which they marketed under the trade name "CELANESE". The wonderful success they had in this field is now common knowledge. The company was later renamed British Celanese Ltd., and as a subsidiary of Courtalds still produces acetate rayon.

About 1919 Eichengrün (11,12) in Germany adapted the die-casting methods used for the manufacture of small brass and zinc parts, to the injection moulding of CA granules. Eichengrün succeeded in injection moulding a few small samples. This method became more popular in the late 1920's when better plasticisers appeared on the market, and new hot compounding methods of processing CA were evolved which no longer required the use of solvents.

The injection moulding process began slowly in both the U.K. and the U.S.A. about 1930 and was becoming established in the latter country by 1933. CA and the injection moulding machine were initially each responsible for the success of the other. The main economic reasons for their rapid growth in the U.S.A. were:-

1. CA was becoming a cheaper commodity because of the expanding acetate rayon industry.
2. Legal restrictions were imposed relating to the buying and selling of CN.
3. The lower cost moulding process enabled the home industry to compete against imported Japanese CN products.

When the original Eichengrün patents finally expired in 1936 the injection moulding industry received a fresh impetus and developed rapidly. Today the process is established as the main method of mass-producing thermoplastic articles, and faster and larger machines are continually being introduced.

The interest shown by the plastics industry towards CA in the late 1920's is reflected in a series of technical articles by P.Chaumeton and V.E.Yarsley which appeared in 1929 during the first year of publication of the first U.K. plastics journal "British Plastics". A British Standard Specification on CA was also produced about this time.

The 1930's brought the commercial production of cellulose ethers and mixed esters, particularly in Germany and the U.S.A. Cellulose acetate butyrate was marketed by the Hercules Powder Co. in the U.S. by 1932. Ethyl cellulose and cellulose acetate propionate followed in 1935 and 1937 respectively.

By 1940 the plasticisers most commonly used with cellulose acetate were methyl, ethyl, butyl and methoxyethyl phthalates, triphenyl phosphate, triacetin, tripropionin; and methyl phthalyl ethyl glycolate.

Cellulose plastics were in steady demand up to and including the 1939-45 world war. Large quantities of CN were being produced in the U.S.A., U.K., France, Germany and Japan. The two major producers of CA sheet and moulding powder at that time were the U.S.A. and Germany. The latter country had four major producers of acetylated cellulose namely I.G. Farbenindustrie, Rhodiaceta A.G., Wacker A.G., and Lonza A.G. These companies had a total production capacity of approx. 15,000 tons/year. The acetylation process was different in each case and the acetic acid used was prepared from acetylene which in turn was prepared from carbide. Venditor, an associate of Dynamit, were the only company producing CA moulding powder in Germany during the war. They produced approx. 120 tons of this material in 1944 which was well below the capacity figure of approx. 550 tons. Only three formulas were made, containing 15-30% dimethyl phthalate as plasticiser and were sold under the name "TROLIT".

Venditor had conventional plant for making CA sheet by the established CN method, total capacity for nitrate and acetate sheets being in the region of 2,400 tons/year. The largest sheets available were 80in. x 40in. and the majority of sheet sold was between 0.005in. and 0.500in. thick.

The main applications for CA sheet in Germany at that time were reinforced window glazing and D.D.T. and sulpha drug containers. Moulding powder was used in the production of film spools and telephone parts. The film casting process was also carried out on a large scale, the film being mainly used for photographic purposes and for insulating and recording tapes.

Since the mid 1940's severe competition from wholly synthetic polymers such as PVC, PS and polyethylene (PE) has pushed CA further and further into the background. However, more recently there has been an increasing demand for CA sheet for specialised applications such as spectacle frames.

The extrusion of CA sheet dates from 1927. In that year the Fiberloid Corp. in the U.S.A. developed a solvent extrusion process for the production of continuous sheeting. This found a large outlet in the growing car industry as a safety glass interlayer under the trade name "FIBESTOS". Later the Monsanto Chemical Corp. took over this company. The sheet was biaxially orientated by a stretching device to give it greater strength. However, this application for CA was relatively short lived, mainly because of brittleness at low temperatures, poor moisture resistance, and unsatisfactory adhesion, and by 1939 it was replaced by polyvinyl butyral.

The extrusion of CA in granule form did not begin until approx. 1932 and it was 1939 before it became a commercial proposition. The second world war quickened up the rate of progress in this field. Up to this time all plastics extrusion was conducted by ram type extruders. Heidrich in Germany was the first to build a screw extrusion machine specially designed for extruding dry plastic materials in 1931. By 1938 the Detroit Macoid Corp. in the U.S. were experimenting with this type of machine, as were the National Rubber Machinery Company. They adapted the already established rubber extruders for the specific purpose of extruding thermoplastic materials by lengthening the barrel and screw and increasing the number of heating elements.

The first CA sheet to be produced on a screw type extruder was in tube form. The tube was slit lengthwise as it left the die and the curved sheeting was passed over a series of spreader-bars of gradually increasing radius of curvature and then through a pair of driven pull down rolls. Kodak in the U.S.A. used the latter process.

The film casting process later assumed much more importance and developed into a major industry. The clarity and surface of thin gauge CA sheet produced by this continuous band process is still superior to the quality which can be obtained from the direct extrusion of dry moulding powder. The film casting process is particularly used for producing CTA film which is used in large quantities in the photographic and movie film industries and also for x-ray plates.

The second world war greatly increased the uses of CA sheet, including aircraft windscreens and cockpit covers, fairings, wing tips, pilot seats and storage boxes for dinghies. Other uses were eyepieces for civilian gas masks and safety goggles.

Use of CA sheet as an interlayer in safety glass has already been mentioned. A large quantity of thin film was made into an "unbreakable" window covering by laminating it to a fabric gauze. A thicker variety of "unbreakable glass" consisted of wire gauze laminated between two sheets of acetate. This was marketed as glazing material for greenhouses and garden frames and had the advantage over glass of being transparent to ultra violet light. Acetate sheet was also laminated to paper, wood and metal foil for a variety of applications. Decorative effects were often printed on the sheet surface. CA sheet was often used for the production of toys, spiral hinges for diaries and booklets, collar stiffeners, lampshades, and for the preservation of important documents.

Thin acetate sheeting was used as an electrical insulation and also as a heat insulator. The latter was manufactured by building up alternate layers of corrugated and flat thin CA sheets to an approximate thickness of one inch. This was sold under the trade name "ISOFLEX" and was used as a lining in cold storage rooms and railway refrigerator cars.

4.3. The Production and Applications of Cellulose Acetate Sheeting in the U.K. & U.S.A.

Because cellulose acetate (CA) is of secondary importance in today's plastics industry compared to polystyrene (PS) and polyvinyl chloride (PVC) materials, the commercial production of CA will be given in much less detail.

The PS and PVC industries continue to expand rapidly and new applications for these materials are continuously being found. The tonnage of CA produced on the other hand remains broadly constant though it fluctuates somewhat year by year. The overall CA plastics industry has expanded gradually over the years but the tonnage produced is still comparatively small. Very few new applications for CA are being found and many of its older applications have gone over to the newer and cheaper plastics materials. Consequently the publicity given to CA has been decreasing in recent years, and accurate production details have been increasingly difficult to obtain.

A breakdown of CA production in the U.K. and U.S.A. plastics industries since 1945 will however be given, sheet production will be emphasised wherever possible and plasticised cellulose nitrate (CN) production figures have been included for comparison purposes.

4.3.1. U.K. production of CA and CN in long tons 1945 - 1962

Year	Total	Moulding Powder (M.P.) and Extrusion Compound	Sheet, Rod and Tube	Cost of M.P. pence/lb	CN Sheet, Rod & Tube
1945	3,295	2,160	1,135	38	1,980
1946	5,735	4,295	1,445	35	2,100
1947	5,040	3,730	1,310	36	2,100
1948	4,640	3,100	1,545	34	2,440
1949	5,860*	4,000*	1,860	38	2,250
1950	7,950*	6,000*	1,950	34	2,620
1951	10,600	8,100	2,500	50	2,700
1952	9,500*	7,500*	2,000	46	1,125
1953	8,300	6,000*	2,300*	38.5	2,000*
1954	6,550	3,900	2,850*	33	2,300*
1955	6,500*	3,500*	3,000*	33.5	2,100*
1956	7,200*	4,000*	3,200*	36	2,400*
1957	8,000*	4,500*	3,500*	36	2,000*
1958	10,700	5,000	5,700	36	
1959	11,600	5,600	6,000	36	
1960	13,500	6,300	7,200	36	
1961	12,200	5,400	6,800	36	
1962	14,100	6,400	7,700	36	

Table 19.

N.B.

1. * = estimated by the author.
2. All CA statistics include cellulose triacetate, cellulose acetate butyrate and other cellulose esters and ethers.
3. Since 1957 all CN data has been grouped with CA.
4. Board of Trade statistics since 1960 have been recognised to give greater accuracy. Previously some CA outlets were duplicated.

Sources of information

1. Board of Trade Library, London. Information specific to CA was difficult to obtain.
2. Various extracts from trade journals and other trade literature too numerous to list in detail.

4.3.2. U.S.A. production of CA and CN in long tons 1945 - 1962

Year	Total	M.P. & Extrusion Compound	Rods & Tube	Sheet Under 0.003in.	Sheet Over 0.003in.	Cost of M.P. cents/lb	CN Sheet, Rod & Tube
1945	33,500	28,100	1,800	1,800	1,800	36	6,700
1946	37,100	27,800	2,100	3,600	3,600	34	4,850
1947	31,300	22,100	2,050	3,400	3,750	34	4,450
1948	21,700	13,400	1,100	3,700	3,500	42	3,650
1949	32,000	23,500	1,550	3,150	3,800	42	2,450
1950	48,000	35,700	1,900	5,500	4,900	43	2,750
1951	41,500	27,800	2,300	6,750	4,650	50	2,500
1952	37,900	26,100	2,250	5,100	4,450	50	2,000
1953	50,200	34,300	2,300	7,700	5,900	46	2,900
1954	49,500	33,800	2,350	7,800	5,550	46	2,200
1955	58,700	40,300	3,200	8,450	6,750	46	2,250
1956	60,200	41,000	3,150	8,750	7,300	46	2,350
1957	60,800	43,200	3,200	8,050	6,350	46	1,850
1958	60,900	41,800	3,900	7,300	7,900	48	1,400
1959	67,900	46,000	4,200	8,500	9,200	48	1,050
1960	62,000	39,700	4,500	8,450	9,450	48	650
1961	63,900	38,700	3,700	8,800	12,700	48	Nil
1962	70,600	42,800	4,150	9,150	14,500	48	Nil

Table 20.

Sources of information

1. U.S. Tariff Commission Reports.
2. Various extracts from trade literature too numerous to list in detail.

4.3.3. Production and application trends for CA sheeting in the U.K. and U.S.A. since 1945

4.3.3.1. Production in the U.K.

The production of CA sheet, film, rods and tube in the U.K. was 3,050 tons in 1943. The importance these products assumed in the manufacture of wartime materials is clearly shown by the fall to 1,310 tons which took place between 1943 and 1947. The lower production rates in 1944-45 were caused by a shortage of plasticiser and cotton linters. The most popular plasticiser at that time was dibutylphthalate, and this was extensively used for the preparation of insecticides during the wartime Burma campaign.

Prior to 1960 the U.K. Board of Trade returns were inaccurate because duplication of the tonnage of CA produced occurred, the same material was included in more than one category. The figures published under the heading "Sheet, Rod and Tube Production" are not specific to CA but include all other cellulose esters and ethers. This category further includes production figures for photographic and packaging film and all other profile shapes. However, recent Board of Trade estimates do show a definite trend towards an expanding CA sheet market, packaging film and spectacle frames being mainly responsible for this growth.

Film was produced and sold to capacity in 1962. Estimates of CA and cellulose triacetate (CTA) capacity in the U.K. are put at approx. 10,000 tons. The total production capacity is not expected to increase significantly over the next few years.

4.3.3.2. Production in the U.S.A.

As early as 1937 the production of CA sheet reached approx. 5,900 tons in the U.S.A. This prewar record was not passed until 1946. In 1948 one of the largest producers in the U.S.A. had a major accident which brought a shortage of CA sheet, particularly for the Christmas packaging market of that year. However, the overall U.S. acetate sheet market continued to expand after the war and the production of sheet materials more than doubled between 1950 and 1962. On the other hand the demand for CA moulding powder and extrusion compound was reduced immediately after the war. However, between 1950 and 1955 the latter materials staged a gradual comeback, and the demand since 1955 has been fairly constant.

Although the individual yearly tonnage of CA materials produced fluctuated considerably between 1945 and 1955, an overall growth rate of approx. 100% took place. Since 1955 the total yearly acetate market has been much steadier and a more gradual expansion of approx. 20% occurred between 1955-62.

Since 1960 many CA sheet producers in the U.S. have altered the form of their returns to the U.S. Tariff Commission. Some of the sheet which was previously recorded in the sheets, rods and tube category is now listed in the sheet over 0.003in. category. However, there is no doubt that the production of sheet over 0.003in. thick has expanded rapidly in the last three years. Two of the applications mainly responsible for this recent growth are CA record albums and book covers. The market for CA window boxes is now being replaced by oriented toughened polystyrene (TPS). Very little CA sheet is now being made by the block pressing and slicing process in the U.S.A. Approx. 50% of the sheet over 0.003in. thick was made by the band casting process and the other 50% was extruded. Laminating applications, sound tapes and gold foil stamping tapes are now the largest consumers of the under 0.003in. thick sheet category.

In 1962 there were only approx. 10 manufacturers of cellulose acetate flake in the world.

4.3.4. Selling price

The selling price of both CA and cellulose acetate-butyrate (CAB) moulding powder and sheeting has remained stable in recent years, despite competitive efforts by small sheet manufacturers. To lower the selling price appreciably, sheet producers must extrude sheet direct from flake and so cut out the costly compounding process.

Although acetate prices are stable they differ considerably in different countries as shown by the table below. These prices were extracted from European Chemical News, p.11, 10/5/63. Cents were converted to pence using $2.8 = £1.

Material	Selling Price pence/lb				
	U.K.	U.S.A.	France	Italy	Belgium
CA flake in 5 ton lots	33.5	31	-	36.5	32.5
CA sheet 0.05 mm thick in ton lots	84	63.5	67	99.5	-
CA moulding powder in ton lots	36	34	33	29	37
CTA in ton lots	42	44.5	42	-	47
CN in ton lots	90	-	67	155.5	74.5

Table 21.

4.3.5. Applications of CA sheet

Cellulose acetate sheet has long been used for such applications as aircraft navigation lights, cinema and theatre spotlights and sunglasses. The development of vacuum forming techniques since the early 1950's has helped to revive interest in CA sheet materials. Such sheeting possesses good forming properties giving sharp mould details and deep draws. By 1955 rigid CA sheet formed into display units, signs, transparent containers for packaging, toys, novelties, shoe lasts, hearing aids and brush backs, were becoming increasingly popular, particularly in the U.S.A. Other uses such as lampshades, laminated identification cards, and seamless tubes for packaging purposes were also expanding slowly. Acetate sheet is readily embossed and this process was extensively used by lampshade manufacturers. British Celanese have been extruding this CA sheet for the packaging industry since the mid 1950's. Vacuum formed signs for advertising and display purposes have accounted for a gradual growth in the production of CAB sheet. However, methyl methacrylate has the majority of this market because of its greater resistance to weathering and better colour stability.

Most of the above applications use acetate sheet varying in thickness between 0.010 - 0.050in. Before 1955 CA sheet was mainly manufactured by the block pressing and slicing process or by the film casting process, since that year direct extrusion from moulding powder and the film casting process have become increasingly popular. The latter process is the established method for producing X-ray and cinematograph films. It is also now in great demand for the manufacture of folder covers, card tabs, and thin film for the packaging industry.

4.3.6. The spectacle frame market

The largest outlet for thick extruded CA sheet is for spectacle frames in the optical industry. This outlet was first developed by the long established CN sheet materials. However as U.K. and U.S.A. production figures show, this market has gradually gone over to acetate sheet. The production of CN particularly over the past few years has dwindled down to less than 1,000 tons per year in both the U.S. and the U.K., the main reasons for this being

(a) the competitive nature of acetate sheeting, particularly in view of its lower specific gravity,

(b) the safety regulations issued against highly inflammable materials by health authorities.

The thickness of sheet produced for the optical industry is usually between 0.120 - 0.160in. before it is cut and processed into spectacle frames.

A census of the spectacle frame industry was carried out in the U.K. in 1954, 1958 and 1962 and in the U.S.A. in 1958 and 1962. Unfortunately the 1962 census figures will not be published until 1964-65. The census figures for previous years are given below.

U.K. census figures 1954 & 1958

Spectacle Frames	1954		1958	
	Number 1,000 Units	Value in £1,000	Number 1,000 Units	Value in £1,000
Mainly Metal	2,220	1,026	727	587
Mainly Plastic	3,660	1,273	3,780	1,515

U.S.A. census figures 1958

Spectacle Frames	1958	
	Number in 1,000 units	Value in $ 1,000
Gold filled fronts	1,124	2,753
Aluminium or other metal fronts	428	1,553
Plastic fronts	16,417	20,027
Combination fronts	7,108	16,929
Temple pieces, all types	52,288	25,736

According to "Manufacturing Optician", July 1962, 5,470,000 pairs of spectacles were sold in Britain in 1961. Approx. 80% of these are thought to have been plastic frames.

Although the above figures show us the size of the market for plastic spectacle frames, they do not indicate the proportion of this market held by CA sheet. To obtain an idea of this proportion the development of extruded CA sheet for the optical market will be given in detail.

4.3.6.1. The production of CA sheet for spectacle frames

It was 1949 before the first American company broke into this market with extruded CA sheet. This breakthrough was mainly due to the efforts of the Roland Brothers who were at that time working for American Celanese.

By 1953 CA sheet was used by over 25 leading manufacturers of spectacle frames in the U.S.A. and accounted for 5% of the overall market. By 1955 the number of manufacturers using acetate had risen to 35 and they accounted for 20% of the overall market. Today it is estimated that over 90% of all plastic spectacle frames manufactured in North America are of CA, although South America still use mainly CN.

The Nixon Nitration Works entered this field of CA sheet production in the early 1950's, although their CN sheet for this application had long been established.

The rest of the world have been very slow in following the example shown by America. In the late 1950's the Roland Brothers broke away from American Celanese and formed their own company. They introduced and patented a dual extrusion process which produces a continuous rectangular clear CA base sheet 4½ - 5in. wide, on top of which is extruded a thin continuous triangular wedge shaped strip of coloured CA approx. 2in. wide. The base sheet is approx. 0.220in. thick in the centre and tapers to approx. 0.160in. at the edges. The wedge shaped strip is approx. 0.050in. thick at its thickest point which is in the centre of the sheet. The sheet is then split down the centre and after further processing produces two spectacle frames the upper half only of which is coloured.

Nixon Nitration have developed a similar dual extrusion process, but they produce a flat CA strip approx. 4½ - 5in. wide, the centre 2in. strip of which is coloured. Many new techniques have been developed by both companies to try to imitate the original CN patterns.

The Roland Bros. process was sold under license to Dynamit in Germany and Courtaulds in the U.S.A. The latter company transferred the patent rights to Britain in 1962. The Roland Bros. have also recently formed a joint company with Artbangle in India for the manufacture and sale of this product.

Other companies developing an extruded CA sheet for this application are BX Plastics in England, Rhone Poulenc in France and Laminate in Italy.

The original CN patterns are still being matched in CA by combining a "block" overlay and a transparent base sheet. The latter may either be extruded or produced by the old block pressing and slicing method.

There are still relatively large quantities of CN sheet used for spectacle frames, particularly in Europe, South America, India, South Africa, Japan and Australia. However, there is little doubt that CN producing companies are having to focus more attention on the CA extrusion process because this is developing into a major outlet for cellulosic sheeting.

4.3.6.2. The advantages of CA over CN spectacle frames

The main advantages are:-

1. Non-inflammable
2. Less tendency for crystal material to yellow
3. Less tendency for colour to fade
4. No solvent loss and therefore excellent dimensional stability
5. Cheaper
6. Shorter delivery times

The latter point is very important because it enables small trial batches of spectacle frames of a new design to be marketed. If these are successful then larger supplies will quickly follow while the style is still popular. With CN the retailer often has to decide whether to risk buying a large batch of a new design and be overstocked if it did not sell. On the other hand if he bought a small trial batch and it was popular he may have a long wait before he could restock that particular design. Such delays may be damaging to business because spectacle frames are now sold on a basis of individuality and beauty. Women particularly in the U.S.A. are becoming increasingly fashion conscious and multiple sales of spectacle frames to a single user have been on the increase. The production of a coloured CA clip on plaque has become increasingly popular in recent years, these may be attached to the upper half of womens spectacle frames and altered at will to suit their personal appearance.

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5. MARKET REQUIREMENTS AND CONTROL TESTING

5.1. Introduction

In this section, the market requirements of sheet materials are considered with special reference to the use of TPS sheet for refrigerator door and tank liners, PVC sheet for roofing and other building applications, and CA sheet for spectacle frames. The routine control testing of extruded sheet materials which is necessary to maintain adequate quality standards is also described in relation to the same three applications.

5.2. Selection Of Sheet Materials

The requirements of an extruded plastics sheet material in any application is dependent on the needs of further fabrication processes and of the end product. These requirements are controlled by the design of the product, and are met by properties derived from

(i) the basic raw material
(ii) the sheet extrusion process
(iii) further fabrication processes

The design of the product is obviously itself controlled largely by the application but the suitability of a particular plastics material for a particular application may be affected by design details. For instance, one of the problems most frequently met is that of stress concentration at critical points, often at points joining two components, or at sharp corners. If the product is carefully designed so the maximum stress concentration is reduced by the use of load spreading fixings or rounded corners, it may be possible to use a less strong and cheaper plastics material, or it may make possible the use of plastics in place of conventional materials. It is important to recognise when inferior design is making undue demands on a material as this may prevent its use altogether or require a superior grade of material which cannot compete on price. Having emphasized this point it is beyond the scope of this thesis to consider product design further.

Probably the most important of all properties of an extruded plastics sheet is its price. However, it is not within the scope of this report to explore in detail the complexities of assessing the costs of producing sheet materials and the profits which may be expected to be added in order to arrive at a selling price. In any case, the story would be incomplete without a similar review of the costs of fabrication, which are still more inaccessible. It is considered sufficient to note here some of the salient points which affect the main applications considered.

5.3. Properties Derived From The Basic Raw Material

It is often difficult to distinguish completely between the properties derived from the basic polymer and modifications derived from the sheet extrusion and subsequent fabrication processes to which it is subjected. However, the type and grade of polymer is the major factor controlling the basic raw material properties. Many different grades of TPS, PVC and CA materials have been developed to suit particular market applications, the inherent properties of these grades depend on the materials, method of manufacture and processing conditions used to modify the properties of the basic raw material.

Some properties may be common to all grades of plastics materials which are used to produce good quality sheet by the extrusion process, for instance all must possess good flow and extrudability characteristics. Other properties which may be important, depending on the methods of further fabrication used and the application of the final product, include density, colour, heat resistance, cold resistance, inflammability, flexibility, rigidity, hardness, impact strength, tensile strength, elongation, creep, thermal conductivity, light stability, ageing, weather resistance, dimensional stability, permeability, non-corrosiveness, chemical resistance, solvent resistance, odour, tainting, toxicity, clarity, transparency, gloss, surface appearance, formability, and machinability as well as many other physical properties.

5.3.1. Modifying the inherent properties

The basic raw materials are often modified so that their properties are particularly suited to a specific application. For instance the ease of flow and extrudability of TPS and PVC materials can be modified by varying the molecular weight and the molecular weight distribution of these polymers by changing the polymerisation method and/or processing conditions. In the case of CA materials both the molecular weight and the degree of acetylation can be varied. Flow properties are also modified by the use of internal or external lubricants and plasticisers in the mix fed to the extruder. Other important formulation ingredients such as stabilisers and pigments etc. will also alter the properties of the product.

Vinyl chloride may be copolymerised with vinyl acetate or vinylidene chloride or PVC may be combined with synthetic rubber materials so that the properties of the final blend are suitable for particular manufacturing techniques and applications.

Space does not permit a full description of the many ways in which TPS, PVC and CA materials are modified to suit particular applications and the underlying reasons, but the following discussion showing how the impact strength of PS may be modified is a good example of the way in which different grades of the same plastics material are developed.

Crystal PS materials have poor impact strength characteristics and in order to improve these properties TPS grades were developed. These are normally made by either mechanically blending or graft polymerising rubbery materials such as copolymers of butadiene with PS, or by a combination of both these processes. In the mechanical blending process a synthetic rubber is compounded with PS in a banbury or extruder or on a two-roll mill, whereas in the graft polymerisation process a solution of a synthetic rubber dissolved in styrene monomer is polymerised by any suitable method.

Usually such blends contain 5 - 15% of synthetic rubber and normally the toughness and elongation properties of TPS materials increase in value as the percentage of synthetic rubber increases. The method and processing conditions used for combining synthetic rubber and PS can greatly effect the impact strength characteristics of the TPS materials produced, thus the impact strength of two or more blends which contain similar proportions of each constituent may be widely different. However, it is not necessarily the TPS grade with the highest impact strength which is the best material for a particular application, for instance in cases where a heat polishing technique is used to induce a gloss on TPS sheet it is often found in practice that the higher impact strength materials are less suitable.

A recent paper by Vincent (1) attempts to explain why the impact strength characteristics of PS are improved by the addition of copolymers of butadiene or copolymerising with acrylonitile, and why a combination of both these processes has a marked effect which accounts for the remarkable toughness properties of ABS blends.

5.3.2. Testing the raw materials

Specifications and test methods for determining the inherent properties of the more commonly used plastics materials have been developed for production control and classification purposes to suit particular applications, and usually material suppliers work to those developed by a national body e.g.

British Standards Institution (BS)
American Society for Testing Materials (ASTM)
Official German Standards (DIN)

Such data on plastics raw material properties as supplied by the manufacturer are often helpful to the raw material user, but the latter may conduct more tests and even devise new tests to investigate properties which are specific to a particular application or processing technique. For instance to overcome excess moisture troubles during the sheet extrusion process a sheet manufacturer may devise a test to ensure the moisture content of the feed material is within the limits necessary for good quality sheet production. Also, for particular applications in which plastics materials are likely to come into contact with food stuffs, tests are devised to determine the residual monomer content or other volatile materials in a polymer if this is likely to transfer an unpleasant odour or taste to the foodstuff.

5.4. Properties Derived From The Sheet Extrusion Process

Because of space restrictions the following discussions and examples are again restricted to TPS materials. Obviously all properties of an extruded plastics sheet derive from properties inherent in the basic raw material or compound, but certain properties are particularly susceptible to modification by the extrusion process, and may even be said to be controlled by it. These properties which can be controlled and modified by the sheet extrusion process include surface appearance, reversion, impact strength, and sheet dimensions. These sheet properties vary depending on the grade of plastics material extruded, the extrusion technique and processing conditions used. The way in which these factors affect the sheet properties is very complex, however, such relationships have been studied in great detail by many workers for TPS sheet materials (2,3,4,5,6,7, 8,9).

Clearly, where the extruder operator can modify properties in important respects it is essential to exercise control, where necessary through routine testing, so that these properties remain within acceptable limits. It is, therefore, appropriate to consider jointly these properties and their control, see sections 5.4.1 - 5.4.6.

5.4.1. Surface appearance

Most TPS materials used for sheet production have an inherently low gloss. Different techniques are used to improve the gloss characteristics of TPS sheet to give it added sales appeal, and these techniques are outlined in Part II sections 11.2.2. to 11.2.4. The degree of surface gloss obtained using any one of these methods with a particular grade of TPS is dependent on the technique used and the extrusion and take-off operating conditions.

Other properties such as impact strength and elongation at failure may be effected by a change to the surface appearance, for instance the values of both these properties are normally decreased when crystal polystyrene film is laminated to one surface of the base TPS sheet to improve the gloss.

The gloss level of extruded sheet is normally measured with a 60° glossmeter and expressed in photovolts, Gardner units or as a percentage based on a black glass standard (52 Gardner (Gloss) units = 100%) ASTM D523 - 53T. Other checks such as colour matching and the inspection of the sheet surface for lines, nibs, pock marks, silver spotting and foreign matter etc. are normally performed by visual inspection.

5.4.2. Reversion

The ability to revert is induced in extruded sheet materials by stretching the sheet in the longitudinal or transverse direction or both as it is cooling, which orientates the polymer molecules. When the sheet is reheated it tends to revert to its original shape and in doing so shrinks in one or two directions. Reversion is reduced to a minimum by synchronising the speeds of the calender stack rolls and the rubber pull-off rolls (see part II section 12.2.14.). Uneven reversion characteristics may also be induced in the sheet by the extrusion process (see part II section 12.2.15).

Reversion characteristics can greatly alter the thermoforming properties of extruded sheet, for instance, sheet with high or uneven reversion may not form evenly and this may give rise to a product with the wrong dimensions. The formed article may also contain thin spots which weaken its structure and make it unsuitable for the desired application or the sheet may even split during forming. Extruded TPS sheet with low and even reversion characteristics is normally required for the production of refrigerator door and tank linings. The degree and direction of orientation can also change the ultimate elongation and tensile strength properties of extruded sheet.

Reversion is normally measured by immersing samples of the extruded sheet in hot oil at a fixed temp. for a fixed period of time, and the reversion value is usually expressed as the percentage shrinkage observed in the longitudinal direction (see BS 3290 App.B). Samples from a number of positions across the width of the sheet are normally tested to check evenness of reversion.

5.4.3. Impact Strength

Two methods are used for measuring the impact strength of sheet materials, the drop-weight impact test (see BS 2782 Method 306C) and the notched bar Izod impact test (see BS 2782 Method 306A), the former is the more popular method used for testing sheet materials because it more nearly represents the type of impact experienced in service. The values of impact strength are normally expressed as ft. lb. or ft.lb./inch of notch depending on the method used.

The measured drop-weight impact strength can be markedly dependant on the nature of the sheet surface on the face <u>away</u> from the blow, i.e. the face that is stressed in tension by the impact. In the case of brittle failure, the important type, the initiation of a crack in this surface is a major factor in determining the energy required to cause a break, so that a brittle surface can greatly weaken an otherwise tough sheet.

This is particularly important in the case of TPS sheet where the methods used for producing an improved gloss on one surface of the sheet are known to considerably lower the impact resistance of that surface. The value of the impact strength of extruded TPS sheet is also dependant on the thickness, reversion and temperature of the sheet under test.

5.4.4. Sheet dimensions

The sheet extruder operator is responsible for the thickness, width and length dimensions of extruded sheet, he must also make sure that the sheet is cut squarely and that all dimensions are controlled to within the normal commercial tolerance limits. For economic and further fabrication reasons these dimensions are now controlled to within very fine limits. For instance if a minimum sheet thickness of 0.200 inches is specified, a supplier producing sheet with an average thickness of 0.205 inches and a tolerance of ± 0.005 inches offers a financial saving to a final product manufacturer compared to one producing sheet averaging 0.210 inches with a tolerance of ± 0.010 inches, if the sheet is sold on a weight basis. Also the thermo-forming process normally produces better results if the sheet thickness is controlled to within fine limits.

Micrometers and steel tape measures are normal equipment for measuring the dimensions of sheet materials and in the U.K. these values are usually quoted in inches.

5.4.5. Formability

The sheet manufacturer often develops a thermo-forming test to investigate the properties of an extruded sheet material. If this test is well designed, it will act as a combined test for the four properties previously mentioned. For instance the faults associated with poor gloss retention, pock marking, high or uneven reversion, and poor sheet dimensions will all be immediately apparent and tests could be conducted to determine the impact strength of the formed article.

5.4.6. Other tests for extruded sheets

The tests conducted and the standards and tolerances set are chosen by sheet manufacturers to ensure that the sheet will undergo any further fabrication process necessary and at the same time impart the essential properties to the final product. Most of the tests used for production control purposes have already been mentioned in previous sections, however, other tests to determine the tensile strength, elongation and flexural strength of the sheet material at low temperatures, may be conducted if these properties are particularly important to the final product manufacturer. Longer term tests may also be developed by the sheet producing companies to investigate stress-cracking, creep, ageing and weathering properties etc.

5.5. Properties Derived From The Fabrication Process

Further modifications to the basic raw material and extruded sheet properties may be made by the sheet fabrication process, and these may reduce or emphasise those modifications already made by the sheet extrusion process. Properties which may be modified include surface gloss, definition and eveness of formings, thickness, dimensional stability, impact strength and rigidity etc.

Control of these properties is to some extent within the power of the fabricator but it is considered to be outside the scope of this report.

5.6. Practical Applications

In table 22 below an attempt has been made to summarise the relative importance of various properties in the selection of the sheet materials used for the three important applications under consideration i.e. refrigerator door and tank liners, roofing and building applications, and spectacle frames.

PROPERTY	APPLICATIONS		
	Refrigerator Linings	Roofing	Spectacle Frames
General toughness	B	B	A
Impact strength	B	B	B
Rigidity	B	B	B
Flexural strength	B	B	B
Fatigue resistance and stress relaxation	A	A–B	C
Low Temp. performance (0°C)	C	B	B
High temp. performance (60°C)	C	B	B
Non-toxic	B	C	B
Non-taint	*A	C	C
Freedom from odour	*A	B–C	B
Chemical resistance	B	A–C	C
Colour stability	B	B	B
Clarity/transparency	C	*A–C	*A–C
Surface appearance of sheet	A–B	A–C	B–C
Weathering	C	*A	B–C
Low burning rate	C	*A	*A–C
Forming	*A	B	C
Machining	A	B	*A

Table 22

Where A represents – property very important
B represents – property important
C represents – property not important

* indicates properties of major importance for specific applications.

5.6.1. TPS sheet for refrigerator liners

A fuller account of the development of this application is given in part 1 section 2, but it is of interest to review the salient points here. The most important feature is undoubtedly the ability of TPS sheet to be thermoformed into deep-drawn and quite intricate shapes, giving wide scope to the refrigerator designer in the provision of a deep cabinet, shelf runners, shaped egg-trays etc. The sheet is rigid with a pleasing appearance and a smooth easy to clean surface. Other properties include freedom from odour or taint, and the ability to be trimmed and prepared for assembly by machining, but TPS is not very different from other plastics in these respects.

Even the apparently obvious requirement of strength at low temperatures is not of prime importance, because the stresses encountered in service are comparatively low, but with certain designs (including those using foamed-in-place insulation) higher loadings occur, and formulations with improved low temperature toughness and resistance to environmental stress cracking have been developed.

Because the door and tank liners are not a major expense in the production of refrigerators, the present day low prices and profits for TPS extruded sheet materials can be mainly attributed to the severe competition between sheet suppliers.

As stated previously it is those properties of TPS sheet which govern its "formability" which are of dominant importance. Foremost among these are the reversion and thickness tolerance characteristics which must be good for the sheet to form evenly, while other essential properties include the retention of surface smoothness and gloss during further fabrication of the extruded TPS sheet material. Routine checking of reversion must be carried out, and a qualitative forming test in which samples of sheet are heated and formed into a simple shape, and then carefully inspected, is conducted to assure consistent good quality.

In addition to the normal examination e.g. for surface appearance and colour etc. long term tests for fatigue, creep, stress relaxation and other properties are often developed by the refrigerator manufacturer to make sure that the plastic liner is capable of withstanding normal assembly stresses without cracking. These tests are usually tedious and costly and therefore must be carefully designed to give the most significant information. For instance, door slam tests (4) have been developed to make sure that the plastic liner when assembled into a finished door possesses the combined high fatigue and impact resistance necessary to withstand repetitive door slamming shocks. Cyclic racking tests (4) have also been devised to investigate the fatigue properties of a refrigerator door assembly.

Under present competitive conditions exact cutting to size of sheets is important because many refrigerator manufacturers are now working to extremely close tolerances with modern thermoforming machines to ensure minimum wastage of time and materials.

5.6.2. Rigid PVC sheet for roofing and other building applications

A fuller account of the development of this application is given in part 1, section 3, however it is of interest to review the salient points here. Corrugated PVC sheet has become a serious competitor to other materials in the building industry because it is a cheap, light-weight roofing material.

Undoubtedly one of the major reasons for its success as a roofing material is the fact that it can be made transparent to light and can therefore permit access of natural daylight into buildings where this is required. Checks on the clarity of transparent sheet must therefore be conducted at regular intervals. Other important properties include good weathering and light stability characteristics, rigid PVC sheets also satisfy many of the low burning rate safety specifications which are now enforced in many countries to prevent fires spreading rapidly in buildings but their use is still restricted because burning brands e.g. from a nearby building on fire, may melt the sheet and pass through it thus causing a fire hazard. Wire reinforced sheet has now been introduced on the market as an attempt to overcome this disadvantage. The excellent chemical resistant properties of rigid PVC sheet materials make it particularly useful in cases where there is a high concentration of corrosive gases or vapours as in the atmosphere of some chemical plants.

The sheet does not warp in use and possesses adequate strength and low temperature and high temperature performance characteristics to withstand the normal stresses and environmental conditions imposed on it by different climates in many parts of the world. For instance when used as a roofing material in Scandinavian countries the sheet possesses the necessary flexural strength properties to resist heavy loadings of snow for extended periods. Such properties have been investigated at low temperatures by supporting the sheet specimens at their corners and then uniformly loading the surface with sand or snow, in this way sheet deflection against load data is obtained and the purling distances of the supporting framework necessary to support the sheet are determined.

Good forming, machining, and surface appearance characteristics come high on the list of properties, but these qualities are of secondary importance with sheet used for roofing purposes because the latter is usually viewed from a distance. However, formings must be good enough to prevent rainwater leaks etc. at overlapping points.

5.6.3. CA sheet for spectacle frames

A fuller account of the development of this application is given in part 1 section 4, but it is of interest to review the salient points here. CA sheet has captured much of the CN spectacle frame market particularly in the U.S.A. because it satisfies the safety specifications which have been imposed against the use of highly inflammable materials for this application in many countries, and it can now be manufactured by a cheaper and quicker process.

CA sheet is capable of giving a wide range of colour effects and this together with its ability to be machined and polished into a good looking finished product accounts for its success in this application. The surface appearance properties of spectacle frames are of the utmost importance and because large quantities of transparent sheet are used, regular checks are made to ensure that the clarity of these materials is of an extremely high standard.

Other properties which are important to a lesser extent include general toughness, rigidity, flexural strength and colour stability.

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6. COMPETITIVE SHEET MATERIALS

6.1. Introduction

Extruded TPS, rigid PVC and CA sheet materials sometimes compete against each other for the same market application. Competition between individual plastics will become sharper in the next few years and consequently profit margins may narrow even further as prices are reduced unless improved production efficiency can keep pace with lower prices.

When selecting a sheet material for a particular application it is necessary first to analyse the application and determine which of the properties required are the most important. Many of the major properties which usually have to be considered, excluding price, have been outlined in section 5 but in addition to these such properties as performance in printing, embossing, heat sealing and welding may be of primary importance.

The final choice of a material for a particular application may be severely restricted by the properties required, but on the other hand there may be several suitable materials. In the latter case the final choice usually depends on the material and processing costs. Lower costs of sheet production, further fabrication or finishing may put a more expensive plastics material in a very competitive position against some of the cheaper basic materials which incur heavy processing costs. It is interesting to note that a reduction in a particular plastics sheet material cost usually opens up new market applications within a year, despite the unavoidable time taken to make the corresponding changes or innovations in design and manufacture.

6.2. Table of Competitive Sheet Materials

Materials which compete in some way with TPS, PVC, and CA extruded sheet cover a very wide range as shown by table 23.

Material Type	Sheet Material Classification		
	Homogenous	Micro Heterogenous	Macro Heterogenous
Synthetic Plastics			
a) Thermoplastics	PS,PVC,PVA, PVAL,PVDC, CA,CTA,CAB, CN,PMM,PC, PE,PP,PTFE, Polyamides.	Thermoplastics plus fillers. Polymer blends, e.g. TPS, ABS, Expanded thermoplastics e.g. expanded PS.	Laminates of all types e.g. paper textiles, wood.
b) Thermosets.	MF,PF,UF, PU, Synthetic rubbers.	Thermosets plus fillers. Expanded thermosets e.g. expanded PU.	
Natural Materials	Cellulose derivatives rubber, bitumen.	Cork, wood, mica, asbestos, cardboard, compressed paper and wood pulps and textiles	Hardboard, chipboard, plywood, and other reinforced laminates, sandwich, fluted, ribbed, corrugated and honeycomb structures.
Glasses	Glasses of all types.	Ceramics	Polyester glass laminates and other glass reinforced plastics. Vitreous enamelled metals. Non-woven glass fabric.
Metals	Aluminium, nickel, iron, copper, zinc magnesium, lead, titanium.	Brass, steel and all other metal alloys	Tin plate and all other metal laminates including metal to plastic laminates & painted metal surfaces.

Table 23.

6.3. Main Materials Of Competition

In the space available it is impossible even to summarise the major applications and properties of all the materials given in Table 23, but a short survey of the materials which compete for the major applications of extruded TPS, PVC and CA sheet materials is given. The major applications for TPS, PVC and CA sheets have been discussed fully in part 1 section 2 - 4, and the other main sheet materials which are competing for some of these markets include

6.3.1. Vitreous enamelled or painted steel
6.3.2. Vinyl metal laminates
6.3.3. ABS Sheet
6.3.4. Glass reinforced polyesters
6.3.5. PMM sheet
6.3.6. Celluloid sheet
6.3.7. PP sheet.

6.3.1. Vitreous enamelled or painted steel sheet

Such sheeting competes directly against TPS sheet in the refrigerator industry, though unfortunately no figures are available showing the proportions of plastic to steel sheeting used for refrigerator interiors in either the U.K. or the U.S.A. It is also widely used in the facing of shop display units.

Thin gauge steel sheeting has the advantage of being much stronger than TPS and yet the latter has displaced it almost entirely in refrigerator door liners and also in the smaller refrigerator tank linings. It did so mainly because of the cheaper further fabrication and finishing costs. The door and interior liners of a refrigerator are often of complex design and the tooling up necessary for fabricating steel liners and the vitreous enamelling or paint spraying and stoving processes which follow are more expensive than the relatively simple thermoforming process. The latter is essentially a one stage process but the former is a two or three stage process and requires more physical handling.

In the past TPS sheet has not been sufficiently strong or rigid to withstand the weight distribution encountered in a tank liner of more than six cubic feet capacity. However, the advent of a polyurethane foaming-in-situ process for insulation purposes has altered the situation, in that this latter process gives much more support to TPS tank linings and will permit their use in the large domestic and shop refrigerator markets and display units.

At present there is little sign of a plastics material replacing steel sheet for the exterior of refrigerators or display units, because of the better structural strength, impact strength and resistance to marring exhibited by the latter sheet material. However, future developments of TPS, rigid PVC or ABS sheet may eventually displace enamelled or painted steel in this application. On the other hand a combination such as a thin foil of PVC or polyvinyl fluoride laminated to steel may be the next major development which will give the necessary strength and finish required.

6.3.2. Vinyl-metal laminate sheet

The above material is usually a combination of plasticised PVC bonded to steel and it competes against rigid PVC and TPS whenever a combination of rigidity, strength and pleasing decorative finish are required. It can be used widely in building constructions, vehicles, consumer goods, and industrial equipment, and it competes directly against rigid PVC and TPS sheet in such applications as curtain walling, partitioning, shop fittings, accoustic ceilings, ductings, instrument panels and the interior of transport vehicles, trays and radio and television cabinet facings.

Vinyl-metal laminates are suited to desk tops, shelves and shop fittings; they give a pleasing decorative finish and also often compete directly against vitreous enamelled or painted steel sheets. Previously no vinyl-metal laminate has been able to achieve the appearance and properties of white vitreous enamelled steel, however, recent developments with a white polyvinyl fluoride film bonded to steel have shown very promising results. Such sheeting may soon be competing against the established materials used in the manufacture of refrigerators, washing machines, spin dryers etc.

6.3.3. ABS sheet

This material has received vast publicity in the U.S.A. but as yet the quantities of ABS sheet absorbed by markets in which it competes against TPS and rigid PVC sheets have not amounted to very much. In the U.K. only approx. 4,000 tons of ABS were consumed in 1962 and only a very small proportion of this was in sheet form. ABS possesses similar properties to TPS sheet, has the added advantages of better gloss, rigidity and impact strength, but it suffers from the serious disadvantage of being higher in price.

ABS sheet is finding new outlets in such applications as portable typewriter and camera cases, luggage cases, furniture, trays, panelling in aircraft, railway carriages, coaches and cars, and safety helmets for mining and building construction workers and motor cyclists. It has tended mainly to develop and expand new markets rather than to compete directly against established TPS and rigid PVC sheet markets.

However, there is little doubt that ABS sheet would take most of the TPS refrigerator market if the price was right.

6.3.4. Glass reinforced polyester sheet

The above material has become much more competitive in recent years because of severe price reductions, e.g. between 1959 and 1962 the U.K. selling price fell from 3/10d per lb. to 2/3d per lb. A percentage breakdown of the U.K. and U.S.A. market outlets for this material over the past few years is given in table 24.

Application	U.K.%				U.S.A.%			
	Year				Year			
	1959	1960	1961	1962	1959	1960	1961	1962
Corrugated sheet	30	30	30	28	15	17	20	21
Land transport vehicles	18	18	18	20	14	17	18	20
Boats	16	16	15	18	28	22	18	17
Aircraft & missiles	12	11	10	6	11	11	11	12
Consumer products	10	10	10	12	18	14	14	13
Chemical plant	5	5	5	5	3	5	5	5
Miscellaneous	9	10	12	11	11	14	14	12

Table 24.

Glass reinforced polyester sheet possesses exceptional strength for a comparatively light weight material and this has made it popular as a roofing material in the building industry as shown by table 25.

1962 Consumption figures for glass reinforced polyester sheet in roofing applications.

Country	Weight consumed in long tons	Percentage of total market for glass reinforced polyesters
Italy	10,000	70
Germany	12,000	50
France	-	46
U.K.	4,200	28
U.S.A.	25,000	21
Canada	900	15
Norway	200	15

Table 25

Table 25 shows that the percentage of the total glass reinforced polyester market used for roofing purposes varies considerably in different countries. This application for glass reinforced polyester materials is likely to suffer from severe competition by extruded rigid PVC over the next few years but it is thought that land and sea transport application will more than compensate for any loss in most countries.

Other applications for this material which are still expanding include washing machine tubs, chairs, shower cubicles, baths and coal and log fire effects for electric fires. Glass reinforced polyesters and extruded ABS sheet materials are likely to compete strongly in such applications as sink units, chairs and baths over the next few years.

6.3.5. PMM sheet

Extruded PMM sheet was introduced into the U.S. market in 1958 and the U.K. market in 1960, however, this process seems to have met with many technical difficulties and the majority of PMM sheet is still made by the established casting process. The volume of PMM sheet produced in the U.K. is known to be increasing but it is very difficult to obtain accurate production data.

PMM sheet is moderately tough and possesses extremely good light transmission and weathering properties. It competes directly against polystyrene in the indoor illuminated sign market and other light and display applications, it also competes against corrugated

PVC in roofing applications but suffers from the disadvantages of being inflammable and higher in price. In 1961 a new thin oriented transparent sheet less than 3/32 inches thick entered the U.S.A. market to compete against rigid PVC for rear windows in convertible cars, motor cycle screens, patio shelters, partitions and roofing. Other applications include illuminated ceilings, sun screens and an increase in its use is expected in the fabrication of sanitary ware, baths, shower units, sinks and basins. In many of these applications it will meet increasingly strong competition from PS, TPS, PVC, and ABS sheet materials.

PMM entered the spectacle frame market just prior to 1960 and competing strongly against celluloid and CA in the U.K., it is thought that PMM may have captured up to 15% of the total plastics spectacle frame market by the end of 1961. However, further fabrication difficulties have considerably reduced the popularity of PMM for this application and the recent trend in the U.K. is towards extruded CA sheet.

6.3.6. Celluloid sheet

This competes directly against CA sheet in the spectacle frame market. A full comparison of the production and applications of these two sheet materials has already been made in part 1 section 4.

6.3.7. PP sheet

PP sheet production and application have not yet expanded sufficiently to enable an accurate prediction of the ways in which this material will compete against TPS and rigid PVC. PP production in the U.K. began in 1960 and consumption figures for 1961 and 1962 were approx. 2,000 and 3,500 tons respectively and this figure is likely to double in 1963. The quantity of thick sheet produced in 1962 is not accurately known but it is thought to be less than 5% of the total PP consumed in that year.

However, PP extruded sheet 0.020 - 0.400 inches thick is now available in the U.K., the thicker sheets are already finding an outlet as water tank linings, cisterns and luggage, and thinner sheets are used as kick panels. Rapid expansion is taking place in this field in the U.S.A. (e.g. production of PP rose from 10,000 tons in 1959 to 45,000 tons in 1962). Approx. 18% of the total PP produced in the U.S.A. in 1962 was converted into film and sheet and considerable competition can be expected from this material in the near future when the price of PP is lowered.

6.4. Raw Material Costs

A comparison of the raw material costs and some of the major physical strength properties which characterise various plastics and metals are given in table 26. Numerous attempts have been made by the author to relate these toughness properties with material costs but no simple relationship has been devised.

Raw Material	Specific Gravity (ASTM D 792)	Impact Strength (ASTM D 256) ft. lb. in. notch Izod	Tensile Strength (ASTM D 638) 1000 lb/in^2	Compressive Strength (ASTM D695) 1000 lb/in^2	Flexural Strength (ASTM D 790) 1000 lb/in^2	Weight/cu:ft. lb.	Cost/unit wt. pence /lb.	Cost/unit vol. shillings cu/ft.
PS	1.06	0.2-0.5	5.0-12.0	11.5-16.0	12.0-17.0	66	19	105
TPS	1.04	0.5-3.0	3.5- 6.5	4.0- 9.0	5.0-10.0	65	23	124
ABS	1.04	3.0-12.0	2.5- 9.0	2.5-11.0	3.6-13.5	65	45	244
Rigid PVC	1.4	1.0-3.0	8.5	8.0	13.5	86	13.5	97
CA	1.32	0.4-5.2	1.9-8.5	2.2-36.0	2.0-16.0	82	36	246
CAB	1.21	0.8-6.3	2.6- 6.9	2.1-22.0	1.8- 9.3	76	54	342
CN	1.4	-	-	-	-	86	90*	645*
PE (LD)	0.92	-	1.0- 2.3	-	-	57	20	95
PE (HD)	0.94	1.5-12	3.1- 5.5	2.4	2.0- 3.0	59	26	128
PP	0.91	0.6-6.0	4.3- 5.7	8.5-10.0	-	57	28.5	135
Nylon 6	1.14	1.0-3.6	10.2-12.0	7.0-14.0	8.0-16.0	71	92	544
PMM	1.2	0.3-0.5	7.0-11.0	12.0-20.0	13.0-17.0	75	40	250
PF	1.28	0.2-0.4	7.0- 8.0	10.0-30.0	12.0-15.0	80	14	95
UF	1.5	0.2-0.4	6.0-13.0	25.0-35.0	10.0-16.0	94	22	172
MF	1.5	0.2-0.4	7.0-13.0	40.0-45.0	11.0-14.0	94	30	235
# Glass reinforced	1.5	-	-	-	-	94	30*	235*
Cast Iron	7.2	-	-	-	-	450	2.5	94
Mild Steel C.R.	7.85	35	70.6	110.0-130.0	-	490	7	286
Stainless Steel	7.92	-	-	-	-	494	39.5	1626
Aluminium 99%	2.8	20	12.0	-	-	175	19.5	284
Duralumin (Al-Si-Cu-Mg)	2.8	-	35.8	-	-	175	-	-
Copper	8.9	-	-	-	-	556	25	1158
Phosphor Bronze (Cu-Sn-P)	8.9	-	58.2	-	-	556	-	-
Brass (Cu-Zn:70/30)	8.5	66	46.3	-	-	530	28	1237
Zinc	6.8	-	-	-	-	422	8	264
Magnesium	1.8	-	-	-	-	112	22	205

Table 26

This should read "Glass reinforced polyesters".

NB:-
(1) All prices are based on one ton lots.
(2) * Denotes cost in sheet form.
(3) Raw material costs were extracted from
  a) European Chemical News - May 1963
  b) Machinery - May 1963
  c) Machinery Market - May 1963
  d) BXP Ltd. and Shell Chemicals trade literature.
(4) Raw material strength properties were extracted from a Shell Chemicals pamphlet "Properties of Plastics".

The above table is an approximate comparison of some of the major strength properties and raw material costs of a few plastics and metals. These materials in sheet form compete directly against each other in a number of specific applications. However, in the space available it is impossible to make direct comparison between the manufacturing costs of final products produced from plastics and metal sheeting, because these costs are dependent on many factors e.g.

(a) the properties required by the final product
(b) the thickness of sheeting used
(c) the processing costs
(d) the finishing costs

6.5. Cost To Thickness Relationship Of Some Plastics Sheet Materials

A comparison of the selling price/sq.ft. for some of the major plastics sheet materials over a thickness range of 0.010 - 0.250 inches, is given in table 27.

NB:-
(1) All prices are based on one ton lots
(2) Sheet prices were obtained from BXP and I.C.I. trade literature (1962)
(3) All prices are based on unpigmented materials.
(4) * Denotes materials which cannot be supplied in a clear transparent form.
(5) All prices for CA and celluloid sheets are based on optical grade materials.
(6) An extra charge is incurred for pigmented, laminated or veneered sheets.

Selling price/sq.ft. of plastics sheeting

Material	Method of Production	Surface	Thickness of Sheet in Inches								
			0.010	0.020	0.030	0.040	0.060	0.100	0.120	0.180	0.250
PS*	Extruded	Nat.	2d	4d	6d	8½d	1/0½d	1/9d	2/1½d	3/2d	4/5d
PVC (Rigid)	Extruded	Nat.	4¼d	8½d	1/1d	1/5d	2/1½d	3/6½d	4/3d	-	-
PVC (Rigid)	Calendered & Pressed	Polished	8d	1/3d	1/10d	2/5d	3/8d	5/6d	6/7½d	9/11d	13/3d
PVC (Flex.)	Calendered & Pressed	Polished	6d	11d	1/3½d	1/10d	2/11d	-	-	-	10/0d
CA.	Extruded also Block & Slice	Nat.	8d	1/2½d	1/9d	2/4d	3/6d	5/10d	7/0d	10/6d	14/6d
CN.	Block & Slice	Nat.	8d	1/2½d	1/8d	2/3d	3/4d	5/6d	6/8d	10/0d	13/6d
PMM	Cast	Polished	-	-	6/0d	5/0d	4/6d	4/3d	4/6d	6/0d	7/6d
PE* (HD)	Extruded	Nat.	3d	6d	9d	1/0d	1/6d	2/6d	3/0d	4/6d	6/2d
PP*	Extruded	Nat.	3½d	7d	10½d	1/2d	1/7½d	2/8½d	3/3d	4/10½d	6/6d

Table 27

Table 27 shows how the competitive nature of each plastics sheet material alters over a wide thickness range. It is noticeable that cast PMM sheet is the odd one out because the manufacturing costs for thin sheet are very high in comparison to the costs for thick sheet, (e.g. cast PMM sheet is cheaper than calendered and pressed rigid PVC above a thickness of 0.070 in. but is more expensive below that thickness).

PART 11

THE EXTRUSION PROCESS

1. INTRODUCTION

The following sections comprise a critical study of many of the technical factors involved in the extrusion process and as such form the most important part of this report. All the practical work described in these sections was conducted at B.X. Plastics Ltd., because of the limited facilities available to study the extrusion process at the Brunel College of Advanced Technology.

Facilities were made available for the student to operate large single screw sheet extruders for extended periods and thus to gain practical experience of the problems involved in the extrusion of a variety of plastics sheet materials on a commercial scale. However, for production economy reasons it was impracticable for the Company to allow the student free access to these sheet extruders for research purposes or to make modifications to their design which are usually very costly. Because of these circumstances and the fact that it is almost impossible to apply science to such advanced and complicated sheet extrusion processes without first closely examining and understanding the basic single screw extrusion process, the student was directed to conduct his experiments on a medium sized development machine and a large scale production compounding extruder.

Practical work on the development machine was restricted to the extrusion of polypropylene because observations on the behaviour of this material were of particular value to the Company at the time. Although polypropylene itself does not come within the scope of these studies, the opportunity was taken by the student to compare the practical performance of this polymer with theory because molten polypropylene resembles a Newtonian fluid more closely than any of the three plastics materials being studied. Practical work conducted on the production machine was restricted to PS and TPS materials for reasons of economy. The process of Evolutionary Operation (E.O.) was applied in the latter case with the intention of increasing plant productivity, and at the same time of obtaining valuable information on the interdependence of the more important extrusion variables. The method of E.O. ensures that there is the minimum large scale loss of production due to experimentation and the minimum accumulation of large quantities of scrap, but it suffers from the disadvantage that the accumulation of knowledge from a production machine is slow because modifications to the machine have to fit in with production plans and other development work. The use of a large extruder is advantageous in that the experimental data obtained is more directly applicable to large sheet producing machines than work conducted on a small development extruder.

Essential features of the single screw extrusion process including theories of delivery capacity and mixing are described. Where possible the practical work conducted by the student is related to theory and a detailed study is made of the interdependence of major extrusion variables, e.g. screw speed, back pressure, plastic melt temperature, output, power and heat input, and the degree of mixing. Only a small proportion (approximately 10%) of the observations recorded by the student are given in this report, but these are sufficient to demonstrate the important aspects of the single screw extrusion process studied, and other work which gave similar patterns and trends is indicated.

The general design features of single, twin and multi screw extruders are discussed in detail and special reference is made to recent trends in the design of devolatilising, and mixing and compounding extrusion equipment. Sheet die design is discussed in general terms only because of the limited research facilities available. However, with the assistance of the Technical Services Division at the Distillers Co. Ltd. Research Station, an Elliott 803 computer was used to help solve a complex problem on the selection of die sizes necessary to keep the production of edge trim scrap to a minimum, given a good sales forecast of the sizes and dimensions of future sheet orders. The work was reported in full elsewhere, but the essential features are described here.

The design of other extruder components such as ancillary sheet take-off equipment, drive motors, gear boxes, thrust bearings, feed hoppers, breaker plates, valves, heating and cooling equipment, and instrumentation are discussed, together with the problem of maintenance of extrusion equipment. Extrusion colouring techniques are considered and methods of making weighing and mixing techniques fully continuous are given. Problems which arise in the extrusion of TPS sheet materials on a commercial scale are tabulated together with possible causes and remedies.

Other activities which were undertaken by the candidate to obtain the background knowledge and information necessary to this work have been stated already in the preface to this report.

The nomenclature used in these sections to describe the extruder, the screw, and the extrusion process is clearly defined in "Nomenclature of the Screw Extruder" SPE Journal, p.295, March 1962, but an ancillary "Glossary of Terms" is given for the benefit of the reader (see Appendix 2). All mathematical terms used in the equations of these sections in connection with the science of extrusion are defined under "Mathematical Nomenclature", (Appendix 3) unless otherwise stated.

Many of the diagrams used in this thesis to illustrate various aspects of the extrusion process are reproductions from various technical papers, journals, and books. References are to be found at the end of the section to which they apply.

2. THE EARLY DEVELOPMENT OF EXTRUSION MACHINES

2.1. Pre 1900

In 1797 Joseph Bramah (1) invented a hand operated plunger and die for the production of lead piping. This is thought to be the first extrusion machine ever made and from this small beginning has grown one of the most widely used processes in the rubber and plastic industries.

The rubber industry in Germany and the U.K. (2,3) adapted the ram type extrusion process about 1845 for the purpose of insulating copper cable with gutta percha. This enabled the first submarine cable to be laid between Dover and Calais in 1851. The only other instance of the ram type extrusion machine being used commercially before 1850 was in the production of macaroni for the Italian food industry. Since 1850 the ram extrusion process has been used for shaping such diverse materials as soap, graphite, coal, ceramics, waxes, foodstuffs, bricks, metals, rubber and plastics. The extrusion process is presently used in many industries but the majority of extrusion machines are now designed for the plastics industry.

The British Xylonite Co. were the first to extrude plastics material around 1880, when they used a ram type extruder for the "wet" extrusion of cellulose nitrate (i.e. the material was softened with solvent).

The first Archimedean screw type extruder was patented in the U.K. by Gray (4) in 1879 and John Royle and Sons Ltd. in the U.S.A. also developed a single screw extruder about this time. Francis Shaw and Co.Ltd. in the U.K. has produced and sold screw extrusion machines by 1881 and Iddon Bros. Ltd. had also designed a screw extruder by 1886. The gear pump extrusion machine was invented in the following year in the U.K. by Willoughby Smith (5). This latter type of extrusion machine has since made thousands of miles of gutta percha insulated submarine cable for the Telegraph Construction and Maintenance Co.

2.2. Post 1900

In 1919 the "wet" extrusion of casein was begun by the British Xylonite Co. and Erinoid Ltd. and in 1927 these same two companies began the "wet" extrusion of cellulose acetate. Around 1930 experiments in the extrusion of polyvinyl chloride "MIPOLAM" were being investigated in Germany (6,7). The extruders used for these purposes were all short barrelled rubber type extruders and they were mainly steam heated.

Heidrich in Germany was the first to produce a screw extrusion machine which was electrically heated and specially designed for the dry extrusion of thermoplastic materials in 1931. In the following year three brothers named Horning left Germany to settle in Britain and they brought a number of Heidrich extrusion machines with them. Also in 1932 Rockhard Resins Ltd. who are now part of British Resin Products Ltd. successfully produced a special phenolic resin compound which was shaped into rollers for Gestetner duplicating machines, by the hydraulic ram extrusion process.

Francis Shaw in the U.K. and the Detroit Macoid Corporation and the National Rubber Machinery Co. in the U.S.A. had all made extruders specifically designed for use with dry plastics materials before 1940. These companies modified the established rubber extruders, by lengthening the screw and barrel and improving the electrical heat controls. About this same time twin screw extrusion machines were being developed by Roberto Colombo and Carlo Pasquetti in Italy.

The early war years brought a shortage of rubber in the U.K. and forced the cable makers to adopt the extrusion of plasticised PVC as a standard method of insulating cables. By 1941 cellulose nitrate, cellulose acetate, cellulose acetate butyrate, ethyl cellulose, polymethyl methacrylate, polyvinylidene chloride, polystyrene and polyethylene had all been extruded successfully in the U.S.A. It was during the 1940's that the extrusion of plastic materials really became established in the U.K., U.S.A. and Germany. The latter country in particular used both hydraulic ram and screw machines for the production of PVC pipe and PS foil during the war.

In the early 1950's many companies began the production of screw type extrusion machines for plastic materials. Over the past ten years rapid progress has been made in the design of extruders and screws. By the end of 1962 there were over 130 extruder manufacturers in the world, excluding those in the U.S.S.R., Eastern Europe and China.

2.3. References

1.	Anon	Plastics 18	p.404	December 1953
2.	Bewley	B.P. 10,825	1845	
3.	Brooman	B.P. 10,582	1845	
4.	Gray	B.P. 5,056	1879	
5.	Willoughby Smith	B.P. 17,768	1887	
6.	Fikentscher M. & Heuch C.	German P.654,989	1930	
7.	Fikentscher M. & Wolff W.	German P.669,793	1931	

3. THEORIES OF DELIVERY CAPACITY FOR SINGLE SCREW EXTRUDERS

3.1. Introduction

The extrusion process is a complicated one and as often happens in industry theory has lagged far behind practice. We are now in the transition period of transforming the art of extrusion into a science, and because of the complications the transition is likely to be a long one.

A single screw extruder melts and compresses thermoplastic materials before forcing them through a die which is designed to produce a continuous extrusion of desired shape. The heat required to melt the plastics granule feed is usually supplied by a combination of external heaters, and frictional heat produced by the shearing action of the screw on the plastic. Compression of the plastics granules is frequently achieved by decreasing the volume of the screw channel after the plastics material has entered the extruder. A combination of solid and fluid flows take place inside the screw channel and the temperature distribution within the system is very complicated.

Low capital investment, virtually continuous processing and a wide range of operation with different thermoplastic materials and section shapes, have all made extrusion an attractive manufacturing process, and account for its rapid growth. Because of the increasing importance of extrusion as a means of both compounding and fabricating plastics economically, considerable effort has been expended, particularly in the U.S.A., U.K. and Germany, towards the development of a more scientific approach to this problem.

We now have a better understanding of what happens in a screw extrusion machine by observation and scientific experimentation rather than by conjecture. However, because of the complexity of the extrusion process many individual opinions have been expressed in the past on extrusion problems without sufficient scientific basis, and thus any discussion on these particular topics has often led to controversy without contributing to a better understanding of the process.

The intensive effort made in recent years by raw material suppliers, machine manufacturers and custom extruders has greatly increased our knowledge of the extrusion process. These three separate sections of the industry have individually accumulated a wealth of knowledge but until recently there has been little integration of their work and this has severely hindered the rate of progress. Mathematically derived extruder flow equations now exist, but for simplicity they are usually based on assumptions which do not strictly apply in practice. In spite of this, many of the derived mathematical equations have been proved by numerous workers to compare well with the results obtained by experiment. Such theories are therefore useful tools to work with when designing a single screw extruder.

The field of screw design for a single screw extruder is a very large one and because of this the extrusion theories presented here will be specific to constant pitch type screws which are composed mainly of a feed section, a compression section, and a metering section, in that order. There are two major reasons why the extrusion theories will be confined to this type of screw design.

(a) It is the most popular type of screw in general use at present.

(b) Facilities available at BXP Ltd. only allowed the use of this type of screw.

Many of these extrusion theories, which are based on consideration of delivery capacity, power requirements and energy balance equation, have been carefully studied and are briefly explained. Important final equations are stated and the restrictive assumptions which limit the practical value of these theories, when applied to the extrusion of plastics materials are discussed. Space does not permit the full mathematical derivations of these theories in this report, however appropriate references to such detailed information is given in the text.

3.2. Theories Based On The Screw Feed Section

The feed section is of great importance because the output of a single screw extruder relies to a large extent on the quantity of plastic material in the granular state which this section conveys to the compression and metering sections. The conveying capacity of the feed section must be at least equal to the capacity of the fluid metering section on a weight basis.

Considering the importance of the feed section, relatively little is known about solids transport. Much more importance and development work will have to be devoted to this topic if we are to understand fully what happens in the feed section. Decker (1) was the first person to derive formulas for the flow of solid materials in a single screw extruder but these formulas do not relate well to actual behaviour. However, some of the conclusions reached by Decker are very relevant to the better conveyance of solid plastic materials and these are:-

1. Friction between the plastic and the screw must be as small as possible.

2. Friction between the plastic and the barrel must be as large as possible.

3. The lead and profile of the screw must be selected to give the largest axial components of movement and pressure.

4. The bulk density of the solids must be high.

3.2.1. Theory of plug flow of granules

Pawlowski (3) and Maillefer (4) both derived mathematical equations for the extrusion of dry powders. Maillefer recognised the problem as one of frictional differentials but his treatment was so simplified that he did not actually solve the problem for a helical channel where torques are involved. He simplified the problem to that of a straight channel and ignored the effect of the screw flight, consequently the derived equation only holds for very shallow screw channels. In practice the feed channels are usually relatively deep in order to extrude thermoplastic materials at economic rates.

The problem of solids conveying in a helical channel was investigated by Darnell and Mol (5) who took the work of Pawlowski and Maillefer a stage further and derived a "Theory of Plug Flow". The solid plug of small particles in the feed section of the screw is subjected to a pushing force normal to the flight surface when the screw is turning. Thus it will tend to move along a helical path normal to the flight if the only resisting forces are frictional forces on the barrel wall. If these frictional forces were infinite in the plane normal to the axis, the plug would move along the screw without turning at all just as a tightly held nut moves over a turning bolt. Friction, however, can oppose motion in any direction and the "nut" of particles moves off perpendicular to the flight surface which means that the "nut" must turn.

There are also frictional forces trying to drag the plug around with the turning screw. When this happens the forward component of movement of the plug is decreased and the delivery rate of the screw drops. In practice this effect is observed when the screw becomes over heated, the material then sticks to it and merely rides around with the screw with little forward motion.

Expressed in another way we can say that the output rate depends on the angle of movement of the solid plug. The angle of movement as derived by Darnell and Mol is therefore of the utmost importance (see Section 3.2.11).

Thus we see that for maximum forward movement of the plug the barrel surface should be rough and the screw to plastic contact should have a minimum coefficient of friction. Screw surfaces in the feed section therefore must be carefully machined and highly polished. Any uneveness which causes spasmodic sticking of the plastic to the screw could lead to serious surging troubles which if not evened out in the compression section may give a surging output from the metering section, to avoid this screws are usually chromium plated. It is important to remember that a constant feed rate of solids in the feed section of an extruder must be maintained if a constant output rate is required from the metering section at economic rates of extrusion.

In order to simplify the mathematical treatment of this theory Darnell and Mol (5) made the following assumptions:

1. The granular plastic in the screw channel behaves as an "elastic" plug and can have internal pressure. ~~This elastic~~ plug contacts all sides of the screw channel.

2. The pressure may be a constant but may also be a function of the channel length.

3. The coefficient of friction between the plastic and metal surface is independant of pressure and is constant over the entire length of the feed section.

4. No internal shear takes place but the material behaves as a solid plug.

5. The screw is single flighted in the feed section.

6. The effect of the width and radial clearance of the flight is negligible.

7. The screw is of constant pitch and the weight of the polymer stuck to the barrel and the screw is negligible.

For the case where the extruder screw and barrel are both machined to the same degree of smoothness, the coefficient of friction will be the same between the plastic and the surface of the screw or barrel provided the temperature of the screw and barrel are the same. For this case Darnell and Mol (5) derived the following equation.

(All terms in the following equations are defined in appendix 3).

$$\frac{Q}{N} = \pi^2 D\ h\ (D-h) \left[\frac{\tan\theta\ .\ \tan\phi}{\tan\theta + \tan\phi} \right] \qquad \ldots\ldots\ldots\ldots\ldots\ldots(1)$$

This equation although expressed differently is similar to the equation derived by Maillefer (4).

3.2.1.1. The angle of direction of movement in a helical channel

Darnell and Mol (5) derived a second equation which defines the angle of direction of movement (θ) of the solid plastic plug relative to a plane <u>perpendicular</u> to the helix angle of the screw. The following equation was derived in order that θ may be evaluated, this would then enable the delivery rate to be estimated from equation (1).

$$\cos\theta = K\sin\theta + C(K\sin\phi + C\cos\phi) + \frac{2h}{t}(KC\tan\phi + E^2) + \frac{hE}{L_a f_b}\sin\phi\,(E\cos\phi + K\sin\phi)\,\ln\frac{p_2}{p_1} \quad\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (2)$$

$$\text{where } K = E\,\frac{(\tan\phi + f_s)}{1 - f_s\tan\phi} \quad\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3)$$

$$C = \frac{D - 2h}{D} \quad\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (4)$$

The value of θ is usually obtained, however, by solving a shortened version of equation (2), where it is assumed that the polymer is not in intimate contact with the trailing surface of the screw flight. This situation implies no pressure build up in the solid plastic particles being conveyed and is the state which normally prevails in the first few turns of the feed section. The shortened version of equation (2) based on these assumptions is:

$$\cos\theta = K\sin\theta + C(K\sin\phi + C\cos\phi)\ldots\ldots\ldots\ldots\ldots\ldots\ldots (5)$$

N.B. In "Processing of Thermoplastic Materials (6) equation (2) has been misprinted. $k\sin\theta$ should read $K\sin\theta$.

3.2.1.2. Conditions which favour a high output

From equation (2) we see that the following conditions favour a large angle of movement which in turn favours a high output.

a. A deep screw channel. (h is large).

b. A low coefficient of friction on the screw. (f_s is low).

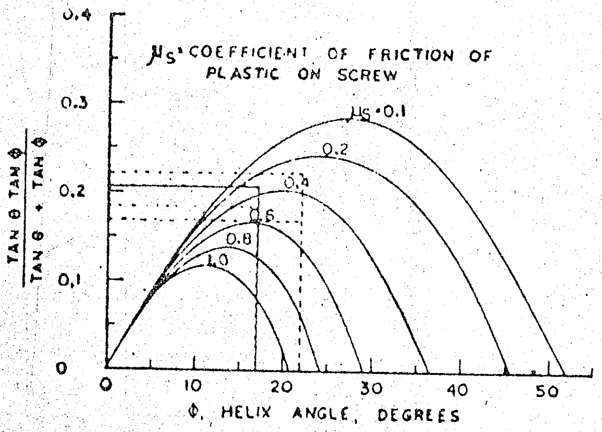
c. A high coefficient of friction on the barrel. (f_b is large).

d. A small helix angle. (low value of ϕ).

Although a small helix angle increases the angle of movement, it will not necessarily increase the volumetric output for a given screw speed because Q is a function of both ϕ and θ.

When ϕ is at a minimum θ is at a maximum and conversely. At both these extremes Q = 0. It is clear therefore that somewhere in between there is a value of ϕ where Q is a maximum, and this value of ϕ depends on the coefficient of friction on the screw.

Darnell and Mol showed in graph form (see fig. 3) the relationship between $\frac{\text{Tan}\,\theta \cdot \text{Tan}\,\phi}{\text{Tan}\,\theta + \text{Tan}\,\phi}$, the helix angle and the coefficient of friction of the plastic on the screw.



Feed screw tan function plotted against helix angle

Fig. 3.

The conclusions to be drawn from this relationship are:

1. As f_s decreases the optimum value of Q & ϕ increases.

2. For plastic materials with various values of f_s between 0.2 and 0.8 a high output is favoured by a relatively small helix angle and a relatively deep feed channel. The optimum helix angle for such a range of plastic materials is approximately that of lead equal to diameter, which is 17.7°. In practice this helix angle is now the one most used in screw design by extruder and screw manufacturers.

3.2.1.3. Special case where f_s is negligible.

In spite of the various assumptions listed earlier which help to simplify the derivations of equations (1) and (2), these equations are still complicated and often require further simplification before they can be used.

For instance in the special case where the coefficient of friction between the plastic and the screw is negligible and where there is no pressure build up, equation (1) approximately reduces to

$$Q = \pi^2 N D^2 h \sin\phi \cos\phi \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (6)$$

substituting $t = \pi D \tan \phi$ equation (6) becomes
$Q = \pi D N h t \cos^2 \phi$

Under these conditions the maximum theoretical volume output for plug flow at open discharge is almost exactly twice the theoretical drag flow of a liquid melt which can be sheared. Compare equation (6) with the drag flow term of equation (10) in the Simplified Flow Theory (see section3.4.1.1).

3.2.2. Comparison between theory and practice

Darnell and Mol extruded Nylon, Polyethylene and Acrylic resins through three different screws and three different barrel surfaces. Thus each material was run on each screw in turn under conditions of $f_b \cong f_s$, $f_b > f_s$ and $f_b < f_s$. The results obtained compared favourably with theory although for the extreme cases of the very rough and the highly polished barrel surfaces the results were respectively higher and lower than those predicted.

Considering the errors involved in determining the coefficients of friction, the bulk densities of the plastics resins and the variation of output with time, the derived equations gave surprisingly good agreement with the experimental results. It was found in practice that for good feeding of the plastic resins the barrel surface should be at least as rough if not rougher than the extruder screw surface.

3.2.3. Length of feed section

If the feed section is too short, insufficient thrust is produced on the (cold) feedstock material to push forward the sticky mass of material in the next (compression) section where it is beginning to flux, and the fluxing material tends to stick to the screw, leading to a low or surging throughput. In addition the conduction of work-heat back along the screw may cause premature fluxing in the feed section and thus reduce its effective thrust-producing length, especially if the rear of the screw is not water-cooled.

A further factor affected by the length of the feed section is the trapping of air, especially in fine powder feedstocks. Here an increased length often gives better results, probably by allowing the air to escape back along the screw.

For plastics materials such as CA and PS in pellet form it was considered normal to have a 1 or 2D feed section a few years ago with a 15D screw, but the present trend is towards 5 to 8D sections with 25D screws. For further details on the design of the feed section see sections 7.1.3.(a). and 7.1.4.(a).

3.2.4. Summary of feed section theories

Since the work of Darnell and Mol there has been no better theoretical interpretation of solids transport in the feed section of a single screw extruder. The theory is relatively sound but a more practical approach to the determination of frictional forces between plastics and various metal surfaces is badly needed. If reproducable friction coefficient results for various well defined metal surfaces could be obtained for different plastics materials over a wide temperature range, then this theory may enable us to design better screws for the extrusion of various plastic materials at specified temperature conditions.

This theory enables us to understand quite clearly the need for a compression section in screw design (see section 3.3.1.). It further shows us that a small helix angle, a smooth screw and a rough barrel are all very necessary if high output rates are desired from the feed section of a single screw extrusion machine.

3.3. Theories Based On The Screw Compression Or Transition Section

This is the only section of an extruder screw which has so far defeated any mathematically derived flow equation. The compression section of a screw is the zone in which the channel dimensions are reduced from those of the feed section to those of the metering section. It is in the compression section that the polymer theoretically changes from a solid to a fluid phase, this is not entirely true in practice because many feed sections supply a partially fluid composition, and metering sections often accept a partially solid mix. However, in the compression section the temperature of the polymer is raised by both conducted heat and frictional work heat and the main transition from solid to fluid melt takes place.

It is not difficult to imagine the complexity of even a simplified flow theory for a section in which the channel dimensions, the plastic temperature and the viscosity of the mix are continuously changing. Superimposed on this the movement and nature of the plastic mass changes in a complex manner from a tumbling and sliding action, through an alternate slipping and sticking action, to laminar shear flow. It is little wonder that a mathematical solution has not yet been supplied for this section and indeed it is doubtful if a useful complete solution will ever be found.

The work of Darnell and Mol has also shown us that the optimum helix angle for the conveyance of plastics solids in the feed section of an extruder is approximately that of lead equal to diameter. In recent years the trend in screw design for single screw extruders has been towards the use of this helix angle. Design trends have also reverted to screws of constant pitch throughout their whole length. The change from the feed depth to the metering depth which takes place in the compression section is normally a continuous one; stepped changes are not favoured in practice.

There are two important factors which govern the design of the compression section of a constant pitch screw. These are:

3.3.1. The compression ratio of the screw.
3.3.2. The length of the compression section.

3.3.1. Compression ratio of screw

It must be remembered that equations (6) and (10) refer to volumetric flow rate. In practice the bulk density of a solids feed is never equal to the liquid density, in fact it is seldom more than 0.5 - 0.7 of it. The maximum delivery rate of the feed section on a weight basis is thus about equal to the drag flow capacity for the melt if h is constant, and if there is absolutely no friction between the plastic and screw, and no back pressure.

In practice a slight increase in friction on the screw drastically reduces the solids feed delivery rate below the liquid drag flow capacity, although there is back pressure and a back flow effect which reduces the output from the metering section. A decrease in flight (volume) between the feed and metering sections helps to compensate for the former effect and ensures that the metering section is full of the plastic melt. If h in the metering section is half that of the feed section for a constant pitch screw, then the compression ratio of the screw is said to be 2:1. When designing a screw compression ratio it is therefore necessary to bear in mind the bulk density and ease of flow of the feedstock and the coefficient of friction for that material on the screw and barrel.

Screw compression ratios usually vary between 1.5 and 4.0. If the compression ratio is not sufficiently high the metering section of the screw is starved, when it is being operated at or near to open discharge. The machine output is then controlled by the feed section and surging may result because changes in bulk density and friction, due to small temperature variations, would immediately be reflected in the delivery rate.

3.3.2. Length of compression section

No theoretical method of finding the optimum length of the compression section for different plastic materials has yet been found. It is mainly from past experience and trial and error that we have come to realise that for hard thermoplastics with a sharp melting point such as 66 nylon, the transition section should be very short, approximately one turn. For polyethylene the transition section varies usually between one and five turns. For materials such as polystyrene and cellulose acetate the compression section may extend over as much as a third of the screw, and some screws designed for the extrusion of rigid PVC are composed almost entirely of a compression section. For further details on the design of the compression section see sections 7.1.3.(b) and 7.1.4.(c).

In conclusion it is true to say that the optimum design of the compression section can not be determined definitely and neither the theories of solid plug flow or melt extrusion hold true in this region.

3.4. The Metering Section

More flow theories have been based on this section than on any other aspect of the screw extrusion process. For convenience of mathematical derivation the metering section is considered, and normally is, a section of uniform geometry full of a plastic melt. The plastic material is probably a melt somewhere near the end of the compression section and this section could actually be considered as part of the metering section. Derived extruder flow equations for simplicity are usually limited to the laminar movement of a Newtonian fluid.

The derivation of the Navier - Stokes (7) equation in 1822 paved the way for a mathematical approach to flow theories for a single screw extruder by considering the problem of pressure drop and flow in tubes and pipes. Boussinesq (8) in 1868 calculated solutions of the Navier - Stokes equations for the pressure flow of viscous fluids in pipes of elliptical and rectangular cross section.

Rowell and Finlayson (9) solved the general equation for a rectangular groove with three fixed walls and one in motion. This solution described drag flow and leakage flow in an extruder in mathematical terms. Rowell and Finlayson (10) later calculated and drew representative curves for the isovels (lines connecting points of the same velocity) in the transverse direction of the groove of the screw for both the extreme conditions of maximum and zero output and also for several intermediate conditions. Similar work was conducted by Rogowsky (11) in the mid 1940's but the isovels he derived are not considered to be accurate.

Eirich (12) in 1947 considered the single screw extruder as a mechanism for developing high pressures and derived an expression for the maximum obtainable pressure. Pigott (13) experimented with oils and rubber on a single screw extruder around 1950 and made an attempt to account for the effect of shear rate on the viscosity of the material passing through the machine.

About this time Grant and Walker (14) investigated the velocity distributions in screw channels. They drew the profile of isovels for the zero output condition and gave a more reliable result than earlier workers, these results were later confirmed by Carley and Strub (15). Grant and Walker also drew attention to the importance of transverse flow and back pressure for the mixing of plastic materials and they outlined the differences between the practical and theoretical volumetric efficiences of single screw extruders.

Maillefer (16) in 1954 derived in full mathematical equations for use with polymer melts. He verified his general equation for output by experiments with oil and polythene and investigated the pressure drop along an extruder barrel at five different points.

3.4.1. Theories based on isothermal conditions

One of the major breakthroughs in this problem of predicting extruder performance was made in 1953 at a symposium presented by a team of Du Pont workers. The full mathematical derivation is given in a paper by Carley, Mallouk and Mc.Kelvey (17), and therefore only the important equations will be stated here.

The net rate of discharge from the extruder is equal to the algebraic sum of the drag flow, the pressure flow and the leakage flow. The latter two types of flow act against the drag flow and therefore the material balance may be given as:

$$Q = Q_D - Q_P - Q_L \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (7)$$

The derived Navier - Stokes equation for steady state flow of a Newtonian fluid in the rectangular channel of an extruder is:

$$\frac{dP}{dz} = \mu \left(\frac{d^2v}{dx^2} + \frac{d^2v}{dy^2} \right) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (8)$$

where z is the distance measured along the helical axis of the channel and v is the velocity of any point x. y. in the channel. This equation is the general equation of flow for Newtonian liquids and its derivation can be found in most books on the mechanics of viscous flow.

In order to simplify the mathematical solution of the above equation the problem can be reduced to one of two dimensions because the speed of the plastic melt along the screw channel in the direction z is constant and can neither increase nor decrease. Solutions of the duo-dimensional equation however are still difficult to manipulate mathematically although attempts to solve this equation have been made by Carley and Strub (15), Squires (18) and Jacobi (19) (see section 3.4.1.2.)

With square-cut screws having a flight depth less than 10% of the flight width it is claimed by Carley, Mallouk and McKelvey (17) that the x component on the velocity gradient can be ignored without incurring an error of more than 10%. In other words it is assumed that the effect of the channel walls on the velocity distribution is negligible and therefore the special case of infinite parallel plates can be applied. The problem thus becomes uni-dimensional.

In a deeply cut screw channel the melt is retarded in the vicinity of the flights so that the flow lines are deflected. In the uni-dimensional model, however, the only purpose of the flights is to guide the flow. Therefore, if the effect of the screw flights is ignored it is only necessary to determine the velocity gradient in the y direction which is the channel depth.

3.4.1.1. Uni-dimensional theory

The differential equation for uni-dimensional velocity distribution simplifies to

$$\frac{dP}{dz} = \mu \frac{d^2v}{dy^2} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (9)$$

Carley, Mallouk and McKelvey (17) first solved the problem assuming no leakage flow, they then derived a separate equation for leakage flow. They assumed isothermal conditions so that μ is constant and approached the problem by considering the case of a stationary screw and a rotating barrel. This in theory does not make any difference to the output of an extruder but does make the mathematical approach simpler. The velocity of the liquid relative to the screw was considered to be a maximum at the barrel surface and zero at the screw surface. The direction of motion is not, however, in a straight line because the screw channel is inclined at an angle ϕ. In their treatment of the flow rate in the screw channel the above authors split the velocity of the fluid into two components, one which acted directly down the channel (drag velocity) and the other which acted at right angles to it (transverse velocity). As stated in the definiton of this terms the transverse flow does not contribute to the forward movement of material but is a circulating flow which takes place between the flights. It contributes towards the mixing efficiency of the screw but does not enter into these calculations.

ASSUMPTIONS

The assumptions the du Pont team made in deriving their equation were thus

1. Isothermal conditions
2. The fluid is incompressible and Newtonian in behaviour.

3. The plastic flows in a streamline and lamina manner.
4. Only the metering section is considered and this is assumed to be completely filled with liquid melt.
5. The effect of the channel walls on the velocity distribution is negligible.
6. The radius of curvature at the base of the flight is ignored. The flight is assumed to be at an angle of 90° to the screw root.
7. The depth and the pitch of the metering section is constant.

DERIVATIONS

Bearing these assumptions in mind the solution of equation (7) is:-

$$Q = \frac{n \pi D h \left({}^{t}/n - e\right) \cos^{2} \phi}{2} \, N - \frac{n h^{3} \left({}^{t}/n - e\right) \sin \phi \cos \phi}{12 \, L \, \mu} \, \Delta P$$

$$- \frac{\pi^{2} D^{2} \delta^{3} \tan \phi}{10 \, \mu \, e \, L} \, \Delta P \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (10)$$

This equation is often expressed as:-

$$Q = \alpha N - \beta \left(\frac{\Delta P}{\mu}\right) - \gamma \left(\frac{\Delta P}{\mu}\right) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (11)$$

where α, β and γ are constants depending only on the geometry of the screw.

A full definition of all the above mathematical terms is given in appendix 3.

It is important to realise the significance of equation (10). To summarise, therefore, this equation states that for isothermal conditions the following laws apply.

1. The drag flow is independent of fluid viscosity and therefore applies equally well to Newtonian and non-Newtonian fluids.
2. The drag flow increases in direct proportion to:-
   a) the screw diameter
   b) the channel depth
   c) the speed of the screw rev/sec.
3. The pressure flow is affected by a change in viscosity.
4. Unlike drag flow the pressure flow is directly proportional to the cube of the channel depth and is therefore strongly dependent on the channel depth.
5. The leak flow term is small for a well-fitting screw, however, it is dependent on the cube of the distance between the top of the screw flights and the barrel. Therefore, this term

can reach significant proportions if the screw is badly worn.

6. The pressure flow and leak flow terms are both dependent on the axial length of the flighted section of the screw across which a pressure drop of Δ P takes place.

In practice, equation (10) has been found to be fairly accurate for shallow depth metering type screws for such materials as corn syrup and polyethylene, see papers by Carley (20), Gaspar (21), McKelvey (22) and Maddock (23). This theory therefore does agree with observed results for certain conditions. This theory also indicates the relevant points to watch in scale up work.

Carley, Mallouk and McKelvey (17) also derived equation for the volumetric output of a screw of varying channel dimensions. The latter equation is based on a solution of Δ P, but it will not be given here as it is thought to be only of academic interest. When a theory, based on Newtonian flow behaviour and isothermal conditions, is applied to a plastics material passing through a screw section of varying channel dimensions (e.g. screw compression section) then the results compared with practice became very inaccurate indeed.

The main reason for this is that the mechanical work heat evolved in this section make near-isothermal operation impossible and also the velocity distribution within the screw channel becomes very complicated.

Recently Jacobi (19) took the uni-dimensional flow equation a little further, by introducing a fourth term Q_4 which he calls expansion flow. This is derived to account for the fact that in practice the plastic melt expands with an increase in temperature and is compressed by the pressure generated in the screw channel. Jacobi illustrates this point for measurements made on polythene (24, 25) and polystyrene (26,27).

3.4.1.2. Duo-dimensional theory

As stated earlier the uni-dimensional theory is not sufficiently accurate for deep-cut screws, because the retarding action of the screw flights on the melt flow cannot be ignored.

Carley and Strub (15) were the first to derive a duo-dimensional theory. The differential equation which describes the duo-dimensional drag-flow in which the edge effects of the screw flights are considered is:-

$$\frac{d^2 v}{dx^2} + \frac{d^2 v}{dy^2} = 0 \quad \ldots\ldots (12)$$

and that for pressure flow is given by equation (8)

The mathematical solution of the duo-dimensional theory is very complex, however, the final equation arrived at by Carley and Strub (15) may be summarised as follows:

$$Q = F_d \propto N - F_p \; \beta \; \frac{\Delta P}{\mu} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (13)$$

where Fd and Fp are the drag-flow and pressure-flow shape factors respectively. These shape factors are complicated functions of the Fourier series involving channel depth to width ratio terms. Their full derivation is given in papers by Carley and Strub (15), Squires (18) and Jacobi (19).

Squires (18) obtained good agreement for Fd between theory and practice with polyisobutylene and polythene using screws with helix angles $\emptyset$ of 23.8° and 30° and channel depth to width ratios (h/w) from 0.04 to 0.955.

As yet no one has proved by experimentation the validity of the pressure flow shape factor Fp. Squires took the theory one stage further by considering the influence of channel curvature on drag flow. He derived the expression

$$Q = F_c \; F_d \propto N - F_p \; \frac{\beta}{\mu} \; \Delta P \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (14)$$

where Fc is the drag flow channel curvature factor. Squires (18) derived Fc and found it to be dependent only on the ratio of the channel depth to the outside screw diameter. He recognised the fact that such channel curvature effects exist for the extruder pressure flow term but did not derive them.

3.4.1.3. Other extensions of the uni and duo-dimensional flow theories

1. Mohr and Mallouk (28) have derived expressions for the transverse flow in the screw channel.
2. Theories have recently been derived for the extrusion of non-Newtonian melts.

In 1958 Mori and Matsumoto (29) gave the final flow equations for a Non-Newtonian Flow Theory without derivation. In 1960 Kennaway and Weeks (30) gave a method for designing screws based on such a theory but they also did not publish equations. In 1962 however, Glyde and Holmes - Walker (31) published in full a derivation for a Non-Newtonian Flow Theory.

3.4.2. Theories based on adiabatic conditions

In practice true adiabatic operation cannot be obtained although under certain conditions we can get very near to it. With true adiabatic extrusion all the heat is provided by internal friction in the melt and no heat is gained or lost through the walls of the flow channel. In practice "Autogenous Extrusion" and "Polytropic Extrusion" can be achieved (for definition of these terms see appendix 2).

However, the adiabatic theory is useful because it enables us to estimate the temperature rise of the plastic melt which occurs in the extruder, and it also leads to a better understanding of the extrusion process. Some of the energy supplied by the extruder motor to turn the screw is converted by a shearing action in the screw channel into heat energy and this increases the temperature of the plastic melt.

McKelvey(32) was the first person to derive an equation for the adiabatic extrusion process. He made similar assumptions of Newtonian Flow etc. to those used in the Isothermal uni-dimensional flow theory. McKelvey assumed that the energy necessary to produce flow could be ignored compared with the heat energy.

In order to calculate the temperature rise of the plastic melt in the axial direction McKelvey derived an energy balance equation. The fluid specific heat and viscosity must be known and a constant specific heat was assumed. The variation of viscosity with temperature was expressed as:-

$$\mu = \mu_i \, e^{-b(T - T_i)} \quad \ldots\ldots (15)$$

where T is the temp. °C and the subscript i refers to inlet conditions.

The increase in heat energy of the fluid as it passes through the metering section of the screw is equal to the sum of the power dissipated as shear energy in the screw channel and in the clearance over the flights.

The increase in heat energy of the fluid is given by

$$C_a \, Q \, \Delta T \quad \ldots\ldots (16)$$

Where C_a = Average specific heat of plastic melt expressed as in lbs/in^3/°C

Q = Extrusion rate - in^3/sec.

ΔT = Temp. rise in plastic - °C.

The sum of the power dissipated as shear energy in the screw channel and in the clearance over the flights was taken from power equations which are dealt with in the next section (see equ. (22)).

McKelvey (32) therefore derived the following equation:

$$C_a\, Q\, \Delta T = \frac{\pi^3 D^3 N^2 \mu L}{h} \left({}^{t}/n - e\right)\left[\cos^2\phi + 4\sin^2\phi\right] + \frac{n\,\pi^2 D^2 N^2 \mu_L\, e\, L}{\delta \tan\phi} \quad\ldots\ldots\ldots\ldots \tag{17}$$

The 4 $\sin^2\phi$ term is attributed to transverse flow and an appreciable error would be introduced if this were ignored.

μ_L = the viscosity in the land clearance; high temp. effects often cause this value to be lower than the viscosity of the melt in the screw channel proper.

Equation (17) then reduces to:

$$Q\, C_a\, \Delta T = E\, N^2 \mu\, L \quad\ldots\ldots\ldots\ldots \tag{18}$$

$$\text{where } E = \frac{\pi^2 D^2}{\tan\phi}\left[\frac{(t - ne)(1+3\sin^2\phi)}{h} + \frac{ne}{\delta}\right] \quad\ldots\ldots\ldots\ldots \tag{19}$$

In general the flow through any die (see section 8) may be expressed as:

$$Q = \frac{k\,\Delta P}{\mu} \quad\ldots\ldots\ldots\ldots \tag{20}$$

McKelvey (32) manipulated equations (14), (15), (18) and (20) and derived equations for the discharge rate, the pressure drop across the screw and the temperature increase of the extrudate in the metering section. The original paper should be consulted for the full derivation of these equations. A later paper by McKelvey and Bernhardt (33) shows good agreement between theory and experimental results for low density polyethylene.

Colwell (34) plotted the discharge rate against back pressure at different screw speeds for an extruder operating at approximately adiabatic conditions. He obtained fairly close agreement with the straight lines representing theoretical isothermal conditions at low back pressures but the discharge rate was reduced considerably at high pressures.

Jacobi (19) has recently derived complex equations which also take into account the effect of back pressure in adiabatic extrusion and also the quantity of heat lost in practice by heat radiation from the metallic extruder components.

Characteristics of Adiabatic Extrusion

There are two main advantages of supplying heat to the material in the screw flight by means of shear.

1. Energy is put in and heat is generated at the point where it is most required in the material, furthermore the amount of heat dissipated from the external barrel surface is less and therefore a higher thermal efficiency is achieved.

2. The thermal conductivity of plastics materials is low. Therefore a high temperature gradient is required across the plastic stream in order to make heat flow, by conduction methods. Consequently a more uniform temperature distribution is expected as adiabatic conditions are approached.

3.4.2.1. Autogenous extrusion*

In practice polythene, cellulose acetate, polypropylene, and polystyrene can all be extruded at near adiabatic conditions, although it is usually first necessary to heat the material and soften it so that the plastic is capable of being sheared. PVC will generate heat but care must be taken to avoid too much work heat being produced otherwise the PVC will degrade.

On the other hand, materials with a narrow softening range require most of the heat to be supplied by conduction methods through the barrel. For instance the amount of energy required to shear molten nylon is small and it is not sufficient to heat the oncoming plastic material.

The adiabatic extrusion theory therefore underlines the fact that screws should be designed from an energy basis as well as an output basis. For certain plastic materials the output is no longer limited by the fixed heat transfer area of the barrel. In the case of large extruders where the ratio of the area available for heat transfer to the extrusion rate is small it is particularly important to design the screw so that the heat supplied is predominantly mechanically generated.

In order to get the optimum balance of output rate and extrudate temperature it is often necessary to change simultaneously the screw speed and the die restriction or valve. This suffers from the disadvantage that such simultaneous changes make it difficult to adjust melt and flow rates accurately and rapidly.

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* For definition of Autogenous Extrusion see Glossary of Terms (appendix 2).

3.4.3. Theories based entirely on screw geometry

This theory was first derived by H.E.Buecken and it is given in detail in a book on extrusion by Simonds, Weith & Schack (35). It is based on the volume of the last flight of the metering section and can only be applied to an extruder operating under open discharge conditions because it does not account for the effect of back pressure or leak flow. This theory does however consider the effect of the volume taken up by the radius of curvature at the base of the flights.

Unfortunately the equation given in the above reference is incorrect; the corrected equation as derived by the author is given below. Volume extruded per rev. of screw

$$= \frac{\pi \quad (t-e)(R_2^2 - R_1^2) - 1.348\, R_1\, (r_1^2 + r_2^2) - 0.3011\, (r_1^3 + r_2^3)}{2} \ldots (21)$$

3.4.4. Power and heat energy theories

Rowell and Finlayson (9) were the first to tackle this problem of estimating the power requirements of a single screw extruder in 1928. Later Rogowsky (11) worked on the problem, but it was 1953 before a more complete and logical analysis of screw power requirements was available.

There are two main theories, the first is based on Isothermal operation and the second on Adiabatic operation.

3.4.4.1. Theories based on isothermal conditions

Mallouk and McKelvey (36) assumed

1. Isothermal conditions
2. Newtonian flow
3. Effect of channel flights can be neglected
4. The channel dimensions are constant

The derivation of the power equation was split into two parts. They first derived an expression for the power used to force by drag flow, the polymer melt through the screw channel against a particular back pressure. A second term was then derived for the power dissipated in the radial clearance between the tops of the screw flights and the barrel wall.

The sum of these two terms gives the total power. The equation they finally derived was:

$$Z = \frac{\pi^3 D^3 N^2 \mu L}{h} + \frac{Q_D \Delta P}{\cos^2 \phi} + \frac{\pi^2 D^2 N e \mu L}{\delta \tan \phi} \ldots\ldots\ldots\ldots\ldots\ldots (22)$$

This equation shows that the power consumption increases as the back pressure increases. The power dissipated in the radial clearance between screw flight and barrel may be two or three times as great as that dissipated in the main part of the screw channel, this causes an increase of the plastic temp. and consequently a decrease in the viscosity of the material in the radial clearance. The actual power consumed is therefore less than the calculated amount. In practice there is often a variation in viscosity along the length of the screw and both the flow and power equations are often expressed in terms of the average viscosity $\bar{\mu}$

$$\bar{\mu} = \frac{\int_0^L \mu \, d\lambda}{L} \qquad \dots\dots(23)$$

The viscosity of material under shear in the radial clearance between the screw flights and the barrel may be represented by μ_L .

3.4.4.2. Theories based on adiabatic conditions

There is a second approach to this problem given in detail by Squires (37). The total screw power is expressed as the sum of the power dissipated as viscous heat in both the screw channel and the radial clearance between the flights and the barrel, and the power required to raise and maintain the melt back pressure. Squires finally derived the following formula which includes the effect of transverse flow.

$$Z = \frac{\pi^3 D^3 N^2 \bar{u} L}{h} (t/n - e) \left[1 + 3 \sin^2 \phi\right] + Q_D \Delta P + \frac{n \pi^2 D^2 N^2 \mu_L e L}{\delta \tan \phi} \qquad \dots\dots(24)$$

Gore and McKelvey (38) have developed a modified form of equation (24) in which the effect of the flight faces of the screw channel on the power dissipated is also accounted for. Their results may be expressed as

$$Z = \frac{\pi^3 D^3 N^2 \bar{u} L}{h} (t/n - e) \left[F_z \cos^2 \phi + 4 \sin^2 \phi\right] + Q_D \Delta P + \frac{n \pi^2 D^2 N^2 \mu_L eL}{\delta \tan \phi} \qquad \dots\dots(25)$$

where F_z is a function of the depth to width ratio of the channel. For shallow channels, F_z approaches unity and equation (25) becomes equivalent to equation (24).

3.4.4.3. Energy Relations

If the effect of the energy required to raise and maintain the melt back pressure is ignored for an extruder operating under adiabatic conditions, then the energy required to raise the material temperature is given by equation (18) which may be expressed as

$$Z = Q\,C_a\,\Delta T = E\,N^2\mu\,L \ldots\ldots\ldots\ldots \tag{26}$$

If this energy is generated by frictional working the following relationship applies

$$Z = Q\,\mu\,\Gamma \ldots\ldots\ldots\ldots \tag{27}$$

where μ = melt viscosity of plastic lb - sec/in^2.
Γ = rate of shear sec^{-1}.

The shear rate at a particular point in a screw channel is given by

$$\Gamma = \frac{\pi D N}{h} \ldots\ldots\ldots\ldots \tag{28}$$

By equating equations (18), (27), and (28) we arrive at

$$\Delta T = \frac{\pi D N \mu}{C_a\,h} \ldots\ldots\ldots\ldots \tag{29}$$

The temperature rise due to frictional working for a given diameter extruder is thus a function of the screw speed, material viscosity and channel depth. In practice this relationship does not apply directly because it must be remembered that the viscosity of most plastic materials decreases with increasing rate of shear and increasing temperature in a complex manner. However, practical experience has shown that for autogenous extrusion a rise in the plastic temperature is obtained with any increase in screw speed or decrease in channel depth.

If now the energy required to build up and maintain the back pressure of an adiabatic extrusion process is also considered the derived equation for the total screw energy is

$$Z = Q\,C_a\,\Delta T + Q\,\Delta P \ldots\ldots\ldots\ldots \tag{30}$$

or

$$Z = E\,N^2\mu\,L + \alpha\,N\,\Delta P \ldots\ldots\ldots\ldots \tag{31}$$

3.4.5. Summary of metering section theories

An understanding of both the Isothermal and Adiabatic Flow Theories leads to a better understanding of the extrusion process. In practice neither isothermal nor adiabatic extrusion conditions can be attained although under certain conditions it is possible to get near to them.

The 'actual' output capacity of the single screw extruder is a complex quantity to derive. Many items have to be considered e.g. the available drive power, maximum screw speed, the back pressure caused by a die or some other resistance to flow, barrel heater capacity, thermal conductivity, thermal stability, melt viscosity, and screw geometry.

In practice heat is supplied mainly by conduction in the rear section of the screw where the channel depth is greatest and the plastic material has not yet melted. Frictional work heat gradually increases from the compression section onwards and it is usually predominant in the metering section of the screw where the channel depth is at a minimum. These two energy mechanisms of conducted heat and frictional work heat are closely related. The greater the energy supplied through the barrel to raise the plastic temperature the lower the energy required to turn the screw and vice versa. If possible it is usually desirable to supply frictional work heat via the screw rather than conducted heat through the barrel, because the former is supplied more accurately to the places where it is required and provides a more even form of heating. Screws should therefore be designed from an energy basis as well as an output basis.

In theory the output is proportional to screw speed and the power input is proportional to the square of the screw speed.

As the depth of the flight in the metering section is reduced the power necessary to turn the screw increases and therefore in practice one would expect autogenous extrusion* conditions to be reached at lower screw speeds for shallow flighted screws rather than for deep ones. Provided the depth of the feed section was kept constant a higher compression ratio screw would be expected to approach these conditions earlier than a low compression screw.

*For definition of Autogenous Extrusion see appendix 2.

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4. EXPERIMENTAL WORK ON DELIVERY CAPACITY THEORIES

4.1. Introduction

Theories of extrusion are normally based on Isothermal* or Adiabatic* extruder conditions. In practice extrusion is normally conducted under Autogenous* or Polytropic* operating conditions.

This section compares theories of volumetric flow and heat and power relationships with practical observations for a two inch extruder working under autogenous and polytropic conditions. This work was initially conducted for the B.X. Plastics Research and Development Department. The object of this work was to find methods of increasing the output of a 2 in. extruder equipped with a lay-flat film die, when extruding I.C.I. polypropylene, (Propathene grade LWF 31). Only one screw was available for use and the work was confined to a study of this particular grade of polypropylene. There were no facilities available to enable a separate study of the feed section to be made, and therefore this work was further confined to a study of melt flow and heat and power relationships.

In addition, a study of the inter-relationship between screw speed, output, back pressure and plastic temperature etc. was made for the autogenous extrusion case and an empirical equation was derived which fitted the observations.

4.1.1. Equipment used

A three stage, constant pitch single start screw was used, and a sketch showing the screw shape and dimensions is given in fig.4.

The diameter of the extruder bore was 2 in. The barrel was fitted with three thermocouples which recorded and controlled the barrel temperature in three separate zones which corresponded approximately to each of the three screw sections. A dynamometer developed by B.X.Plastics Ltd. was fitted to the point of final drive to the screw shaft which measured the torque necessary to turn the screw in lb ft. By changing the ratio of the drive pulleys and using a manually controlled variable speed drive motor a screw speed range of 15-60 rev/min was available, and a tachometer gave a direct reading of screw speed in rev/min.

A pressure transducer, melt thermocouple and valve were also available for use, a sketch showing the possible positions for this equipment is given in fig. 5. The heater band positions in relation to the thermocouple positions are shown in fig. 6.

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* for definitions of these terms see "Glossary of Terms"

SCREW DESIGN USED FOR POLYPROPYLENE EXTRUSION WORK

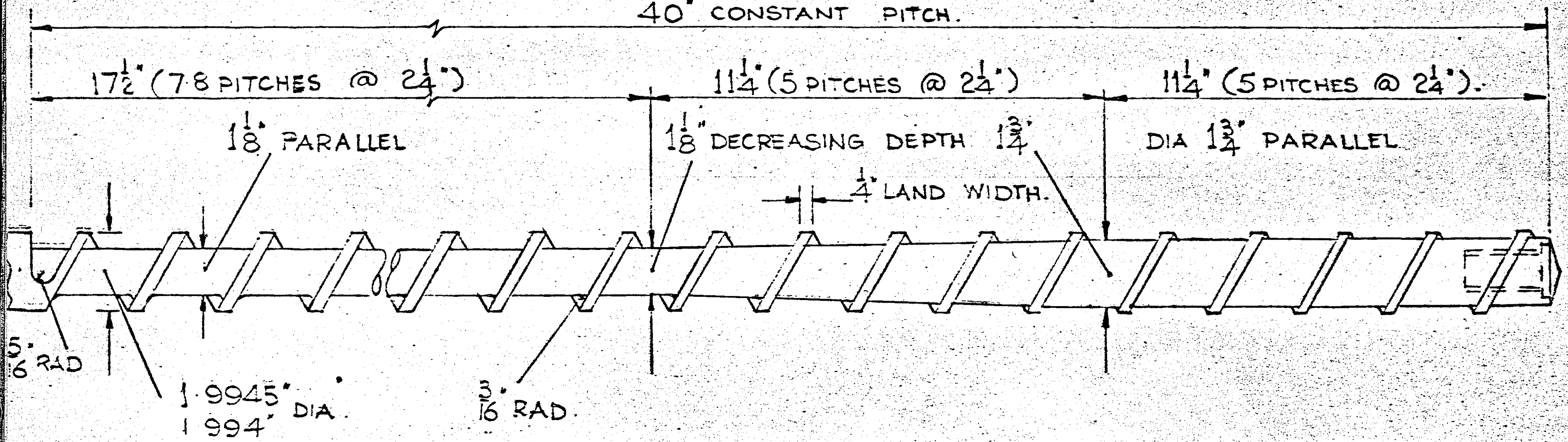


FIG. 4

Diagram of 2 inch Development Extruder
Used on Polypropylene Work.

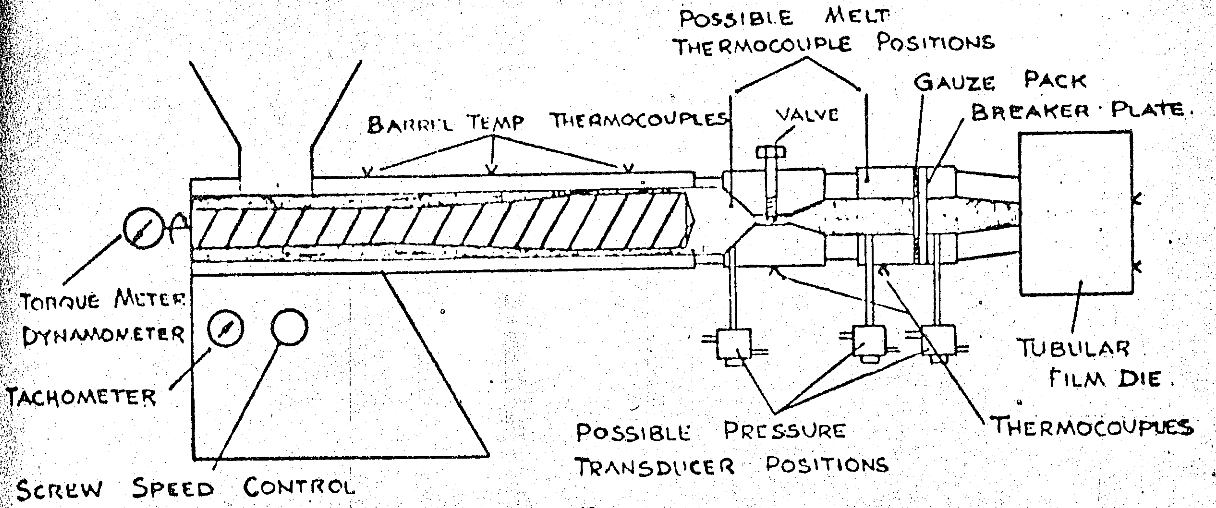


Fig.5

Diagram Showing Position of Heater
Bands on the 2 inch Machine.

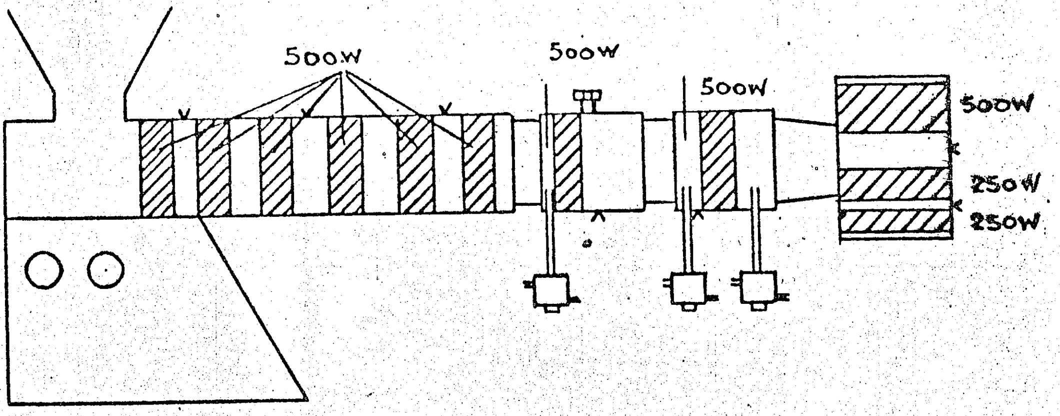


Fig.6

The extruder 'set-up' combinations used during this series of trials are listed below

A. Extruder only (open discharge case)
B. Extruder + breaker plate
C. Extruder + breaker plate and screen pack
D. Extruder + valve
E. Extruder + breaker plate and screen pack + die

The screen pack used for conditions C and E consisted of 1 x 60 mesh, 2 x 120 mesh and 2 x 200 mesh screens.

The screens used were of the following dimensions

Type (mesh)	Diameter of wire (inches)	Size of opening (ins)
60	0.0070	0.0097
120	0.0036	0.0052
200	0.0021	0.0029

Table 28

Die design will be dealt with more thoroughly in a later chapter (see section 8), but the effect of die temperature on output will be considered here.

For this series of experiments on the autogenous extrusion of polypropylene it took 5-10 min. for the extruder to return to the "steady-state" condition after either the rev/min or back pressure had been altered. In practice 15 min. were allowed before any of the following experimental observations were recorded.

4.2. Comparison Of Observations With Extrusion Theories

4.2.1. Isothermal flow theories

The uni-dimensional Simplified Flow Theory is given by equation (10).

$$Q = \frac{\pi n D h (t/_n - e) \cos^2 \phi N}{2} - \frac{n h^3 (t/_n - e) \sin \phi \cos \phi \Delta P}{12 L \mu} - \frac{\pi^2 D^2 \delta^3 \tan \phi \Delta P}{10 \mu e L} \qquad \ldots\ldots\ldots\ldots (10)$$

The theory based entirely on the screw geometry for the open discharge condition is given by equation (21).

$$Q = N \left[\frac{\pi (t - e)(R_2^2 - R_1^2) - 1.348\ R_1 (r_1^2 + r_2^2) - 0.3011 (r_1^3 + r_2^3)}{2} \right. \qquad \text{......................(21)}$$

The viscosity data used in equation (10) was derived from a graph supplied by I.C.I. for Propathene LWF 31 (see fig. 7). The derived value was based on a theoretical shear rate and the observed plastic melt temperature.

$$\text{Theoretical shear rate} = \frac{\text{Shear Stress (dynes/cm}^2)}{\text{Viscosity (poises)}} = \frac{\pi D N}{h}\ \text{sec}^{-1}$$

$$\frac{\text{Viscosity (poises)}}{68{,}947.} = \mu \text{ in lb/in}^2\text{/sec.}$$

Q is a volume expressed as in^3/sec., converted into lb/hr. by the following equation.

$$\frac{Q \times 16.39 \times 0.905 \times 3600}{453.6}\ \text{lb/hr.}$$

where 0.905 is the specific gravity of polypropylene L was assumed to be the length of the metering section = 11.25 in.

Viscosity Data for Polypropylene
(Grade Propathene LWF 31)

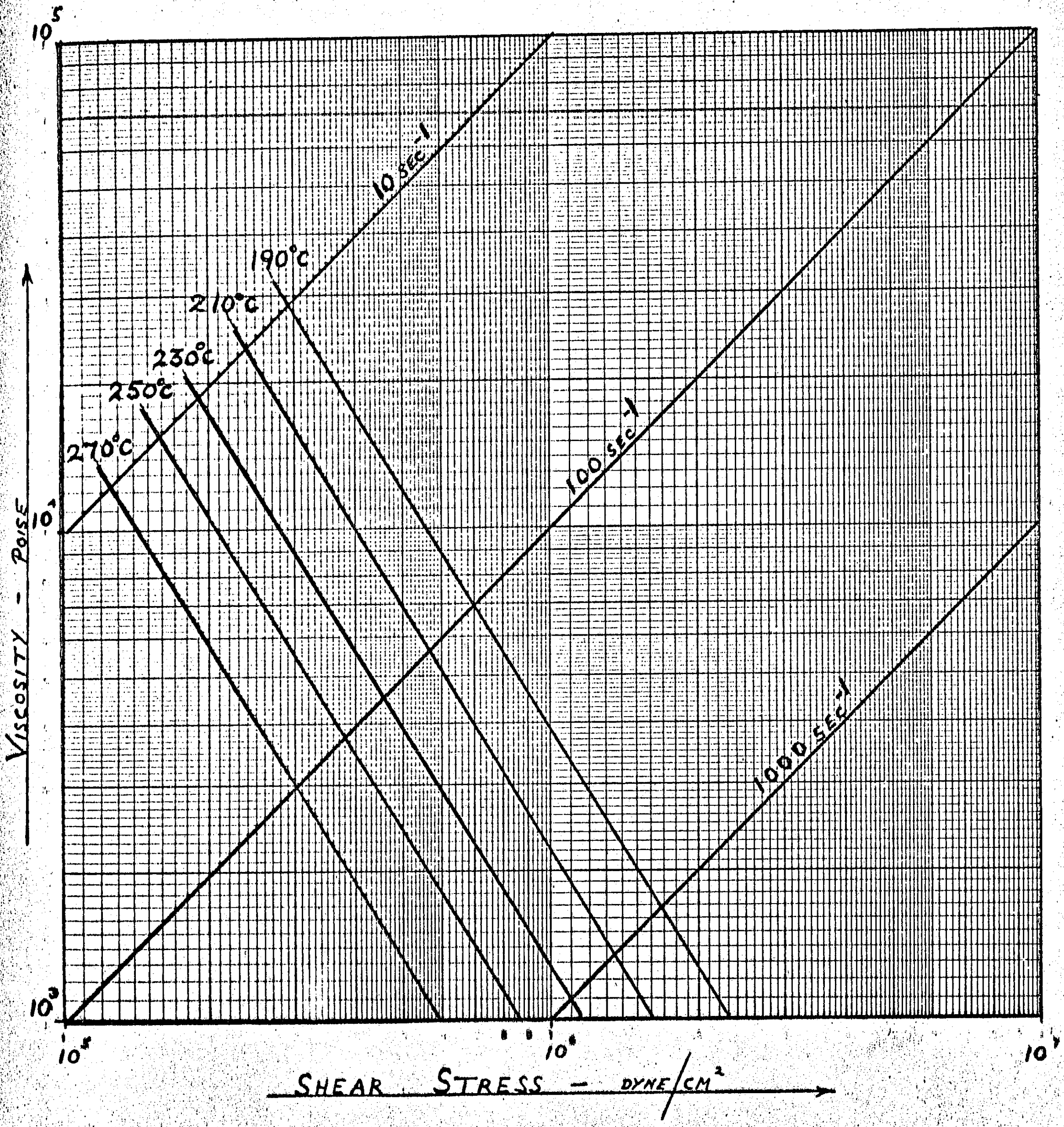


Fig.7.

4.2.1.1. Results

(a) Autogenous Extrusion Case

Extruder Operating Condition	Screw Speed rev/min	Torque lb ft.	Recorded Temps °C				Press lb/in^2	Shear Rate Sec^{-1}	μ lb in^2 sec.	Observed Output lb/hr	Theoretical Output	
			Barrel 1	Barrel 2	Barrel 3	Plast. Temp.					Equ.10 lb/hr	Equ.21 lb/hr
A	15	310	147	173	180	–	–	12.6	–	15.0	20.4	17.[illegible]
B	15	330	148	173	181	195	100	12.6	.35	14.6	20.1	
C	15	345	148	174	182	195	460	12.6	.35	13.9	18.9	
D	15	375	148	175	183	195	950	12.6	.35	11.1	17.1	
D	15	375	150	175	190	195	1320	12.6	.35	10.2	16.2	
D	15	380	150	177	195	197	2410	12.6	.35	8.3	12.8	
D	15	390	151	179	198	198	3260	12.6	.34	7.3	9.7	
A	25	320	152	180	189	–	–	21.0	–	20.8	34.0	28.9
B	25	340	152	182	189	198	150	21.0	.25	20.5	33.4	
C	25	365	154	182	190	199	520	21.0	.25	20.1	31.7	
D	25	390	155	182	190	199	1160	21.0	.25	17.6	27.8	
D	25	390	157	182	193	200	2280	21.0	.25	14.4	23.9	
D	25	395	160	183	199	205	3350	21.0	.23	11.5	18.0	
D	25	400	168	187	203	210	4560	21.0	.22	7.8	11.1	
A	35	325	152	185	189	–	–	29.3	–	26.4	47.6	40.4
B	35	345	152	186	191	207	200	29.3	.19	26.2	46.5	
C	35	380	154	187	192	208	590	29.3	.17	25.4	42.5	
D	35	395	156	188	195	210	1340	29.3	.17	22.2	40.7	
D	35	395	167	189	197	212	2820	29.3	.17	17.1	29.4	
D	35	395	175	192	205	216	4380	29.3	.17	12.6	19.2	
A	45	335	158	186	192	–	–	37.7	–	32.2	61.2	52.0
B	45	350	159	188	196	208	250	37.7	.16	31.9	59.5	
C	45	390	160	189	197	210	650	37.7	.15	31.0	56.5	
D	45	400	162	190	200	212	1500	37.7	.15	26.1	50.1	
D	45	400	170	192	205	215	2600	37.7	.15	22.3	42.0	
D	45	400	175	193	208	218	3800	37.7	.14	18.2	31.3	
A	55	340	166	189	197	–	–	46.1	–	37.7	74.8	63.6
B	55	350	166	190	198	210	280	46.1	.13	36.7	72.5	
C	55	395	166	192	200	212	700	46.1	.13	35.1	68.9	
D	55	405	168	193	204	215	1400	46.1	.13	32.6	62.9	
D	55	410	170	194	206	217	2850	46.1	.12	26.7	48.5	
D	55	410	176	196	210	220	3960	46.1	.12	22.2	38.3	

Table 29

The temperatures of the breaker plate assembly and valve were preset at 200°C and they were controlled to within ± 2°C.

(b) Polytropic extrusion case

Extruder Operating Condition B	Screw Speed rev/min	Torque lb ft.	Recorded Temps. °C					Press lb/in^2	Observed Output lb/hr	Theoretical Output Equ.10 lb/hr
			Barrel			Breaker Plate	Plast Temp.			
			1	2	3					
Low barrel temps.	15	430	121	175	190	197	197	500	12.4	18.9
	25	450	130	178	190	197	196	520	17.5	31.7
	35	390	141	181	191	196	190	560	21.2	44.8
High barrel Temps.	15	145	211	207	218	217	213	500	16.5	18.6
	25	200	205	215	217	224	220	560	27.9	30.9
	35	240	202	212	210	222	213	590	38.6	43.8
Cooling the feed zone at constant rev/min	25	200	205	207	218	224	220	560	27.9	30.9
	25	210	180	211	214	222	221	540	27.3	31.0
	25	210	160	211	214	222	222	520	26.6	31.2
	25	230	140	211	214	222	222	510	25.3	31.2
	25	250	130	206	215	222	223	500	24.6	31.3
	25	260	120	204	215	222	223	500	24.3	31.3
	25	260	110	204	215	222	222	500	24.1	31.3

Table 30

4.2.1.2. Discussion of results

The best agreement between theory and practice was obtained for the polytropic extrusion case of high barrel temperatures. This was to be expected because both the high barrel temperature and low torque readings signify that much less work heat (approx. a half) was being developed by the screw when compared to autogenous extrusion, or polytropic extrusion at low barrel temperatures. High barrel temperatures are known to bring us nearer to the isothermal state, while autogenous extrusion is the nearest we can get to adiabatic extrusion in practice.

(a) Autogenous extrusion case

Agreement between the isothermal uni-dimensional simplified flow equation and the observed output varies for different values of screw speed and back pressures. One reason for this is because L was considered to be a constant length, but in practice we know that the length of screw full of plastics melt (L) must increase as the back pressure increases also the value of L will vary with screw speed. Unfortunately facilities were not available for measuring L, which would entail boring small holes through the barrel at regular intervals down its length, the last hole to leak the plastic melt would then give an approximate value for L.

The discrepancy between the observed and theoretical output figures given in table 29 varies between approximately 25% to 50%. This is not surprising when it is realised that the data available for L and μ is very approximate. Further we know that autogenous extrusion is nearer to adiabatic rather than isothermal conditions. In the case of autogenous extrusion the temperature gradient across the metering section and the heat losses via the screw and barrel become greater as the screw speed is increased. Further the drag flow and pressure flow shape factors and channel curvature factors of the duo-dimensional isothermal melt flow theory have not been accounted for. Also, as the plastic temperature and screw temperature increases, the coefficient of friction between the plastic and the screw increases, and there is a greater tendency for the plastic to stick to the screw in the feed zone, thus causing a reduction in output.

These facts help to account for the discrepancy between theoretical and actual outputs in the autogenous extrusion case. The theoretical output derived for the open discharge condition using equation (21) shows better agreement with the observed output than the simplified flow equation, because the former accounts for the volume taken up by the radius of curvature at the base of the screw flights. Equation (21) however, does not take into account the drop in output caused by a restriction in the flow path and therefore it can only be applied to the open discharge case.

(b) Polytropic extrusion case

Considering the approximate nature of the uni-dimensional simplified flow equation and the μ and L data, it is surprising that such a good agreement between the theoretical and observed output of polypropylene melt was obtained for the polytropic extrusion case of high barrel temperature, although these operating conditions are known to approach isothermal conditions.

It is noticeable that under these operating conditions the temperature of the feed zone and the barrel greatly effect the output of the extruder, high outputs being favoured by high barrel temperatures. Similar results were obtained when the extruder was equipped with a lay-flat tube die, but the output dropped considerably when the feed zone temperature rose above 220°C, presumably because the plastic material began sticking to the screw surface and slipping over the barrel surface, thus effectively starving the feed.

The observations further show that the energy required to drive the screw depends greatly on the heat energy supplied through the barrel wall. These results therefore verify many of the qualitative conclusions formed on this subject of heat energy under the heading "Theories of Delivery Capacity for Single Screw Extruders".

It must however be remembered that although high barrel temperatures often give rise to greater outputs, they also lead to poorer mixing of the plastic melt because there is less shear taking place in the screw channel. In practice it is usually necessary to obtain a balance between mixing efficiency and output.

The temperature, particle size and lubrication qualities of the feed stock as well as the barrel and screw zone temperatures, are all very important factors which have a considerable bearing on the output obtained from an extruder. They are among the most important and yet least understood of all extrusion problems, and constitute one of the largest gaps in our knowledge of the extrusion process. Provided the necessary facilities are made available to study the feed section as a separate entity, it is strongly recommended that some future studies by other workers should be directed along this course.

4.2.2. Adiabatic flow theories

In practice autogenous extrusion conditions resemble the adiabatic state more closely than polytropic conditions and therefore only the former results are studied here. The adiabatic flow theory is given by equation (18)

$$Q\,C_a \Delta T = \frac{\pi^2 D^2}{\tan \phi}\left[\frac{(t - ne)\,(1 + 3\sin^2\phi)}{h} + \frac{n\,e}{\delta}\right] N^2 \mu L \;\ldots\ldots(18)$$

for Autogenous extrusion conditions this equation should be

$$Q\,C_a \Delta T + M = \frac{\pi^2 D^2}{\tan \phi}\left[\frac{(t - ne)(1 + 3\sin^2\phi)}{h}\,\mu + \frac{ne}{\delta}\,\bar{\mu}\right] N^2 L \;\ldots\ldots(32)$$

where M = heat lost from system by the normal methods of conduction, convection and radiation.

μ = average viscosity in screw channel

$\bar{\mu}$ = average viscosity in land clearance (considerably lower than μ).

This equation may be applied to find the temperature rise over a screw section of constant dimensions and therefore it is theoretically possible to find the plastic temperature rise over the metering section because ΔT is the only unknown in equation (32).

Assuming no heat energy leaves the system ΔT is given by

$$\Delta T = \frac{\pi^2 D^2 N^2 \mu L}{C_a Q \tan \phi} \left[\frac{(t - ne)(1 + 3 \sin^2 \phi)}{h} + \frac{n e}{\delta} \right] \ldots\ldots\ldots\ldots (33)$$

where L = length of metering section 11.25 in

C_a = average specific heat of polypropylene 0.46 B.T.U./lb/$^{\circ}$F or 230 in lbs/in^3/$^{\circ}$C

μ and $\bar{\mu}$ are based on the plastic temperature of the polypropylene melt at the end of the metering section.

Q = Observed output in ^{3}/sec.

N = Observed screw speed rev/sec.

All dimensions used in these equations are given in the introduction to this section.

4.2.2.1. Results for autogenous extrusion conditions

Extruder Operating Condition	Observed Screw Speed rev/min	Plastic Temp ^{o}C	μ lb in^{2}sec	Back Pressure lb/in^{2}.	Observed Output lb/hr.	Calculated ΔT ^{o}C
B	15	195	0.35	100	14.6	59
C	15	195	0.35	460	13.9	62
D	15	195	0.35	950	11.1	78
D	15	195	0.35	1320	10.2	85
B	25	198	0.25	150	20.5	84
C	25	199	0.25	520	20.1	85
D	25	199	0.25	1160	17.6	97
D	25	200	0.25	2280	14.4	119
B	35	207	0.18	200	26.2	88
C	35	208	0.17	590	25.4	90
D	35	210	0.17	1340	22.2	103
D	35	212	0.17	2820	17.1	134
B	45	208	0.16	250	31.9	108
C	45	210	0.15	650	31.0	112
D	45	212	0.15	1500	26.1	128
D	45	215	0.15	2600	22.3	150
B	55	210	0.13	280	36.7	118
C	55	212	0.13	700	35.1	123
D	55	215	0.13	1400	32.6	133
D	55	217	0.13	2850	26.7	161

Table 31

4.2.2.2. Discussion of results

These results for the theoretical plastic temperature rise over the metering section of the screw are approximately up to 3 times greater than the values expected. The reasons for this inaccuracy are:

1. The heat energy (M) lost by conduction in the metering section of the screw and barrel was considerable.

2. The viscosity data available was inaccurate e.g. if $\bar{\mu}$ was known for each set of operating conditions, the second term inside the brackets of equation (18) would be considerably smaller, particularly at high screw speeds.

3. The screw and the barrel surfaces were both slightly worn and therefore the land clearance is probably larger than the 'blueprint' figure of 0.003 inches. The land clearance distance will vary slightly down the length of the metering section and in this equation its value is very important e.g. if δ was doubled ΔT would be reduced by approximately 40%.

However, these results do reveal two very important facts:

a. Raising the screw speed and thus the rate of shear of the plastic material, increases the value of ΔT.

b. Raising the back pressure increases the residence time of the plastic material in the metering section, and this also increases the value of ΔT.

4.2.3. Power and heat relationships

Equation (30) gives an energy balance relationship for an adiabatic extrusion process.

Equation (30) is:

$$Z = Q\,C_a\,\Delta T + Q\,\Delta P$$

where Z is the total power input to the screw

$$= \frac{2\,\pi \times \text{r.p.m.} \times \text{torque} \times 6{,}600}{33{,}000} \text{ in. lb/sec.}$$

C_a is the average specific heat of polypropylene

$= 0.46$ B.T.U./lb/°F or 230 in lbs/in^3/°C

ΔT is the difference between the initial and the final plastics temperature = (plastic temperature - 20°C).

The $Q\,C_a\Delta T$ term is the energy required to raise the temperature. The $Q\Delta P$ term is the energy required to displace the plastics material against the back pressure and thus produce forward flow.

For autogenous extrusion the equation should be

$$Z = Q\,C_a\,\Delta T + Q\,\Delta P + M \quad\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots(34)$$

where M is the energy lost by the conduction, convection and radiation of heat from the system, but this is not known.

The theoretical output of plastic material and the apparent energy lost from the system were calculated using equation (30) and equation (34) respectively.

4.2.3.1. Results for "autogenous" extrusion conditions

Extruder Operating Condition	Screw Speed rev/min	Torque lb ft.	ΔT °C	ΔP lb/in^2	Observed Output lb/hr	Calculated Out/put lb/hr	Total Energy Supplied to system (Z)H.P.	Apparent Energy lost from system (M)H.P.
B	15	330	175	100	14.6	18.2	0.94	0.19
C	15	345	175	460	13.9	18.9	0.99	0.26
D	15	375	175	950	11.1	20.2	1.07	0.48
B	25	340	178	150	20.5	30.6	1.61	0.54
C	25	365	179	520	20.1	32.4	1.74	0.66
D	25	390	179	1160	17.6	34.1	1.86	0.90
B	35	345	187	200	26.2	40.8	2.30	0.85
C	35	380	188	590	25.4	44.3	2.52	1.11
D	35	395	190	1340	22.2	44.8	2.63	1.35
B	45	350	188	250	31.9	52.9	3.00	1.23
C	45	390	190	650	31.0	57.8	3.34	1.58
D	45	400	192	1500	26.1	57.6	3.43	1.90
B	55	350	190	280	36.7	63.9	3.67	1.59
C	55	395	192	700	35.1	70.7	4.14	2.12
D	55	405	195	1400	32.6	70.4	4.24	2.30

Table 32

4.2.3.2. Discussion of results

Apart from a slight discrepancy at 15 rev/min the calculated results for the apparent energy lost from the system at differing screw speeds and back pressures are in a regular pattern. These results indicate that the apparent energy lost from the system is strongly dependent on the screw speed and back pressure. This would be expected since it has been established from theoretical considerations in the previous section and also from experimental observations shown above, that ΔT tends to rise as the screw speed and/or back pressure are increased. The heat energy lost, mainly by conduction through the back of the screw and barrel must also increase as ΔT rises.

A comparison between the calculated and observed outputs indicates that up to half the total energy supplied to turn the screw, for the extrusion of polypropylene under these conditions, is lost from the system. This conclusion is not surprising when one considers the heat energy lost via the screw and barrel.

Higher back pressures up to 4,000 lb/in^2 make very little difference to the torque readings (see table 29) and therefore any further increase in the apparent energy lost from the system is caused by a reduction of Q, although the ΔT and ΔP terms become more prominent at these high back pressures.

In the case of true adiabatic extrusion M would be zero.

It is interesting to note that for back pressures below 300 lb/in^2 the calculated outputs for a screw speed range of 15-55 rev/min are in agreement with the theoretical figures derived using the simplified flow equation, compare with results given in table 29.

4.3. Derivation Of Empirical Extrusion Equations

4.3.1. Interrelationship between pressure, output and screw speed

This series of experiments was also designed to show up any interrelationships which may exist between the major variables in the autogenous extrusion of polypropylene. The results, given in table 29 were obtained by varying the screw speed and/or back pressure and observing the effect this had on the other extrusion variables.

On analysis these results showed that a linear relationship existed between back pressure and observed output at various fixed screw speeds (see fig. 8). This linear relationship holds equally well for the calculated output results based on the uni-dimensional simplified flow theory. At low screw speeds (i.e. 15 rev/min) it will be noticed that the relationship is not quite linear.

These results have also been expressed in three dimensional form, see fig. 9. The outer surfaces of this three dimensional plot are based on the results obtained when the extruder was operating under conditions A and D, see table 29. Other authors (1,2) have obtained similar relationships for polythene and polystyrene but they have not expressed their result in this three dimensional way. This type of graph (fig. 9) has considerable value in practice because provided two of the three variables are known the third may be predicted. Also provided this type of planar relationship can reasonably be assumed for future experiments of this type, it will only be necessary to make six carefully chosen observations, and

Graph Showing Relationship Between Output, Back Pressure and Screw Speed for Polypropylene (Grade Propathene LWF 31).

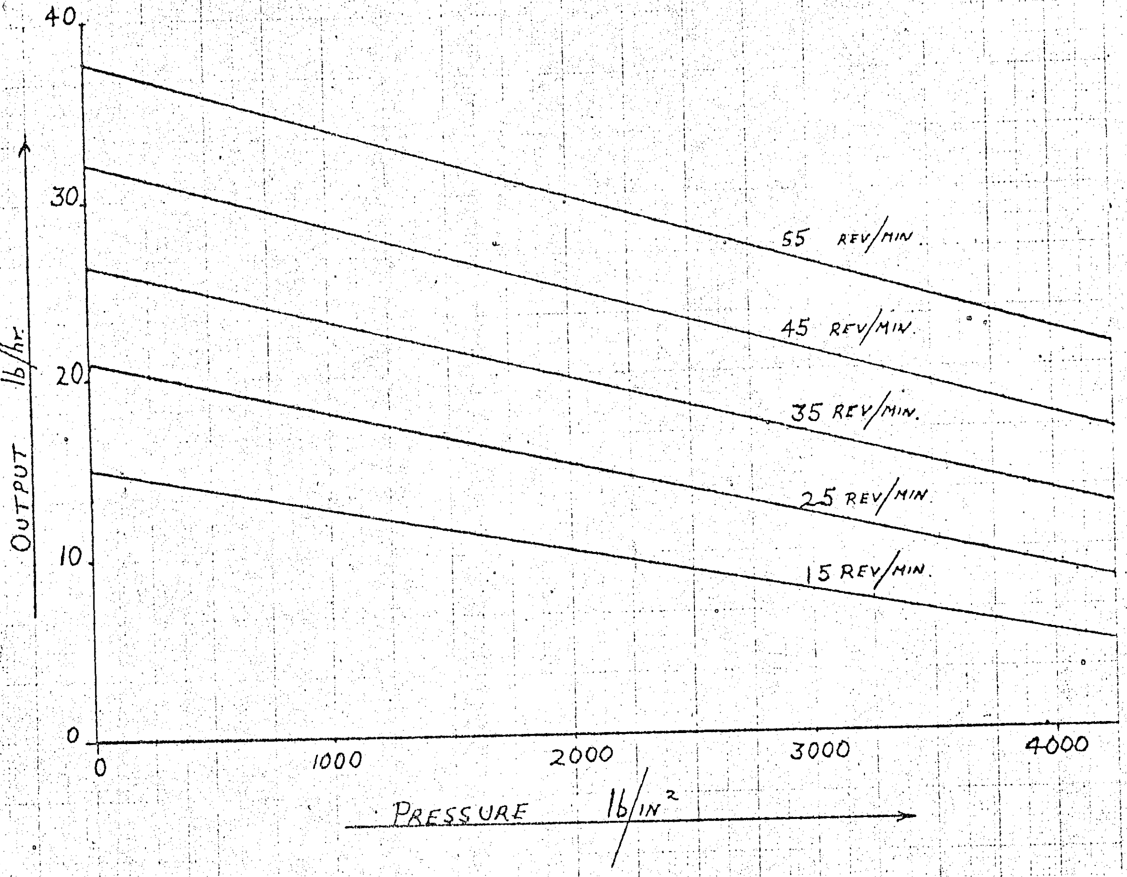


Fig. 8.

Inter-relationship Between Screw Speed, Back Pressure and Output for the Extrusion of Polypropylene.

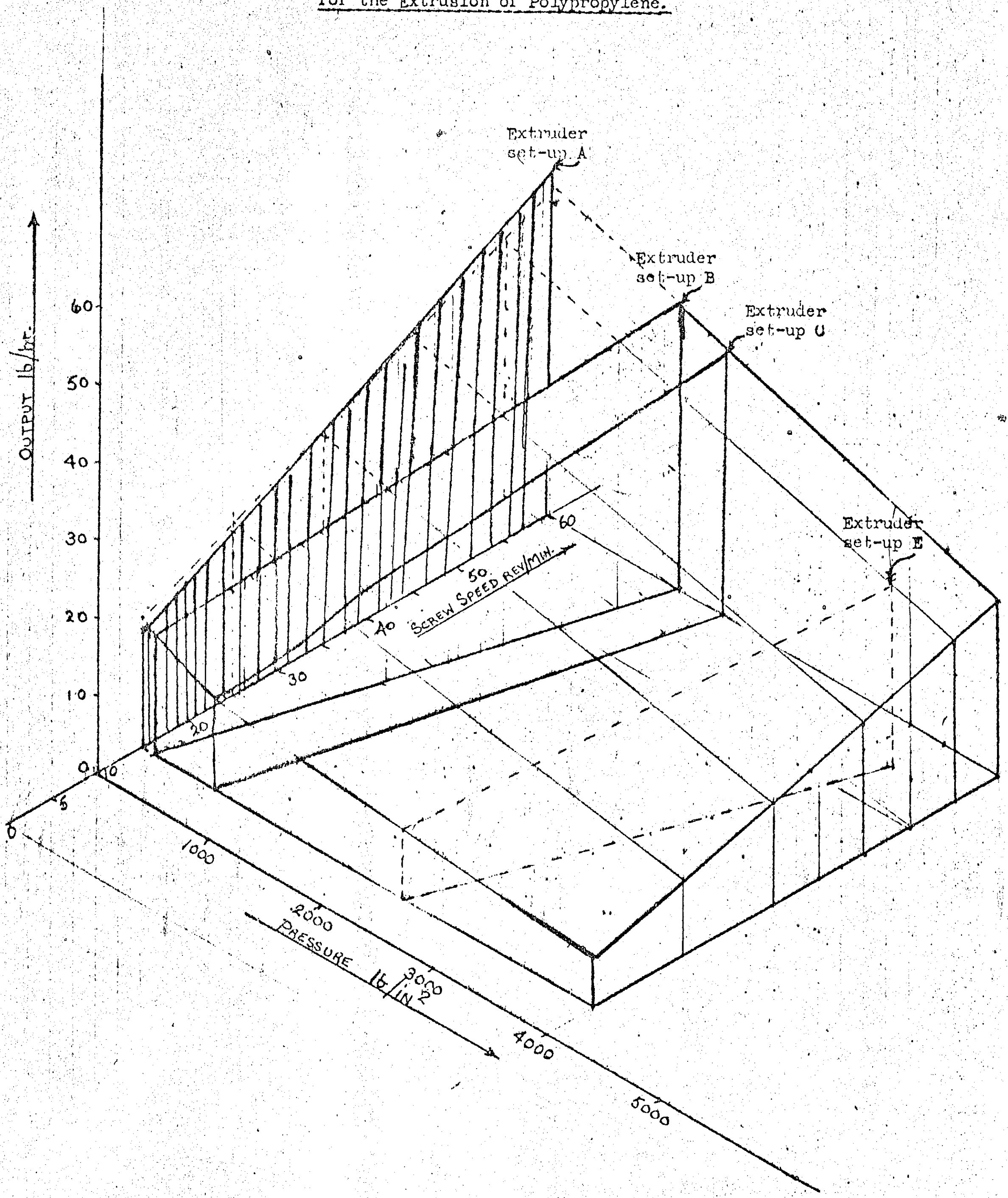


Fig.9

all intermediate operating conditions could then be clearly defined.

4.3.2. Empirical extrusion equations

An Elliott 803 computer was used to conduct a number of arithmetic calculations which verified that the surface of the three dimensional plot (fig. 9) was in fact planar, and an attempt was made to fit an equation to this plane.

A brief outline of the method used is as follows:

If y = output in lb/hr.
x_1 = screw speed in rev/min
x_2 = recorded pressure in lb/in^2.

A linear relationship was suspected between y, x_1 and x_2, thought to be of the form:

$$y = b_o + b_1 x_1 + b_2 x_2 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (35)$$

where b_o b_1 and b_2 are numerical constants, y is a dependent variable, and x_1 and x_2 are independent variables (i.e. controllable).

It was necessary to find b_o b_1 and b_2 such that equation (35) fits the observations. This was achieved by using the method of least squares; as the name implies, values of b_o b_1 and b_2 are found such that the sums of the squares of the distances of the observations from the plane are minimum.

These distances are:-

$$y_j - b_o - b_1 x_{1j} - b_2 x_{2j} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (36)$$

The sum of the squares

$$= \sum_{j}^{N} (y_j - b_o - b_1 x_{1j} - b_2 x_{2j})^2 \quad \ldots\ldots\ldots\ldots\ldots\ldots (37)$$

where N = the number of observations and

$$j = 1, 2, 3 \ldots\ldots\ldots\ldots\ldots N$$

An Elliott 803 computer was programmed by P. Chandler (then of B.X. Plastics Ltd.) and his results show that the observations listed in table 29 give a well fitting linear relationship between y and x_1 and x_2 over the area covered by the observed values of x_1 and x_2.

This relationship was

$$y = 5.9343 + 0.5872\ x_1 - 0.0035\ x_2 \ldots\ldots\ldots\ldots\ldots\ldots\ldots (38)$$

It was further shown that a consideration of higher order terms in x_1 and x_2 (e.g. $x_1^2\ x_2^2$ etc) does NOT lead to a function which fits the observations significantly better.

From equation (38) we see that when the screw speed and back pressure are zero, the output is 5.93 lb/hr. This is of course impossible and could arise only because no observations were made at low values. Obviously the true relationship must alter rapidly near the origin and extrapolation from the area of observation in this direction should not be carried out.

Similar work to this was conducted by Donald, Rieke and Humes (1). From their experimental results they derived the following empirical relationship.

$$Q = a + bN + cPN + dP \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (39)$$

where a, b, c and d are constants.

It would seem therefore that important aspects of the extrusion process can, under these circumstances, be described by experimentally determined equations.

4.3.3. Interrelationship between pressure, output, screw speed and plastic temperature.

A further study of the observations in table 29 shows that the plastics temperature is positively correlated with the screw speed and the recorded pressure i.e. as the screw speed or back pressure increases, the melt temperature increases. However, the melt temperature does not appear to have any direct relationship with the output, but will be correlated with it in as much as it is correlated with screw speed and pressure.

4.3.4. Effect of breaker plate and screen pack

It should be noted that the observations recorded for extruder operations B and C as defined in the supplementary introduction to this section exactly fit the three dimensional grid which relates the back pressure and screw speed to output (see fig.9). Further these planes representing conditions B and C are diagonally situated because the back pressure rises as the screw rev/min is increased.

It is also interesting to note that the plane of observations for different back pressures enabled the output at open discharge to be accurately predicted before measurements of extruder operation A were conducted.

4.3.5. Effect of the die on output

The following results were obtained for the 2" extruder operating under conditions E. (i.e. extruder + breaker plate + screen pack + die).

The pressure transducer was situated on the screw side of the breaker plate assembly and the screen pack consisted of 1 x 60 mesh, 2 x 120 mesh and 2 x 200 mesh gauzes. The extruder was operated under autogenous conditions but the temperatures of the breaker plate assembly and die were preset at 200°C.

Results

Screw Speed rev/min	Torque lb ft.	Recorded Temperatures °C							Press. lb/in^2	Output lb/hr
		Barrel			Breaker Plate Ass.	Die		Plastic Temp.		
		1	2	3		1	2			
15	380	149	179	185	200	200	200	197	2320	8.9
25	395	158	180	192	200	200	202	200	2620	13.0
35	395	168	186	199	200	200	205	208	2900	16.9
45	400	171	188	202	200	200	205	212	3180	20.6
55	400	175	192	204	200	200	206	216	3450	24.5

Table 33

These results were plotted on the three dimensional diagrams relating screw speed, back pressure, and output, as given by the results of table 29, see extruder operation E fig. 9 and 10. The plot of the above results was seen to fit exactly on to the plane representing the extruder performance at different back pressures, when no die was used. These results therefore indicate that under these conditions of autogenous extrusion with polypropylene the output from an extruder and die is directly proportional to the screw speed and back pressure irrespective of actual design of the die itself. Because of this conclusion, it is not considered necessary to quote the layflat film die dimensions in detail.

4.3.6. Effect of altering the die temperature

In the preceding section the temperature of the breaker plate assembly and the die were arbitrarily set at 200°C and yet the results fitted perfectly into the three dimensional graph defining

Effect of Die and Die Temperature on Output of Polypropylene.

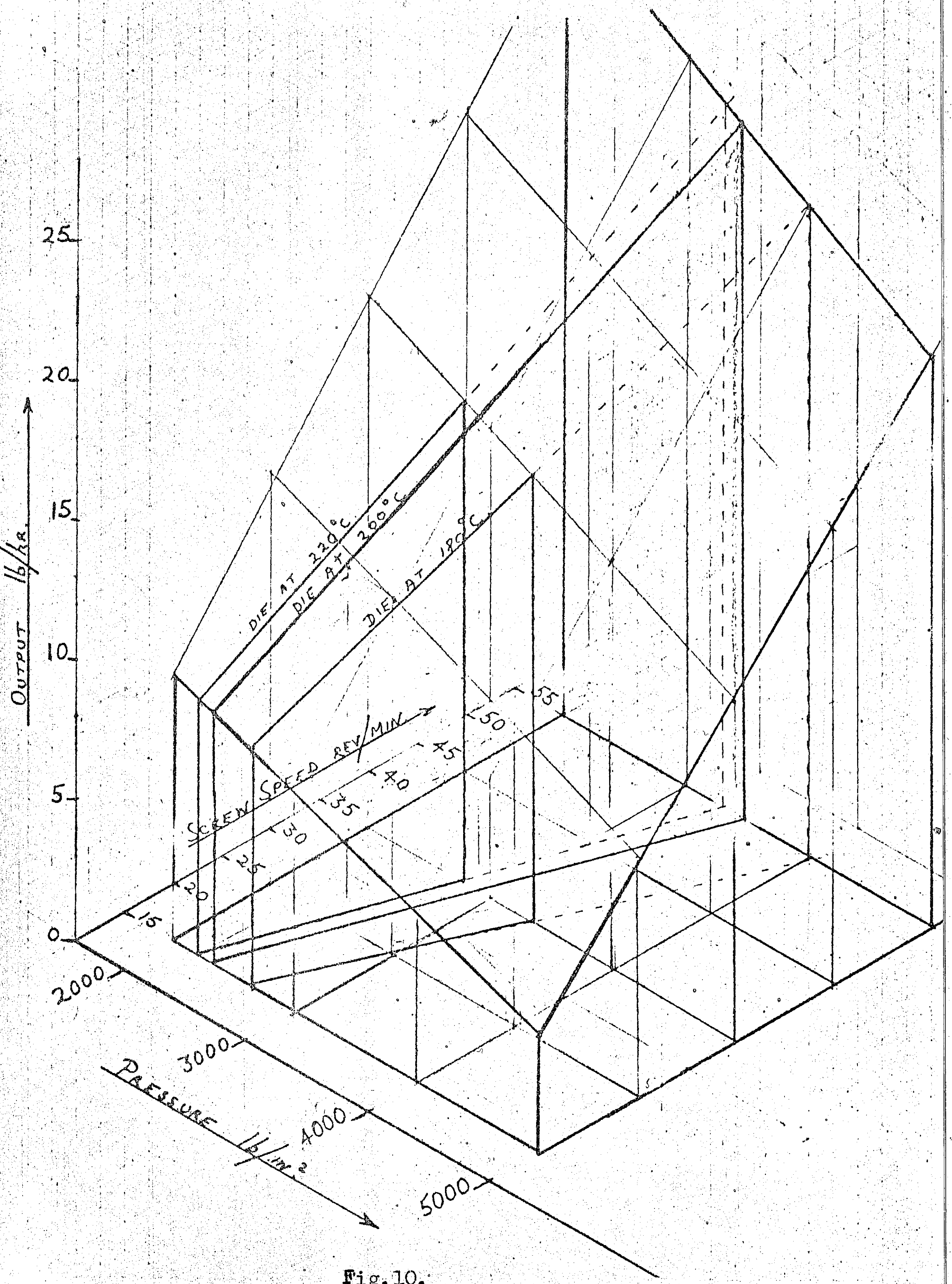


Fig. 10.

the autogenous extrusion of polypropylene at different back pressures without a die. It was therefore decided to investigate the effect of die temperature on output and back pressure at different screw speeds and trials were conducted at high and low die temperatures for extruder "set-up" conditions E (i.e. extruder + breaker plate + screen pack + die).

Results

Screw Speed rev/min	Torque lb ft.	Recorded Temps. $^{\circ}$C							Press lb/in^2	Output lb/hr.
		Barrel			Breaker Plate Ass.	Die		Plastic Temp.		
		1	2	3		1	2			
Lower die temperatures										
15	360	154	181	192	200	180	180	208	2640	8.7
25	380	165	181	194	200	180	180	211	2970	12.6
35	380	170	186	195	200	180	180	213	3340	16.0
Higher die temperatures										
15	380	147	180	185	200	220	220	208	2200	9.1
25	390	162	182	186	200	220	220	211	2500	13.6
35	395	167	183	187	200	220	220	212	2790	17.2

Table 34

These results for different die temperatures were also found to fit on to the plane of the three dimensional diagram relating screw speed, back pressure, and output, as given by the results of table 29. However, the line showing the interrelationship between screw speed, back pressure, and output becomes more oblique as the temperature of the die is lowered (see fig. 10).

The latter effect is observed because the difference in back pressure over a range of screw speeds increases as the die temperature falls, and decreases as the die temperature rises.

There must, therefore, be an empirical equation which relates screw speed, back pressure and the die temperature to the output for an extruder operating under these conditions.

In the above experiments it was noticed that there was a constant pressure drop of approximately 50 lb/in^2 across the breaker plate assembly.

4.4. Conclusions

Most theories are based on assumptions of isothermal or adiabatic extrusion conditions to keep the mathematical derivations within workable limits. In practice neither of these two operating conditions are obtainable and extrusion is normally conducted under autogenous or polytropic conditions.

Using the equipment described in the introduction to this section the polytropic extrusion of polypropylene at high barrel temperatures gave the best agreement between theory and practice. These particular operating conditions are known to resemble the isothermal extrusion case. Even so, the agreement between the theoretical output derived using the uni-dimensional simplified flow equation and the observed output was surprisingly close when one considers the approximate nature of both the practical data and the theoretical equation used.

The experimental observations obtained for the autogenous extrusion of polypropylene were also compared to theories based on rheology or energy balance equations. These particular operating conditions are known to resemble the adiabatic extrusion case. Both isothermal and adiabatic theories showed signs of agreement with experimental observations for the autogenous extrusion case, but they can not be applied quantitatively.

In all cases these theories have helped to explain in a qualitative manner the observations which occur in practice, and their application has led to a much better understanding of the single screw extrusion process.

Empirical equations which quantitatively define the output characteristics for the autogenous extrusion case were successfully derived. It is thought that similar equations could be derived for an extruder operating under polytropic conditions.

Derived theoretical equations based on isothermal or adiabatic extrusion conditions can therefore be a useful guide to the engineer wishing to design an extruder even though they often do not strictly agree with the results obtained in practice. However, once an extruder and screw are in use, it is possible to derive quantitative empirical equations which clearly define the extrusion characteristics of some particular polymers under certain operating conditions.

The experimental observations recorded in this section show that a variation in the barrel zone temperature can have a great effect on the observed output for an extruder operating under constant conditions of screw speed and back pressure. The temperature, particle size and lubrication qualities of the feed stock are also known to affect greatly the output characteristics of an extruder. Future research directed toward a quantitative study of the effect of these variables in an isolated feed section, would do much to fill one of the largest gaps remaining in our knowledge of the extrusion process.

4.5. References

1. Donald H.J., Rieke J.K., and Humes E.S. SPE Journal p.689 July 1960
2. Maddock B.H., Bakelite Co. pamphlet No.98, May 1957

5. THE MIXING ACTION OF SINGLE SCREW EXTRUDERS

5.1. Introduction

This section outlines the theory derived and the important experimental work conducted by various workers to determine the mixing mechanism of constant pitch single screw extruders.

Theories of mixing based on derived mathematical equations do not in themselves lead to a clear understanding of the mixing action undergone by the plastic melt, or the melting action undergone by the solid plastics feed material. However, numerous workers have furthered our understanding of these mixing and melting actions by visual analysis methods and it is this work which is emphasised in this section.

Mixing Torpedoes are also mentioned briefly.

5.2. Theory of Mixing

The theories which derive expressions for the velocity profiles of the down channel flow and the transverse channel flow will not be given in full, only the final conclusion will be stated. A full derivation of these theories is given by various workers. 1.2.3.

The assumptions made are:-

1. The viscosity is constant at all points in the screw channel.
2. The bounding surfaces are flat.
3. The channel is very wide compared with the depth (uni-dimensional flow theory).
4. The barrel surface is assumed to move relative to the stationary screw surface.
5. Leakage flow over the flights is zero.

The velocity profile in the direction of the channel is given by:-

$$V = U \cos \phi \left[H - 3a \ (H - H^2) \right] . \qquad (40)$$

The velocity profile in the transverse plane to the screw channel is given by:-

$$S = U \sin \phi \; (3H^2 - 2H) \quad \ldots\ldots\ldots\ldots\ldots (41)$$

The speed of the fluid at any depth in the channel is given by the vector sum of these two components.

$$m = \sqrt{V^2 + S^2} \quad \ldots\ldots\ldots\ldots\ldots (42)$$

$$m = \sqrt{\left[U \cos \phi \left[H - 3a\,(H - H^2)\right]\right]^2 + \left[U \sin \phi \; (3H^2 - 2H)\right]^2} \quad \ldots\ldots\ldots\ldots\ldots (43)$$

The mathematical nomenclature used above is defined in appendix 3.

The mixing efficiency of a single screw extruder operating without a torpedo is dependent to a large extent on the transverse flow. It is therefore not surprising that the first detailed study of transverse flow was in connection with a paper on the Theory of Mixing in a Single Screw Extruder by Mohr, Saxton & Jepson. (2)

Illustrations taken from a book on extrusion (1) show the theoretical down-channel flow and transverse flow profiles (see fig 11).

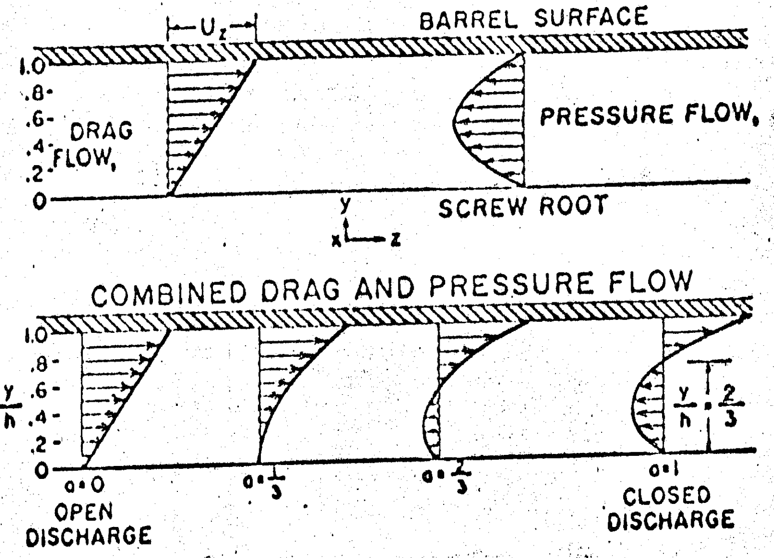
The down-channel flow is a combination of the drag flow and the pressure flow profiles and depends on the screw channel geometry, screw speed, back pressure and plastic melt viscosity. The transverse flow on the other hand does not depend on the back pressure; it is a closed circuit flow.

5.3. Visual Analysis Using Extruders With Transparent Barrels

5.3.1. Experimental determination of down-channel and transverse channel velocity vectors

Eccher & Valentinotti[4] confirmed the theory relating to down-channel and transverse channel flow. They used an extruder with a stationary screw and a transparent rotating barrel to measure the velocity of minute aluminium particles dispersed in polyisobutylene mixed with paraffin oil at 20°C. The extruder was set up in a vertical plane, the fluid was extruded vertically downwards and returned to the feed port via the hollow screw. The barrel was rotated at a velocity

Down Channel Flow Profiles



Transverse Channel Flow Profile

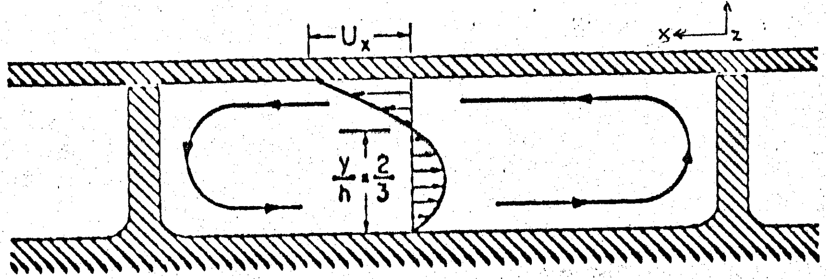


Fig. 11.

of 0.146 rev/min and the viscosity of the fluid used was 1530 poises at 20°C. All optical and mechanical errors were reduced to a minimum by careful design of the apparatus.

The velocity and direction of movement of the aluminium particles was measured at different depths in the screw channel and at different distances in the axial direction from one of the screw flights.

The velocity vector measured in each case was broken down into two components: V in the down-channel direction and S in the transverse channel direction. Profiles of these two velocity components at varying channel depths are shown for three different positions across the channel width in the original paper[4]. These profiles were obtained for three different working conditions of maximum flow, intermediate flow and no output.

The shape of these profiles is similar to those given in fig.11. Three different profiles were obtained for the down-channel flow component at three different output conditions. Only one profile was obtained for the transverse flow component this was valid for all three working conditions. It must, however, be remembered that the shape of all these profiles differs slightly at different distances across the channel width.

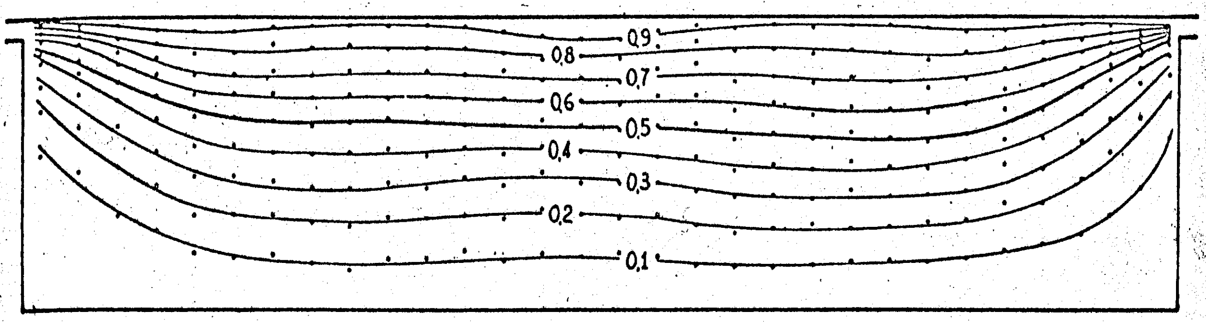
5.3.1.1. Down-channel flow vectors

Eccher and Valentinotti[4] expressed the down-channel components in the form of isovels viewed in the cross sectional plane of the screw channel (i.e. at right angles to the helix angle). These isovels or lines joining points of constant velocity show the fluid velocities for the down-channel flow conditions at various heights in the screw channel. Fig.12. shows the profile of these isovels for the three working conditions of maximum flow, intermediate flow and no output. Each isovel in these figures represents a layer of the fluid moving at a constant velocity relative to the barrel wall $(\frac{m}{U})$

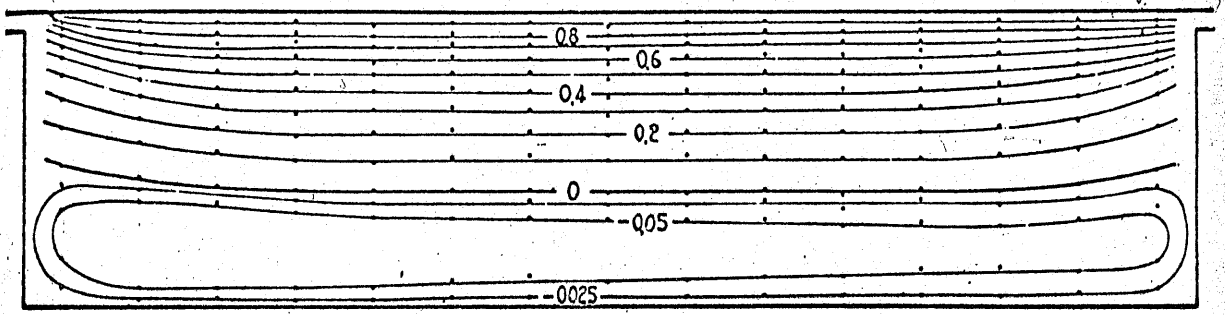
Particles adhering to the barrel surface therefore have a maximum velocity $m/U = 1$, and particles not in motion have a velocity $m/U = 0$.

These isovels which were determined from experimental observations show the effect of the screw flights on the down channel fluid flow vectors. This effect is ignored in the uni-dimensional flow theory because the channel is assumed

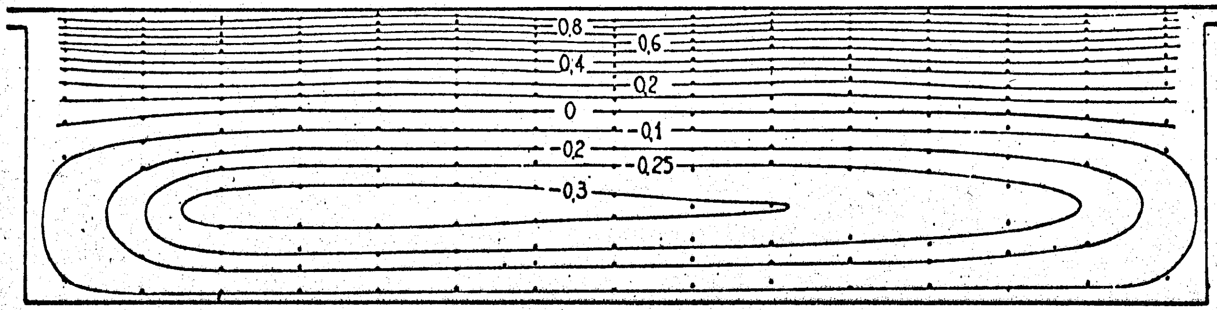
Profile of Isovels Representing Down Channel Flow Velocities at Different Heights in the Screw Channel for Three Working Conditions.



(a) Maximum Flow



(b) Intermediate Flow



(c) No Output

Fig.12.

to be very wide compared with the depth. For the case of maximum flow the isovels have a simple profile, the velocity being zero at the fixed walls and having a maximum positive value at the dragging barrel surface. In the case of intermediate flow the channel section is divided into two zones with positive and negative velocities. In the no-output case the zero line moves up to a position about two thirds the height of the screw channel and the area with negative velocities is about twice the positive one.

In all three cases the effect of the screw flights is less noticeable as the ratio of height to width of the positive zone becomes smaller. Also in all three cases the fluid near the base of the screw flights is almost stagnant and thus the use of generous root fillets is advisable for heat sensitive polymers in order to eliminate this semi stagnant region and reduce polymer degradation.

5.3.1.2. Transverse-channel flow vectors

As already stated the transverse channel flow was similar for all three cases of maximum, intermediate and no-output. The transverse channel flow is, therefore, not influenced by the down-channel flow. The maximum positive value of the transverse component is obtained at the dragging barrel surface. Both the down-channel flow and the transverse flow components are reduced to zero at the screw root.

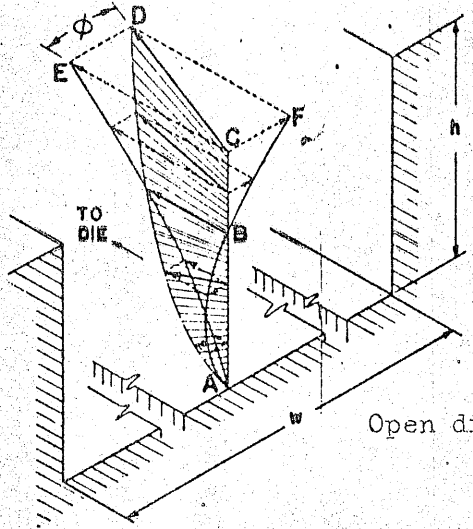
The transverse channel flow is directed downwards when it meets the leading edge of the screw flight and upwards when it meets the trailing edge of the screw flight. The transverse channel flow component is thus a circulation flow.

5.3.1.3. Combination of the down-channel and transverse-channel flow vectors

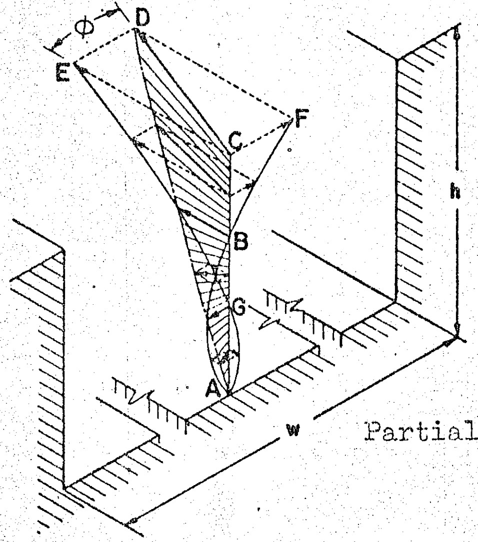
So far, only the components in the two principle directions have been discussed; their resultant is the velocity vector which was initially measured directly.

Three-dimensional sketches are the best way of showing the variations in magnitude and direction of the velocity vector m and its components V and S with channel depth. Fig. 13, which is reproduced from a technical bulletin (5), shows how the transverse and down-channel velocities can be combined to give the resultant velocity of particles at various depths in the centre of the screw channel, for the three working conditions of open discharge, partial restriction to flow and completely closed discharge. Fig. 13 is itself an improved version of a diagram given in the paper by Eccher and

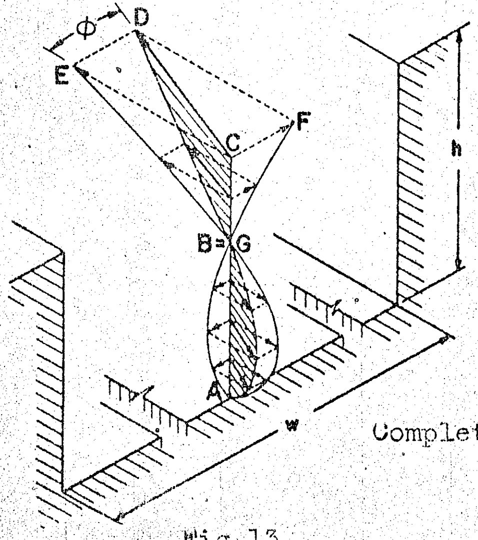
Diagram Showing the Transverse-channel and Down-channel Components and the Resultant Velocity Vectors for Plastics Flow at Three Different Output Conditions.



Open discharge case



Partial restriction to flow case



Completely closed discharge case

Fig. 13.

Valentinotti (4).

The length and direction of the arrows in fig.13 represent the velocity and direction of movement of particles at various depths in the centre of the screw channel. For instance the resultant velocity and direction of movement of a particle at point C is represented by the arrow CD. Smaller arrows CE and CF represent down-channel and transverse channel flow respectively.

Positions B and G represent points of zero velocity for transverse flow and down-channel flow respectively. The transverse flow is seen to be a circulation flow which contributes to the mixing action of the screw and its flow profile is the same for all three output conditions. The transverse flow is shown moving towards the right at the top of the screw channel and towards the left at the bottom of the screw channel. At the flight edges the top layer flows down towards the screw root and the bottom layer flows up towards the barrel surface.

In the completely closed discharge case it is noticeable that the cross over point G for down-channel flow is the same as point B for transverse flow i.e. there is no movement at all of the particles at this point relative to the screw.

For each output condition the resultant velocity vectors form a maximum rotational angle which varies continuously from approx. 90° for maximum flow to 180° for the no-output condition. These conclusions do not hold for the zones close to the walls of the flight, because of the reduction in the transverse component. At these points the angle of rotation is less than 90° for maximum flow and more than 180° for the no-output condition.

5.3.2. Back flow

In the past it has been inferred by many workers, particularly those considering down-channel flow, that under certain conditions of restricted outlet, material can actually flow back up the screw channel. It is true that at some points in the circulatory movement, the velocity can be resolved to have a component backwards in the direction of the helical screw channel. But when account is taken of the helix angle, it is found that no backward movement of the fluid occurs on the direction of the screw axis. Provided there is an output from the extruder, no matter how small, all the fluid is always moving towards the output end of the extruder and the material motion may be described as a spiral within a spiral. This fact has been clearly established by Mohr, Clapp, Squires and Starr (6.7).

These workers too conducted experiments with a stationary screw and a transparent rotating barrel, they considered the influence of transverse flow on the movement of both Newtonian and non-Newtonian fluids in the screw channel.

The Newtonian liquid studied was a corn syrup which had a viscosity of approx. 1500 poises at room temp. For this series of experiments the transparent barrel was rotated at 1 rev/min. The non-Newtonian liquid studied was an 8% solution of hydroxyethyl cellulose in water and the barrel was rotated at 2 - 3 rev/min for this series of experiments. The slope of the viscosity shear rate curve for the hydroxyethyl cellulose solution was approx. the same as that for a polyethylene melt. The back pressure was altered by changing the size of the restriction in the outlet pipe. A small diameter hollow probe was fitted perpendicular to the screw root and at the centre of the screw channel. A number of short smaller tubes branched off at right angles to the probe, the highest of which was 1/32 in. from the barrel surface. The probe and the screw were hollow and through them an intense black fluid could be injected into the moving fluid in a number of streams at various heights in the screw channel.

A circumferential line was drawn on the rotating barrel just to the upstream side of the probe. The black fluid injected through the probe formed into a fan shape pattern and the angular spread of this pattern increased as the ratio of pressure flow to drag flow was increased. Photographs of these patterns presented in technical papers (6.7) demonstrate conclusively that no fluid in the screw channel ever moves axially towards the rear of the extruder (i.e. no injected black fluid ever crosses the circumferential line of the barrel). The flow distribution of both the Newtonian and non-Newtonian fluid were found to be qualitatively the same. The angle of motion of the fluid was always in the range 0 - 180 degrees to the plane perpendicular to the screw axis. For the condition of zero output the only components of fluid velocity are in a plane parallel to the circumferential line of the barrel (angle of motion 0 or 180 degrees) and the axial velocity component is zero at all depths in the chnnel.

These findings verify the work of Eccher and Valentinotti because fig. 13 shows us that the maximum angle of rotation of the resultant velocity vectors is never greater than 180 degrees to the plane DC i.e. particles always move either on the plane DC or in a direction to one side (the outlet side) of the plane DC. Therefore, particles never move towards the rear of the extruder because the plane given by DC is parallel to any plane given by a circumferential line drawn on the extruder barrel. The only exception to this may occur near the screw flights but in practice this is not significant.

The cross channel flow has an important component in the direction of the axis of the screw and S may be regarded as negative when V is positive and vice versa. During the time the material at the bottom of the screw channel appears to be going rearwards in the channel, it is also moving towards the leading edge of the screw flight and provided there is any output at all the net motion is towards the output end of the extruder.

The author has also witnessed a film presented by E.I. du Pont de Nemours and Co. Inc. which used this latter transparent barrelled extruder to demonstrate the above facts.

These workers also obtained quantitative data on velocity profiles by removing the probe and measuring, by the use of two cameras, the movement of white resin beads of diameter 25 - 30 mil. They measured the particle speed and the angle of particle travel over a range of channel depths and back pressure conditions for Newtonian and non-Newtonian fluids. It was noticed that the flow patterns produced with the non-Newtonian fluid did not have as great an angular spread as those produced with the Newtonian fluid at comparable values of back pressure. The final conclusions reached were that neither type of fluid flows towards the rear of the extruder, and both fluids give a similar picture of flow distribution in the screw channel.

This latter conclusion adds support to the use of the simplified extrusion flow theories for calculations involving a non-Newtonian polymer melt.

5.4. Visual Analysis Using Coloured Extrudates

Maddock (8) and Street (9) have both studied the mixing process in single screw extruders. Both authors experimented on the extrusion of pigmented polymers. When the extruder had been stabilised at a desired set of operating conditions, it was stopped and the barrel was cooled.

After cooling, the screw was removed from the barrel and various coloured photographs were taken of the full screw and the unwrapped material helix. Photographs were also taken of sliced cross channel sections of the plastic in the melt and fluxing zones.

Maddock conducted mixing experiments on coloured polyethylene and plasticised PVC and Street extruded natural and coloured polystyrene and rigid PVC compounds. The conclusions reached by both these authors are given below.

The mixing efficiency of a single screw extruder is increased by:-

1. decreasing the channel depth
2. increasing the back pressure
3. cooling the screw

The mixing efficiency is decreased by the addition of surface lubricants to the cold granules because they delay fluxing by reducing adhesion to the barrel and screw surfaces. An excessive quantity of lubricant can completely disrupt the normal fluxing and circulating flow behaviour, to the extent that essentially no mixing takes place.

It is noticeable that for a constant screw speed, the mixing efficiency is usually only increased at the expense of loss of output. Theories of melt flow were found to hold approximately true in practice. Street (9) found that the spiral of material which was moving most rapidly towards the die was centred on an area approximately two thirds of the way up from the root of the screw channel and two thirds of the channel width in front of the leading flight.

5.4.1. Melting action of granules

Maddock (8) and Street (9) both thoroughly investigated the forward movement and melting action undergone by the solid plastics granules as they travel from the feed section to the metering section of a constant pitch screw.

The basic mechanism may be defined as follows, the plastic particles are conveyed rapidly along the screw until they become subjected to a shearing action between the root of the screw and a thin skin of molten material along the barrel. This thin skin of molten material is formed by the hot surface of the barrel. The leading edge of the screw flight scrapes up some of this molten material, and a reservoir of molten plastic, which moves along a helical circulating path is built up at the back of the screw channel. In front of this reservoir there is an area of fluxed material and an area of cold granules. These granules move at a faster rate than the melt, and their average temperature gradually rises as they move along the screw as a result of work heat generated and conducted heat which surrounds them on all sides.

Further melting of the cold and fluxed material is caused by the hot surface of the barrel and also the heat given off by the molten reservoir into the adjacent fluxed material. Thus as the free flowing granules and the fluxed material flow past the plastic melt, they are gradually absorbed into it, and the molten reservoir in front of the leading edge of the screw flight (i.e. at the rear of the screw channel) grows increasingly larger.

Under ideal conditions of screw design and applied heat the molten material gradually fills the entire channel as it approaches the output end of the extruder. Both Maddock (8) and Street (9) showed that the path of the polymer melt is a helix within a helix. The photographs taken by these workers clearly show the rotation of the plastic melt in the plane transverse to the channel.

When a back pressure is applied, the material is deflected from its normal course perpendicular to the helix angle, to a more acute angle. Consequently the material travels along a longer helical path, the rate of output drops. Raising the back pressure causes earlier fluxing and gives better mixing, even when the screw speed has been raised to give an output rate similar to the initial output.

5.5. Mixing Torpedoes

A torpedo is a specially designed restriction attached to the front of the screw, and its effect can be threefold.

1. Results in better mixing of the polymer melt
2. Raises the plastic temperature by mechanical working
3. Reduces the surging action of a plastic melt which is sometimes obtained from the output end of the extruder

Torpedoes of widely varying lengths and designs are used in practice. They may be cylindrical or tapered in shape and possess surfaces which are smooth or channelled, the design depends on the melt characteristics and the final use of the plastic being extruded. The range of designs a torpedo may take is almost limitless and for this reason only two of the most popular designs will be discussed in any detail.

5.5.1. The smear head torpedo

This is usually cylindrical in shape with a perfectly smooth surface. The clearance between the torpedo surface and the bore of the barrel is usually less than the depth of the preceding flight. The plastic material which passes over this torpedo is thus subjected to a strong shearing action which raises the plastic temperature, gives improved mixing and lessens the pulsating or surging tendencies of the plastic melt along the screw.

The disadvantage sometimes experienced with a smear head torpedo is that it may result in excessive back pressures and low output rates.

The smear head torpedo has often been used for PVC provided the clearance is not too small, because the shearing and mixing action is not too severe. Thus the plastics temperature is not raised sufficiently to cause degradation of the compound.

5.5.2. The Dow mixing torpedo

As the name suggests, this torpedo was developed by the Dow Chemical Co. It's design consists of helically fluted zones interspaced by smooth relieved radial grooves (see fig. 14(X)).

In this case the molten thermoplastic is carried forward by the torpedo and does not rely entirely on the pressure generated by the screw. The thermoplastic is rolled and forwarded in the helical flutes by the rotation of the torpedo. All the material is thus brought into contact with the cylinder wall. Frictional heat is generated within the plastics by this rolling action and also by the intense shearing action taking place between the flute lands and the cylinder bore. A large land clearance of approx. 0.015 in. is advocated by the initial designers of this torpedo, The Dow mixing torpedo normally has three or four radial grooves along its length, which are really expansion chambers which allow intermixing of the plastic melt. These radial grooves are also thought to minimise any plastic memory effect caused by the torpedo flutes.

The use of this torpedo gives a more uniform temperature, and hence viscosity , throughout the plastic melt. It also helps to overcome surging problems and possesses good mixing properties at high output rates. The output rate is to some extent governed by the helix angle and cross sectional shape of the spiral flutes. The Dow mixing torpedo was primarily developed for the extrusion of ethyl cellulose and polystyrene, it has also been used successfully for the extrusion of cellulose acetate. However, it should not be used for the extrusion of PVC compounds because the temperature developed often causes the degradation of this polymer.

5.6 References

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6. EXPERIMENTAL WORK ON THE MIXING ACTION OF SINGLE SCREW EXTRUDERS

6.1. Introduction

Facilities were made available by the BX Plastics compounding department to investigate the effect of torpedo design on the output and extrudate quality characteristics of a $3\frac{3}{4}$ inch single screw production extruder. Little quantitative data was available on the output and mixing efficiency of the established screw design when fitted with various mixing torpedoes. The present work was limited to the extrusion of PS and TPS materials because nearly all the moulding compound produced by this department is based on these two polymers. Thus, the problem of material cost usually associated with experimental work on large extruders was overcome because good quality extrudate was despatched to customers and poor quality extrudate was re-used in other production orders.

This study of mixing torpedoes was conducted in three stages.

6.2. Investigation of the interdependence of four major extrusion variables, (output, back pressure, screw speed and plastic melt temperature) when mixing torpedoes are used.

6.3. The effect torpedo design has on output, plastic melt temperature, and total power input.

6.4. Visual analysis of the quality of mixing obtained with various torpedo designs.

All important findings of practical value to the compounding department are given.

6.1.1. Equipment used

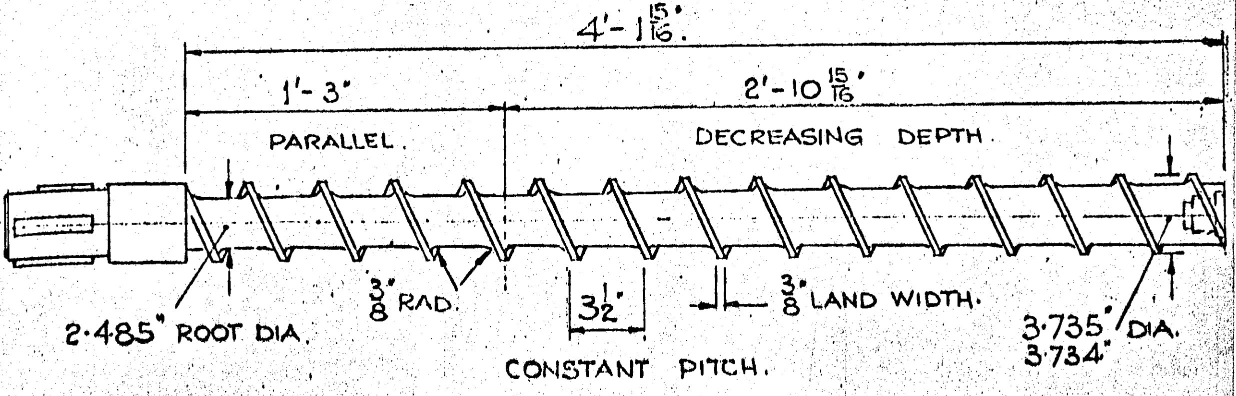
(a) The extruder screw and torpedo designs used are given in Fig.14.

(b) The die and breaker plate design used are given in Fig. 15.

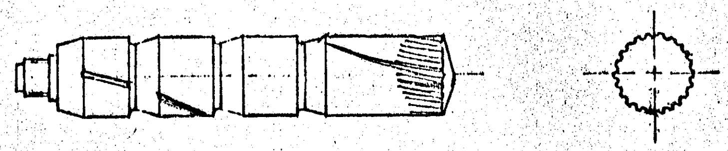
(c) A diagram of the extruder 'set-up' is given in Fig. 16.

Throughout these experiments the back pressure was altered by varying the screen pack, because it was not intended that machine operators would use valves under production conditions. The observations were recorded under the following conditions of back pressure.

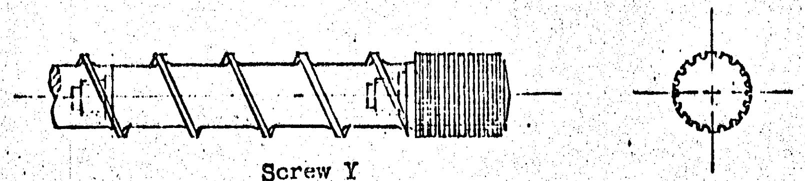
Screw and Torpedo Designs Used With the 3¼ inch Production Extruder



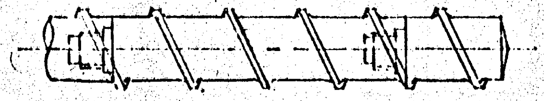
BASIC SCREW DESIGN.



Screw X
DOW TORPEDO



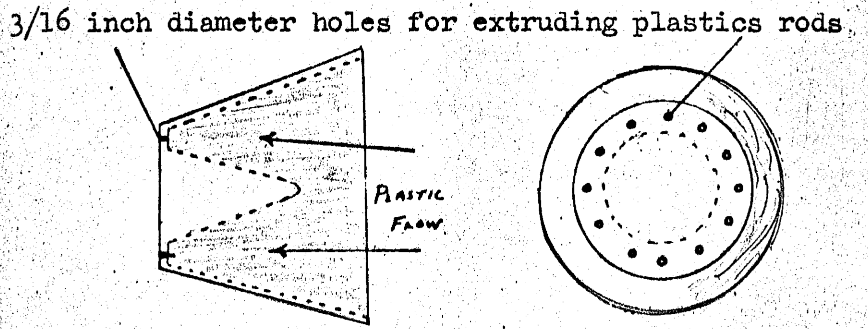
Screw Y
PINEAPPLE TORPEDO AND EXTENDED METERING SECTION



Screw Z
NO TORPEDO (EXTENDED METERING SECTION).

Fig. 14.

Die Design



Breaker Plate Design

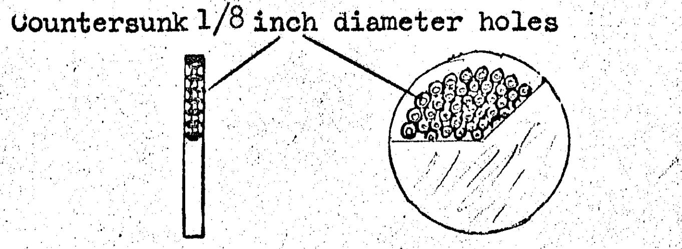


Fig.15.

Diagram of $3\frac{3}{4}$ inch Production Extruder

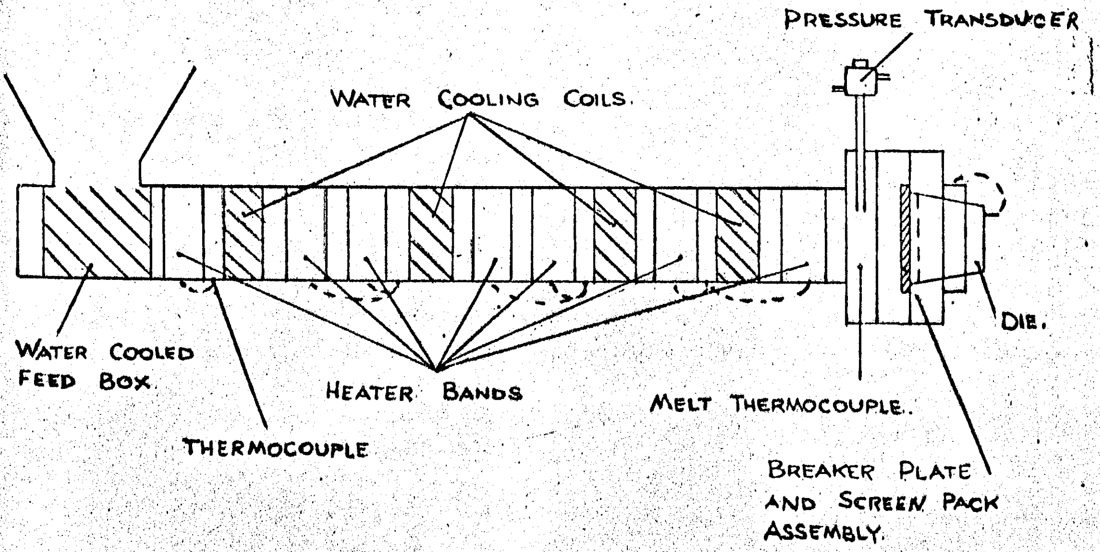


Fig.16.

Breaker Plate Assembly Type	OPERATING CONDITIONS
A	Open discharge condition (zero back pressure)
B	Breaker plate plus die but no screen pack
C (60 mesh)	As B plus 1 x 60 mesh screen pack
D (120 mesh)	As B plus 1 x 60 and 1 x 120 mesh screen pack
E (200 mesh)	As B plus 1 x 60, 1 x 120 and 1 x 200 mesh screen pack

Table 35

The screens used were of the following dimensions.

TYPE (mesh)	DIAMETER OF WIRE (inches)	SIZE OF OPENING (inches)
60	0.0070	0.0097
120	0.0036	0.0052
200	0.0021	0.0029

Table 36

The temperature of the four barrel and the die heat zones were controlled, through thermocouples sunk in the barrel, by five West temperature controllers.

A specially constructed Extrusion Performance Meter, designed by D.C. Nicholas and D.R. Reid of BX Plastics Ltd., was used for recording:

1. The temperature of another set of thermocouples which were situated directly below those already mentioned, and which penetrated down to the outer surface of the barrel liner.

2. The pressure on the screw side of the breaker plate as measured by a Dynisco pressure transducer (Model PT58).

3. The plastic melt temperature in a position opposite the pressure transducer.

4. The screw speed (a tachometer was fitted to the screw driving shaft).

5. The power input to the screw drive motor.
6. The integrated power input to the barrel heaters.

Initially a 30 H.P. motor capable of giving a screw speed range of 18-60 rev/min was used, but later this was replaced with a 36.7 H.P. motor capable of giving 25-94 rev/min.

In all cases take-off equipment consisted of a water bath, pull down rolls, capstans (to aid air cooling and drying) and a dicing machine.

6.2. Mixing Torpedoes And The Interdependence Of Extrusion Variables

The interdependence of four major extrusion variables (output, back pressure, screw speed and plastic melt temperature) when using three screws of different design are studied in this section.

The first series of experiments was conducted with the pineapple torpedo screw (see Fig. 14, screw Y).

The material extruded was TPS (grade Bextrene 235).

The barrel and die temperatures were preset at the following values:

BARREL TEMPERATURES °C ZONES				DIE TEMPERATURE °C
1	2	3	4	
180	190	200	200	180

Table 37

VISCOSITY DATA FOR TPS

(Grade Bextrene 235)

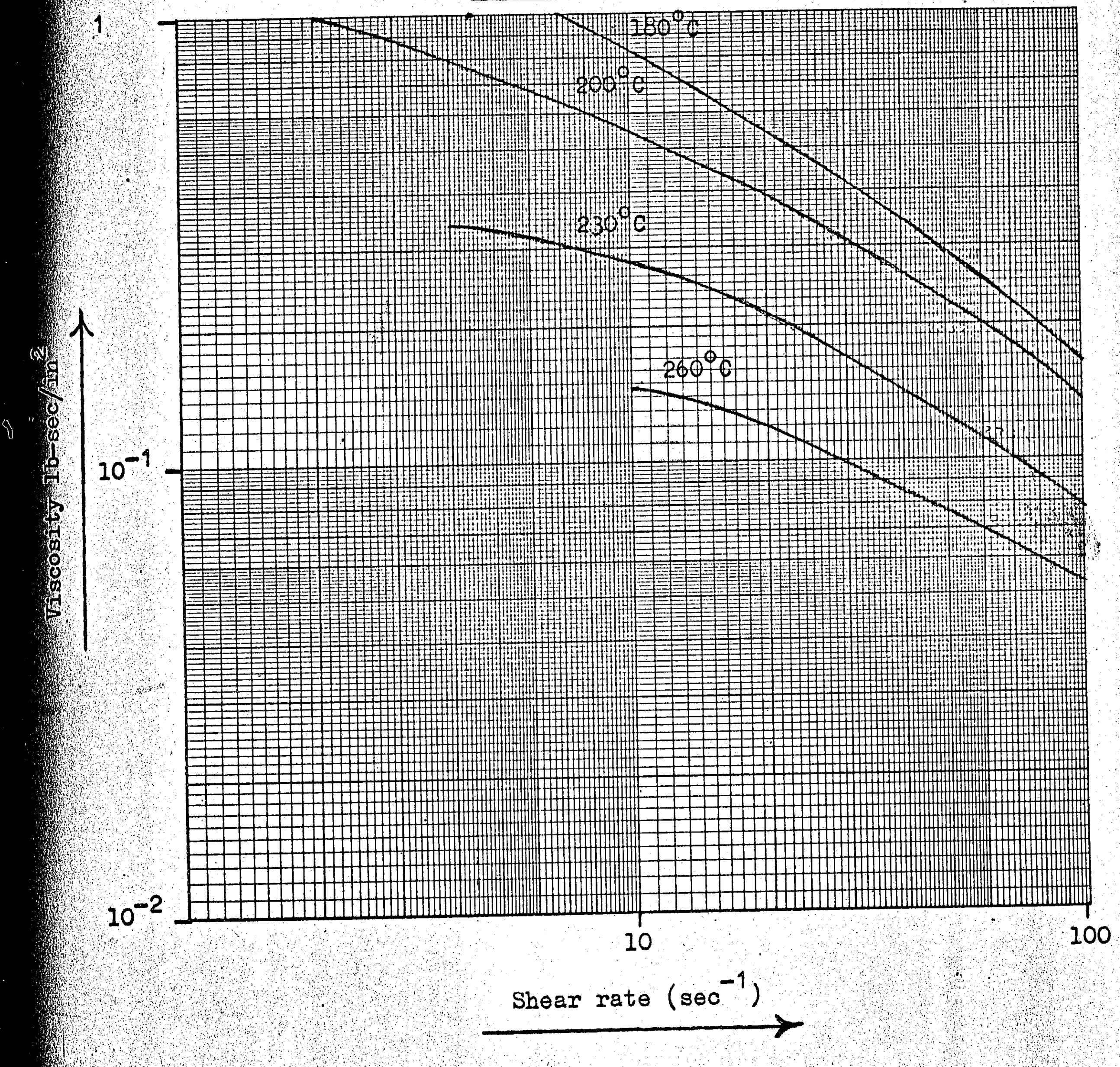


Fig. 17

6.2.1.

Results for the pineapple torpedo screw
(screw Y)

Breaker Plate ass.	Screw Speed rev/min.	Recorded temperatures ^{o}C						Back Press lb/in^2	Shear rate sec^{-1}	Viscosity $lb/in^2/sec$	Observed output lb/hr	Calculated output (equ.10) lb/hr
		Barrel				Die temp.	Plastic temp.					
		1	2	3	4							
A	20	-	-	-	-	-	-	-	10.6	-	190	244
B		157	186	182	173	175	179	1,125		0.80	178	228
C		155	182	179	173	172	181	1,175		0.80	178	227
D		156	183	181	179	172	184	2,000		0.75	168	213
E		164	187	194	192	178	195	4,600		0.59	125	157
A	30	-	-	-	-	-	-	-	16.0	-	240	366
B		170	190	190	182	174	186	1,125		0.57	230	344
C		167	187	187	180	172	188	1,200		0.55	225	342
D		164	185	186	182	175	191	1,950		0.55	215	326
E		177	188	201	200	175	202	4,550		0.42	170	248
A	40	-	-	-	-	-	-	-	21.3	-	285	489
B		180	192	195	186	176	192	1,075		0.45	268	462
C		179	190	195	188	172	193	1,200		0.44	265	458
D		176	188	193	188	172	196	1,900		0.40	255	436
E		187	190	207	204	177	208	4,500		0.32	210	331
A	50	-	-	-	-	-	-	-	25.4	-	325	611
B		191	193	204	194	174	198	975		0.32	305	577
C		187	190	202	195	175	199	1,125		0.32	298	574
D		185	188	201	195	175	201	1,875		0.31	293	544
E		192	194	211	205	177	212	4,450		0.26	250	421
A	60	-	-	-	-	-	-	-	31.9	-	355	733
B		199	199	210	198	174	203	925		0.27	335	695
C		196	195	210	200	176	204	1,060		0.26	330	687
D		194	195	209	200	173	206	1,850		0.26	323	653
E		200	202	218	207	180	215	4,400		0.22	278	509

Table 38

The calculated output was determined by using the Unidimensional Flow Equation (10). The viscosity data used in the pressure flow and leak flow terms was based on rheometry work conducted by D.R. Reid of the Physics Department, BX Plastics Ltd. The viscosity, shear rate and plastic temperature relationship for Bextrene 235 is given in Fig. 17. The length of the screw full of plastic melt (L) used in equation (10) was assumed to be 36 inches (i.e. length of metering section plus torpedo).

NB. These experiments were repeated in full with the Dow torpedo screw (see Fig. 14 screw X) and with the extended metering section screw (see Fig. 14, screw Z). The results of the latter two series of experiments are not given because in all cases similar relationships to those discovered for the Pineapple torpedo were found to exist.

6.2.2. Discussion of results

These results show remarkably good agreement considering the approximate nature of equation (10) and the viscosity data, and also the assumption that L is constant at 36 inch. The calculated output was expected to be greater than the observed output, because the resistance to plastic flow caused by the torpedo was not accounted for in the former case.

It is noticeable that the ratio of the observed output to the calculated output using equation (10) forms an almost identical pattern to the results obtained for polypropylene (compare table 29 and table 38). The reason for this is not clear, although it could be partly due to the fact that L was assumed constant in both cases.

Comparing the recorded barrel temperatures with the preset barrel temperatures indicates that autogeneous conditions are approached as the screw speed and/or back pressure are increased.

6.2.2.1. The relationship between output, back pressure and screw speed

A linear relationship exists between output and back pressure at any given screw speed, as shown in Fig. 18. A three dimensional plot of the above three variables is given in Fig. 19. This plot shows that an almost planar relationship exists for this TPS material similar to that obtained for polypropylene (compare Fig. 19 with Fig. 9). As autogeneous extrusion conditions are approached at higher screw speeds and back pressures, there is a tendency towards a less planar relationship. However, an emperical equation similar to one derived for polypropylene could be applied to this case.

Unlike the polypropylene case, an increase in screw speed usually caused a decrease in the recorded back pressure for any given breaker plate assembly, see Fig. 20. This effect was observed

Fig. 18.

Graph Showing Relationship Between Output, Back Pressure and Screw Speed for Screw Y (Pineapple Torpedo) Using TPS (Grade Bextrene 235).

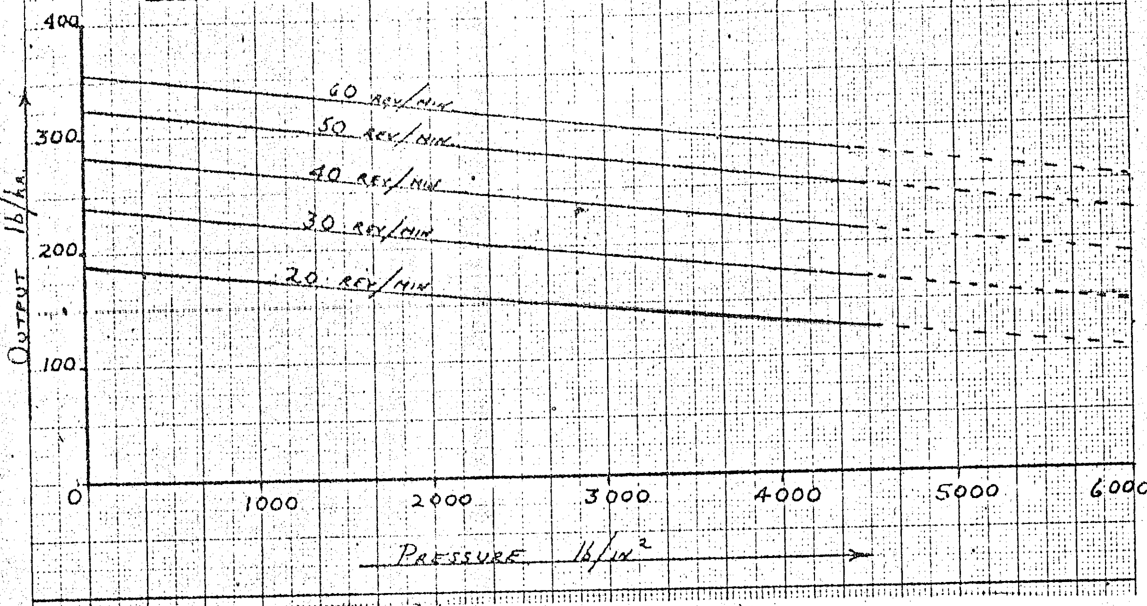


Fig. 19. Three Dimensional Plot of Output, Back Pressure and Screw Speed for Screw Y (Pineapple Torpedo) Using TPS (Grade Bextrene 235).

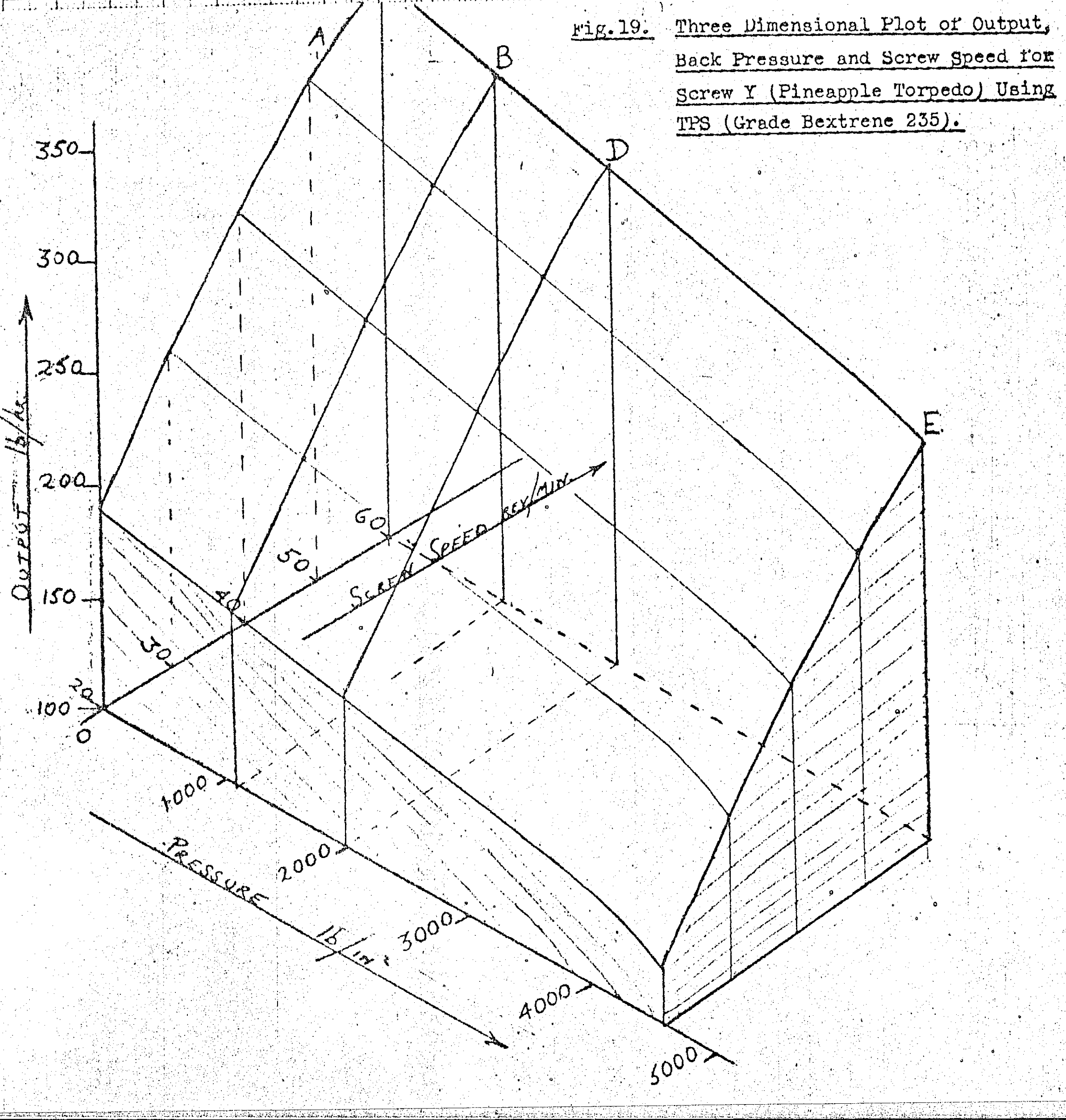


Fig. 20.

Graph Showing Drop in Back Pressure as the Speed of Screw Y (Pineapple Torpedo) is Increased Using TPS (Grade Bextrene 235).

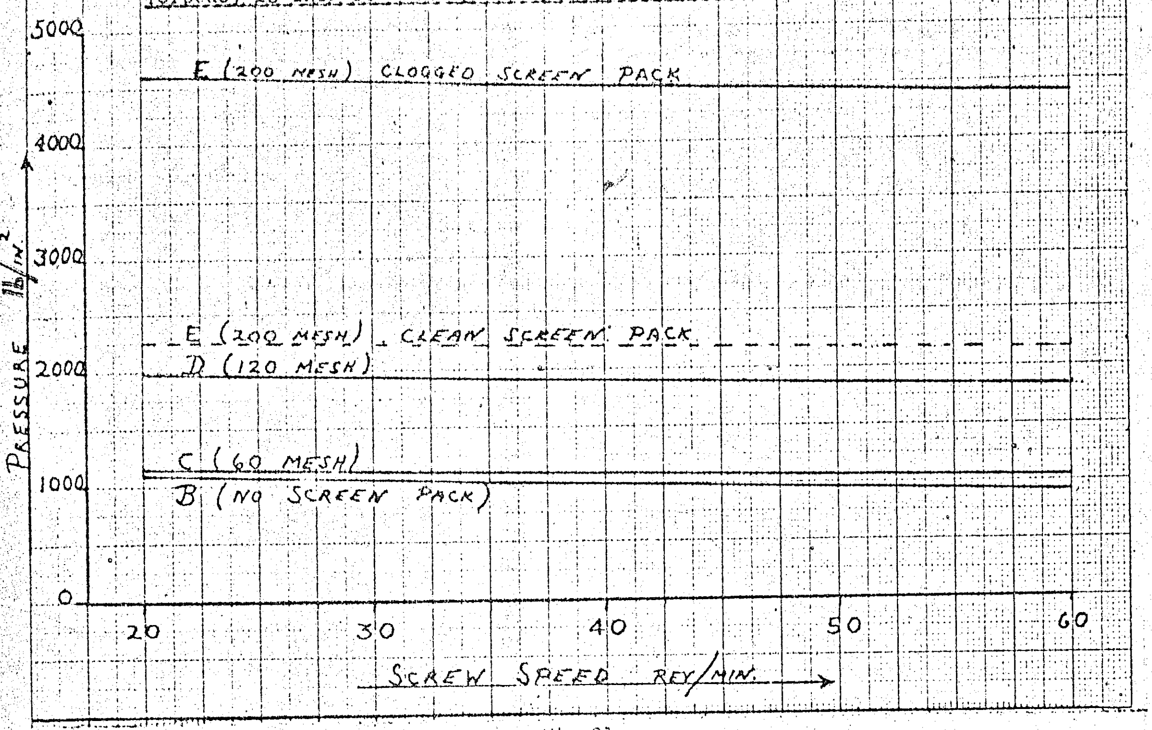
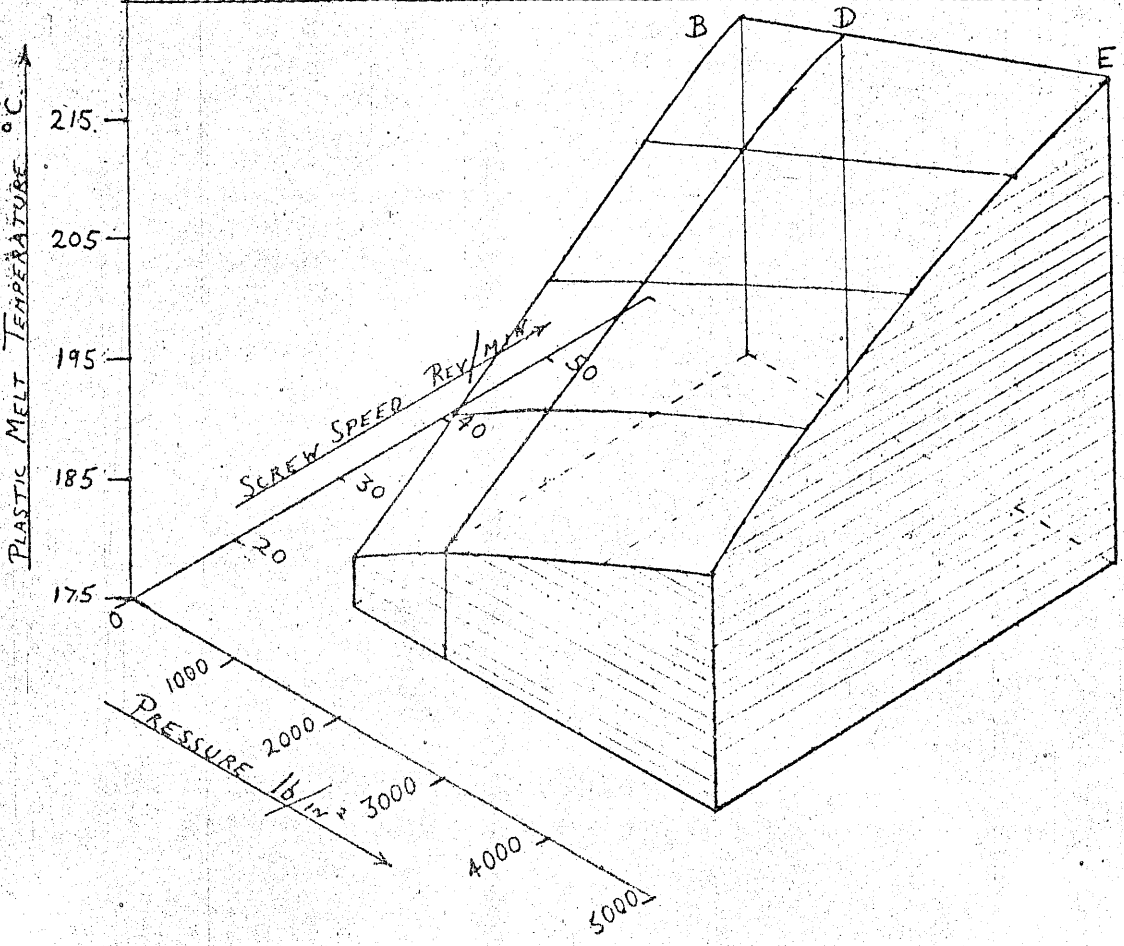


Fig. 21.

Three Dimensional Plot of Screw Speed, Back Pressure and Plastic Melt Temperature for Screw Y (Pineapple Torpedo) Using TPS (Grade Bextrene 235).



even when the screw was not fitted with a torpedo and since the output rose with screw speed, it must be related to an increase in the plastic melt temperature.

NB. Screws X and Z (Dow torpedo and no torpedo respectively) gave similar almost planar relationships between output, back pressure and screw speed.

6.2.2.2. The relationship between screw speed, back pressure and plastic melt temperature

A three dimensional plot of the above three variables is given in fig. 21. This three dimensional plot is based on the observations recorded for breaker plate assemblies type B, D (120 mesh) and E (200 mesh) which are listed in table 38. As expected the plastic melt temperature rose as the back pressure and/or screw speed were increased.

The most significant feature of this three dimensional plot is that the shape is almost a mirror image of fig. 19, the latter shows the observed output in relation to these same back pressure and screw speed combinations. In making a direct comparison between fig. 19 and fig. 21 it must be remembered that the screw speed and back pressure axis scales are slightly different.

NB. Screws X and Z (Dow torpedo and no torpedo respectively) gave similar almost planar relationships between screw speed, back pressure and plastic melt temperature.

In conclusion therefore it may be stated that screws with or without mixing torpedoes give a direct inter-relationship between screw speed, back pressure, plastic melt temperature and output. No doubt an emperical equation could be derived which would connect these four extrusion variables.

Unfortunately the instrument for measuring the integrated heat and power input to the extruder was faulty during these series of experiments and therefore no attempt could be made to correlate these quantities with the above four variables.

6.3. The Effect Of Torpedo Design On Output, Plastic Melt Temperature, And Total Power Input

To simplify the presentation of all the following results, only the observations recorded using a breaker plate assembly type D (120 mesh) will be given. However, the results obtained at other conditions of back pressure will be referred to.

6.3.1. The effect of torpedo design on output

Experiments were conducted with all three screws, over a screw speed range of 30-90 rev/min. Observations of surging phenomena and output were recorded for both natural PS (grade Bextrene 7000) and natural TPS (grade Bextrene 235) polymers, at the open discharge condition and also in conjunction with breaker plate assemblies type C (60 mesh), D (120 mesh) and E (200 mesh).

6.3.1.1. Results

Screw speed rev/min	Observed Output lb/hr.					
	Screw X		Screw Y		Screw Z	
	PS	TPS	PS	TPS	PS	TPS
30	160	150	205	215	225	240
40	190	175	240	255	265	275
50	220	200	280	290	300	310
60	255	230	315	325	330	340
70	290	255	345	355	360	370
80	315	280	370	380	385	395
90	335	305	400	405	410	420

Table 39

6.3.1.2. Discussion of results

In most cases, the output to screw speed relationship was almost linear, for any given breaker plate assembly. The one major exception to this occurred when TPS was extruded at the open discharge condition using screw Z where a plug flow effect of granules was observed.

(a) Surging

A qualitative comparison was made between the surging phenomena experienced with all three screws, a summary of the results is given below in table 40.

Screw design	Material extruded	Surging observed with breaker plate ass. Type D (120 mesh)	Type E (200 mesh)
X	PS	None	None
Y	PS	Bad	Bad
Z	PS	Bad	Bad
X	TPS	None	None
Y	TPS	Slight	None
Z	TPS	Bad	Bad

Table 40

These results suggest that a well designed torpedo and/or an increase in back pressure both help to overcome surging of the plastic melt in the extruder screw.

(b) Bulk density

The bulk densities of the two feed materials used are given in table 41.

Method of determination - German Specification DIN 53468

Feed material (rough ground)	Average weight of 250 ccs (gm)	Range of 10 weighings (gm)	Bulk density (lb/cu.ft.)
PS	129.8	± 0.2	32.4
TPS	117.6	± 0.3	29.4

Table 41

The bulk density of the PS feed material was found to be greater than that of the TPS but only in the case of the Dow torpedo (screw X) was the observed output of PS greater than that of TPS, for similar conditions of screw speed and breaker plate assembly. The different flow properties of these two materials at different plastic melt temperatures probably account for this phenomenon.

The bulk density of $\frac{1}{8}$ inch diameter PS extruded pellets at room temperature was found to be 36.2 lb/cu.ft., this is approximately 10% higher than the initial rough ground PS feedstock. Extrusion of these pellets, using screw X (Dow torpedo), gave a corresponding 10% increase in output for similar operating conditions of screw speed etc. A similar effect was observed when TPS pellets were re-extruded using screw Y (Pineapple torpedo).

(c) Restriction to flow

It was observed in all cases that the output of screw X (Dow torpedo) was less than screw Y (Pineapple torpedo) which in turn was less than screw Z (no torpedo).

The Dow torpedo used caused a considerable restriction to melt flow compared to that observed for the screw without a torpedo, whereas the Pineapple torpedo used only caused a slight restriction. Please note this may not be the case if both torpedoes were the same length. (The Pineapple torpedo used was considerably shorter than the Dow torpedo). The results of the next section on quality of mixing, indicate that with the extrusion of PS and TPS materials, up to 30% of the possible output has to be sacrificed in order to get effective mixing.

The results of this section indicate two other important findings of practical value.

(i) An approximate 30% increase in output can be obtained with a Dow torpedo screw if drive motors capable of giving a screw speed up to 90 rev/min replaced the existing motors which are only capable of giving screw speeds up to 60 rev/min. The thrust bearings and gear ~~bex~~ of the extruder used are capable of withstanding these heavier loadings provided they are not misused.

(ii) It is possible to extrude good quality natural TPS at outputs of 405 lb/hr using screw Y (Pineapple torpedo) at 90 rev/min if no dispersion of different grades or colour is involved). The maximum output rate prior to this work was 230 lb/hr using screw X (Dow torpedo) at 60 rev/min.

6.3.2. The effect of torpedo design on the plastic melt temperature

Experiments were conducted with all three screws over a screw speed range of 30-90 rev/min. Observations of the plastic melt temperature were recorded for both PS (grade Bextrene 7000) and TPS (grade Bextrene 235) materials for breaker plate assembly conditions C (60 mesh), D(120 mesh) and E (200 mesh).

6.3.2.1. Results for the breaker plate assembly type D (120 mesh) case

Screw speed rev/min	Recorded Plastic temperature °C					
	Screw X		Screw Y		Screw Z	
	PS	TPS	PS	TPS	PS	TPS
30	201	192	193	191	180	182
40	210	197	197	196	184	187
50	216	203	200	201	188	191
60	221	208	204	206	192	195
70	225	213	208	210	197	199
80	229	218	211	213	201	204
90	232	223	215	216	205	207

Table 42

6.3.2.2. Discussion of results

In all cases, including other breaker plate assemblies, the plastic temperature was observed to rise when the screw speed and/or the restriction caused by the breaker plate assembly were increased. It was also noticeable that the plastic temperature for both PS and TPS materials using screw X was greater than screw Y which in turn was greater than screw Z, for any given screw speed and breaker plate assembly.

The recorded plastic temperatures of both PS and TPS materials were within ± 3°C of each other for screws Y and Z operating under similar conditions, but not in the case of screw X. In the latter case the plastic temperature of PS materials was always approximately 10°C higher than that of TPS materials, at similar conditions of screw speed etc., the reason for this is not fully understood. Unfortunately the back pressure recording equipment was out of order when the experiments using PS feedstock were conducted. It is thought that such pressure observations would help to explain the higher outputs and higher plastic temperatures obtained with PS materials compared with TPS, in the case of screw X.

6.3.3. <u>The effect of torpedo design on the drive and integrated heat power inputs</u>

Experiments were conducted with all three screws over a screw speed range of 30-90 rev/min. Unfortunately the drive power and integrated heat input measuring equipment was out of order when the TPS experiments were conducted. Data of the drive and heat input is therefore limited to observations recorded when PS (grade Bextrene 7000) was extruded at the open discharge condition, and also in conjunction with breaker plate assemblies type C (60 mesh), D (120 mesh) and E (200 mesh). The temperature controllers were of course set at constant values throughout as recorded in Table 37.

6.3.3.1. Results for the breaker plate assembly type D (120 mesh) case

Screw speed rev/min	Recorded power K.W.					
	Drive input			Integrated heat input		
	Screw X	Screw Y	Screw Z	Screw X	Screw Y	Screw Z
30	24	27	27	5	7	6
40	31	37	35	5	6	5
50	38	46	44	4	6	5
60	45	55	52	3	5	4
70	52	63	59	2	4	4
80	60	69	66	1	3	3
90	67	75	72	0	1	2

Table 43

6.3.3.2. Discussion of results

a) Integrated heat input

In all cases, including other breaker plate assemblies, the integrated heat input decreased with an increase in screw speed. The work heat developed by the plastic material increases as the screw speed increases because of the higher shear rates, and this extra work heat more than compensates for the heat absorbed by the extra output of hot material. Consequently less heat is required from the barrel heaters as the screw speed increases because of the extra work heat developed within the plastic material.

Screw X (Dow torpedo) is capable of developing a larger quantity of work heat within the plastic melt than either screws Y (Pineapple torpedo) or Z (no torpedo). This, together with the fact that less output was obtained from screw X, accounts for the lower integrated heat input figures and the higher plastic melt temperatures, at any given screw speed. A similar comparison between screws Y and Z shows that the integrated heat input results are similar, but the plastic melt temperature for any given screw speed is higher in the case of screw Y.

Integrated heat input measurements were also obtained for the extrusion of TPS (grade Bextrene 235), using screw X and a breaker plate assembly type D (120 mesh). The results showed that the power input through the barrel heaters dropped from 6.5 K.W. at 30 rev/min to 0.5 K.W. at 70 rev/min. Above a screw speed of 70 rev/min autogeneous extrusion conditions were established.

b) Drive power input

In most of the trials conducted the relationship between the power supplied to the drive motor and the screw speed was almost linear. In all cases the drive power input increased as the screw speed increased.

From previous work on polypropylene (section 4) it has been established that the power required to turn the screw normally increases, for any given set of operating conditions, if the output or the restriction to flow are increased. Although more power was required to force material over the Dow torpedo, the observed output for screw X (Dow torpedo) was considerably lower than that of screws Y (Pineapple torpedo) or Z (no torpedo), and the net result was that the recorded drive power input for screw X was less than that for screws Y or Z. A comparison between screws Y and Z shows that the output of screw Y was slightly lower than that of screw Z, but the net recorded drive power input was greater because of the restriction to flow caused by the Pineapple torpedo.

Similar drive power inputs were recorded for the extrusion of TPS (grade Bextrene 235) using screw X and a breaker plate assembly type D (120 mesh). In this case the power drive input at 30 rev/min was 26 K.W. and at 90 rev/min it was 64 K.W.

The cost per pound of output of the total power (heat and drive) input for both PS and TPS materials was approximately unchanged in the case of screw X when the screw speed was increased from 60 to 90 rev/min. Therefore the corresponding increase in output at higher screw speeds considerably increases the production efficiency of the plant in this case, provided no extra production personnel are needed to cope with the increased output.

6.3.4. The effect of a 200 mesh screen on output

6.3.4.1. Extrusion of TPS materials

In the early stages of this work on the extrusion of unpigmented TPS materials (grades Bextrene 234 and 235) the results obtained using breaker plate assembly type E (200 mesh) were not reproducible.

A special trial was therefore conducted with the breaker plate assembly and unpigmented Bextrene 235 at a constant screw speed of 40 rev/min. Observations of the output and back pressure were recorded every fifteen minutes for the first eight hours of extrusion through a new 200 mesh screen pack, and the results obtained are shown in fig. 22. This work was repeated three times with new 200 mesh screen packs and on each occasion it was noticed that the output and back pressure were constant only for approximately thirty minutes before the back pressure began to rise and the output to drop. The back pressure in each case rose from approximately 2,000 lb/in^2 to a value between 4,600 and 4,900 lb/in^2 and then quite suddenly it began falling. Meanwhile the output dropped from a maximum to a minimum value and then began gaining. The time it took to reach the simultaneous values of highest back pressure and lowest output, in each case, varied between 2½ and 4½ hours. However, in all three experiments the back pressure and output settled down to constant values of 4,500 lb/in^2 and 210 lb/hr respectively after 24 hours continuous extrusion.

Graph Showing Variation of Output and Back Pressure when TPS (Bextrene Grade 235) is Extruded at a Constant Screw Speed of 40 rev/min. Through a New 200 mesh Screen

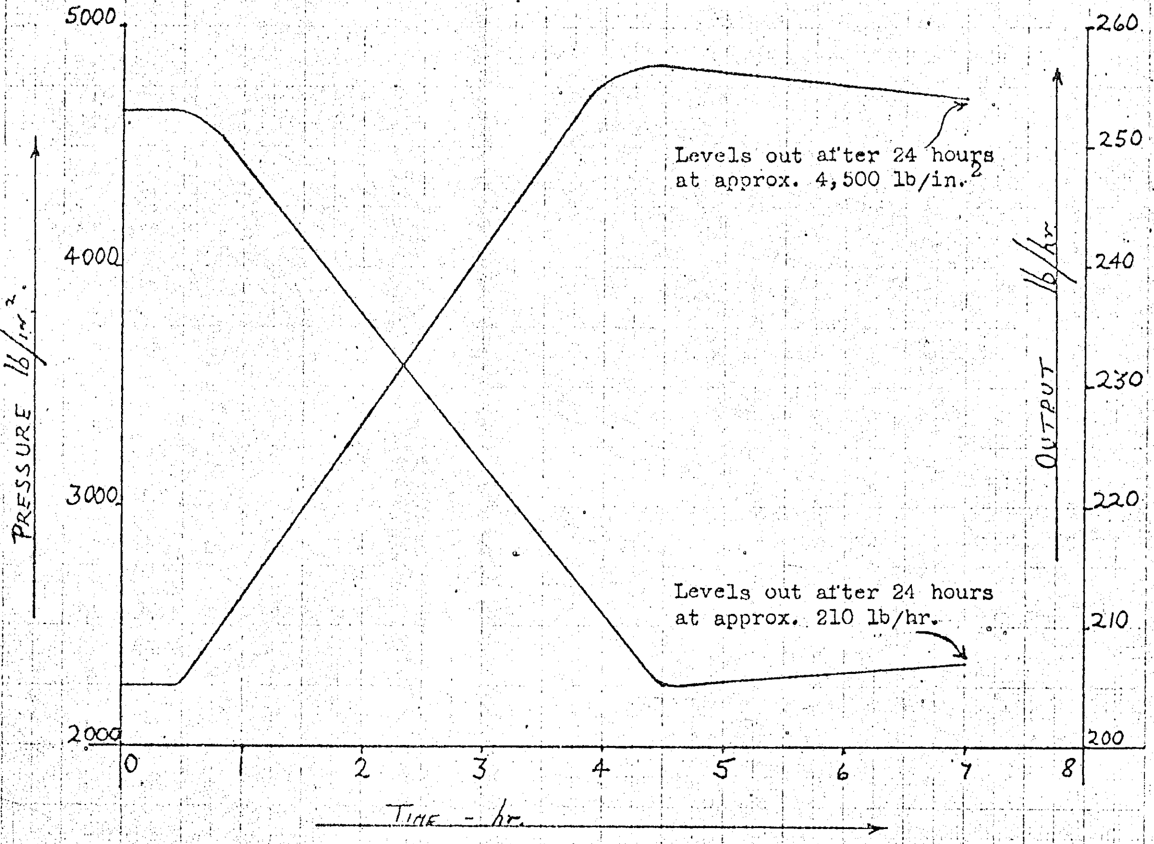


Fig.22.

A further series of experiments, in which 200 mesh screen packs were removed from the extruder after various periods of time and soaked in toluene to dissolve all traces of the TPS, revealed that dirt particles were present. These dirt particles could only have come from the TPS material and they were responsible for the clogging of the 200 mesh screens and the eventual breakdown of some of the wire strands at pressures approaching 5000 lb/in^2. This breakdown continued gradually for approximately 4-6 hours before the back pressure and output became constant.

It is possible that undispersed rubber nibs may help to clog the 200 mesh screen, but because these would also dissolve in toluene their existence was never proved. The time it takes to clog a 200 mesh screen and the highest value of back pressure reached before breakdown occurs will obviously vary, depending on the size and number of dirt particles present, the rate of extrusion and the strength of the screen pack.

For these reasons a new 200 mesh screen pack and breaker plate assembly (type E) was always run in for 18 hours before the observations of back pressure and output etc. given in sections 6.2. and 6.3. were recorded. After this period of time reproducible results were obtained, but of course the fine gauge itself was no longer intact.

Similar results were obtained for another grade of unpigmented TPS (Bextrene 234).

The values of the back pressure developed at various screw speeds for Bextrene 235 if the breaker plate assembly type E (200 mesh) did not clog are given in fig. 20, along with values for breaker plate assemblies type B, C (60 mesh), D (120 mesh) and E (clogged).

This work indicates that 200 mesh screen packs should be changed approximately every 2 hours if they are specifically intended to remove dirt particles of this nature from TPS materials. Otherwise, their only useful purpose is to maintain high back pressures, which are sometimes helpful in overcoming surging or mixing problems.

6.3.4.2. Extrusion of untoughened PS materials

The above phenomenon did not occur when untoughened PS (grade 7000) materials were extruded, even when pigmented, because very few dirt particles were present. Consequently the observed back pressures were very much lower when this material was extruded through a 200 mesh screen pack.

6.4. Visual Analysis Of The Quality Of Mixing Obtained With Screws X, Y And Z.

A visual comparison was made between the quality of mixing obtained using the same three screws (see fig. 14, X, Y and Z). A standard colour formulation (granules of coloured masterbatch plus natural granules) was extruded through each of the three screws in turn at various conditions of back pressure and screw speed. Observations were recorded for the open discharge condition and for breaker plate assemblies type C (60 mesh), D (120 mesh) and E (200 mesh) at screw speeds of 30, 40, 50, 60, 70, 80 and 90 rev/min. However, in order to simplify the presentation of this work only the observations recorded at screw speeds of 30, 60 and 90 rev/min are given.

This work was conducted entirely with untoughened PS (grade 7000) material and only one colour formulation was studied in detail, BXP ref. 151/7070 Pink. The pink masterbatch used consisted of pigments thoroughly dispersed in a natural PS base, and was common to all experiments. Each mix containing 5 parts masterbatch to 100 parts natural PS was tumbled for thirty minutes before being discharged into the extruder hopper.

In each of the following experiments the barrel and die temperatures were preset at the following values

Barrel temperatures °C Zones				Die temperature °C
1	2	3	4	
180	190	200	200	180

Table 44

The pressure transducer measurements were not recording correctly during these experiments because the instrument was faulty, therefore no values of back pressure will be given.

6.4.1. Results

The Dow torpedo screw (X)

Breaker plate ass.	Screw speed rev/min	Recorded temperatures °C						Power KW		Observed output lb/hr.
		Barrel zones				Die zone	Plast. temp.	Drive input	Heat input	
		1	2	3	4					
Open discharge	30	155	180	190	200	180	-	15	8	180
	60	-	-	-	-	-	-	47	5	310
	90	-	-	-	-	-	-	68	2	390
C (60 mesh)	30	160	193	200	208	180	190	20	7	170
	60	185	197	212	222	180	212	47	3	280
	90	200	215	220	225	180	221	68	1	350
D (120 mesh)	30	170	195	210	215	180	201	24	5	160
	60	186	206	215	223	180	221	45	3	255
	90	200	215	220	225	180	232	67	0	335
E (200 mesh)	30	175	196	215	218	180	203	30	5	150
	60	195	207	218	223	180	222	45	3	250
	90	203	220	225	225	180	233	68	0	330

Table 45

The Pineapple torpedo screw (Y)

Breaker plate ass.	Screw speed rev/min	Recorded temperatures °C						Power KW		Observed output lb/hr.
		Barrel zones				Die zone	Plast. temp.	Drive input	Heat input	
		1	2	3	4					
Open discharge	30	145	173	190	200	180	-	-	-	230
	60	-	-	-	-	-	-	-	-	350
	90	-	-	-	-	-	-	-	-	440
C (60 mesh)	30	160	185	190	195	180	185	30	8	215
	60	180	192	195	205	180	195	55	6	330
	90	189	198	200	210	180	205	75	2	415
D (120 mesh)	30	175	192	194	202	180	195	27	7	205
	60	188	203	205	210	180	204	55	5	315
	90	195	210	215	216	180	215	75	1	400
E (200 mesh)	30	175	190	200	201	180	196	27	7	200
	60	190	205	215	214	180	207	55	4	310
	90	200	210	218	225	180	218	75	0	395

Table 46

The screw without a torpedo (Z)

Breaker plate ass.	Screw speed rev/min	Recorded temperatures °C						Power KW.hr		Observed output lb/hr.
		Barrel zone				Die zone	Plast. temp.	Drive input	Heat input	
		1	2	3	4					
Open discharge	30	–	–	–	–	–	–	–	–	–
	60	–	–	–	–	–	–	–	–	–
	90	–	–	–	–	–	–	–	–	–
C (60 mesh)	30	159	190	192	194	180	150-160	28	8	240
	60	163	190	192	192	180	160-172	54	6	340
	90	169	200	200	202	180	175-186	73	3	420
D (120 mesh)	30	185	194	199	203	180	170-180	27	6	225
	60	190	200	205	208	180	183-192	52	4	330
	90	195	210	218	216	180	196-205	72	2	410
E (200 mesh)	30	185	197	201	199	180	174-185	27	6	210
	60	186	206	211	209	180	189-200	51	4	315
	90	195	212	220	217	180	200-210	71	1	400

Table 47

6.4.2. Discussion of results

6.4.2.1. Extrudate appearance

The extrudate emerged as twelve continuous rods from the die, the diameter of these rods varying between approximately $\frac{1}{4}$ inch and $\frac{3}{8}$ inch depending on the extruder output. These rods were normally pulled down to $\frac{1}{8}$ inch diameter before feeding to the dicing machine. It was noticed that the surface of the rods was smoother at low screw speeds than at high screw speeds. Also, the rods had a smoother appearance when finer screen packs were used, for any given screw speed.

The smoothness of the rods was thought to be dependent on three things

a) The degree of mixing of the compound in the extruder

b) The plastic melt temperature

c) The velocity of plastic melt flowing through the die

An increase in the value of (a) and/or (b) or a decrease in the value of (c) gave the extrudate a smoother appearance.

6.4.2.2. Surging of extrudate

In the case of the Dow torpedo screw (screw X) no surging phenomenon was observed and the take-off equipment could cope with the output, consequently all the samples collected in this case were in pellet form. However, in the case of the Pineapple torpedo screw (screw Y) and the screw without a torpedo (screw Z), the take-off equipment could not cope with the extrudate because of surging and/or the increased rate of output. Consequently the samples collected from these two screws were in rod form. A large fluctuation of the plastic melt temperature was observed in the case of screw Z. caused by bad surging.

6.4.2.3. Recorded barrel temperatures and integrated heat input

The recorded barrel temperatures particularly at high screw speeds and high back pressures were equal to the preset values, indicating that the extruder was working under autogeneous conditions. However, the integrated heat input figures showed that although autogeneous conditions were approached as the screw speed and/or back pressure were raised, the extruder was still working under polytropic conditions.

The reason for this discrepancy is that the thermocouples recording the barrel temperatures were situated deeper in the barrel than those controlling the barrel temperatures, and that some heat input was needed to compensate for heat loss outwards from the barrel.

6.4.2.4. Visual evidence of quality of mixing

The reader is asked to extract appendices 4,5,6 and 7 from the back of the report and have at hand for reference.

For every combination of screw design, screw speed and breaker plate assembly shown in the results of tables 45, 46, and 47, a sample of the extrudate was compression moulded into a plaque measuring approximately 6 inch x 6 inch x 0.050 inch. Each plaque was translucent to light and they were each photographed over a strong diffused light source by BXP Physics Department (33 photographs in all). Undispersed pigmented masterbatch shows up in these photographs as dark striations or blemishes, while undispersed clear PS is transparent to light and therefore appears white. The photographs which are most uniform in shade represent better quality of mixing.

To simplify the presentation of the findings only eighteen different photographs are shown. Apologies are made for the quality of these photographs, which do not reproduce accurately the visual contrasts of the actual plaques. It was subsequently found that plaques moulded from rods produced better photographs than those moulded from pellets. However, unfortunately no samples in rod form were collected in the case of the Dow torpedo screw.

To summarise the pictorial evidence an arbitrary numerical assessment of the quality of mixing obtained from each screw based on the author's impressions of the actual moulded plaques is given in table 48 below.

Screw design	Breaker plate assembly	Quality of mixing		
		screw speed rev/min		
		30	60	90
Dow Torpedo (X)	C (60 mesh)	6	6	5
	D (120 mesh)	9	8	7
	E (200 mesh)	10	9	8
Pineapple Torpedo (Y)	C (60 mesh)	3	2	1
	D (120 mesh)	6	4	2
	E (200 mesh)	6	5	4
No Torpedo (Z)	C (60 mesh)	1	0	0
	D (120 mesh)	3	2	1
	E (200 mesh)	4	3	2

Table 48

Where 10 = Very good quality of mixing
7 = Commercially acceptable
4 or less = Very poor quality of mixing

a) The effect of screw speed and breaker plate assembly on quality of mixing

The way in which the quality of mixing of screw Y (Pineapple torpedo) varies with screw speed and breaker plate assembly is shown in appendix 4. Notice that the degree of mixing increases as the resistance of the breaker plate assembly increases (i.e. as the back pressure increases) for any given screw speed. However, the degree of mixing decreases as the screw speed increases, for any given breaker plate assembly.

Therefore, the conditions of lowest screw speed and highest back pressure (i.e. lowest observed output) gives the best quality of mixing, and the case of highest screw speed and lowest back pressure (i.e. highest observed output) gives the poorest quality of mixing.

Screws X (Dow torpedo) and Z (no torpedo) gave similar results.

b) Comparison of the quality of mixing of all three screws at maximum and minimum output conditions

The photographs in appendix 5 show the quality of mixing of all three screws for

(i) the poorest condition of mixing (i.e. max. output)
(ii) the best conditions of mixing (i.e. min. output)

It is apparent from this comparison that the overall quality of mixing of screw X (Dow torpedo) is better than screw Y (Pineapple torpedo) which in turn is better than screw Z (no torpedo).

Only the product obtained with the Dow torpedo screw and breaker plate assemblies D (120 mesh) and E (200 mesh) was commercially acceptable. The observed output of screw X is less than screw Y which in turn is less than screw Z, for any given set of operating conditions.

c) Comparison of the quality of mixing of screws X and Y at various screw speeds

A direct comparison of the degree of mixing obtained with screw X and screw Y using a screen pack assembly type D (120 mesh) at 30, 60 and 90 rev/min is given in appendix 6.

The quality of mixing of both screws decreased with an increase in screw speed, for a given breaker plate assembly.

d) Comparison of the quality of mixing of screws X and Z at various back pressures

A comparison of the quality of mixing of screws X and Z at a constant screw speed of 30 rev/min and at various back pressures (i.e. breaker plate assemblies type C (60 mesh), D (120 mesh) and E (200 mesh) is given in appendix 7.

In both cases better mixing was obtained as the back pressure was increased. A similar effect was observed for any given screw speed up to 90 rev/min. In all cases screw Y gave intermediate results which were nearer the quality of mixing of screw Z rather than screw X. (Compare appendix 7 with the top row of appendix 4).

It is also noticeable that for each screw design a similar quality product can be obtained at an increased output rate, by raising the screw speed and back pressure simultaneously. However it must be remembered that to raise the output appreciably by this method, while maintaining a constant quality product, demands a large increase in the total drive power input.

6.5. Conclusions

1. Screws with or without mixing torpedoes give a simple interrelationship between screw speed, back pressure, plastic melt temperature and output for the extrusion of TPS (grade Bextrene 235).

2. A well designed torpedo and/or an increase in back pressure both help to overcome surging.

3. An increase in torpedo restriction and/or back pressure caused an increase in the work heat generated and this in turn raised the plastic melt temperature.

4. The relationship between the extruder total power input and screw speed is almost linear for the extrusion of untoughened PS on the extruder used.

5. The output of screw X (Dow torpedo) is less than screw Y (Pineapple torpedo) which in turn is less than screw Z (no torpedo) for any given screw speed and breaker plate assembly, because the restriction of each screw is in the same order.

6. For a given extruder operating at a constant screw speed, better quality of mixing can only be obtained at the expense of loss of output. However a simultaneous increase in screw speed and back pressure can increase the output without loss of quality provided the extra drive power required is available.

7. Although the Dow torpedo is a considerable restriction and gives lower outputs than either a small Pineapple torpedo or a screw without a torpedo, it is a very efficient mixing device. By comparison the mixing achieved with screw Y (Pineapple torpedo) and screw Z (no torpedo) was extremely poor. In the case of PS only the product obtained with the Dow torpedo screw and fine gauge packs was commercially acceptable.

8. The quality of mixing increased as the screen pack restriction was increased for all three screws, at any given screw speed.

9. The quality of mixing decreased with an increase in screw speed for all three screws.

10. An increase of approximately 30% in the output of PS and TPS materials was obtained with all three screws when the maximum screw speed of this extruder was increased from 60 to 90 rev/min.

11. Rough ground natural TPS materials were successfully converted into pellets at greatly increased outputs (up to 80% increase) by using a short Pineapple torpedo (screw Y) in place of a Dow torpedo (screw X) and increasing the maximum screw speed from 60 to 90 rev/min. If sufficient demand for these two materials was maintained, converting one extruder in this way may be very profitable.

12. The TPS materials used contained a high proportion of dirt particles which are too large to pass through a 200 mesh screen. These screens are useful for maintaining high pressures which may help in overcoming surging or mixing problems. However, if they are specifically intended to remove dirt particles from these TPS materials, they must be changed frequently otherwise the screen pack will break down under high pressure and the dirt particles will not in fact be removed.

7. PRESENT TRENDS IN SCREW AND BARREL DESIGN

7.1. Conventional Single Screw Performance Characteristics

The optimum screw design varies greatly with die design, extrusion characteristics of the plastics feedstock and extrusion operating conditions such as barrel temperatures, screw speed etc. Under normal production conditions an extruder is required to produce a wide range of products which depend largely on customer requirements and for this reason every screw usually has to be a compromise between several requirements.

This subject has been studied by many workers (1,2,3,4) but usually the conclusions reached are based on very simplified assumptions. In this section a wider and more general account of the effect screw variables have on the extrusion of thermoplastic materials will be given.

In order that the term L/D ratio may be understood in later sections of this chapter, L/D refers to the ratio between the effective screw length (L) from the forward edge of the feed opening, and the barrel bore diameter (D). The length of particular sections of a screw are often referred to in terms of D; all such values quoted in this chapter are based on 25:1, L/D ratio extruders, unless otherwise stated.

7.1.1. The helix angle and pitch of screw flights

7.1.1.1. Variable pitch screws

Variable pitch screws are not very popular at the present time in comparison to constant pitch screws. This is partially because the former, as a consequence of their geometry, usually have deeper flights and a shorter pitch in the metering section compared to the latter. This means that variable pitch screws are more susceptible to changes in back pressure (see section 7.1.4(b)), and also the flight edge effect, as discussed in Part 11, section 3.4., becomes noticeable if the flight depth to width ratio is greater than 1:10.

Variable pitch screws are occasionally used for the extrusion of rigid PVC because shallow flighted screws are liable to generate excessive work heat and give degradation of the product. They are however much more difficult and expensive to make and in view of their disadvantages and limited use the rest of this chapter will be devoted to constant pitch screws.

7.1.1.2. Constant pitch screws

It is normally recognised that constant pitch screws give a more uniform plastic melt temperature and viscosity distribution than variable pitch screws, because the metering section of the former is usually shallower than the latter. Plastic melts have low thermal conductivity characteristics, and the rate of thermal conduction varies inversely as the square of the thickness of the plastic melt.

For a given diameter screw and number of starts the helix angle determines the pitch of the screw, lower helix angles giving smaller pitch but generating a more positive thrust. Darnell (5) has shown that a helix angle within the range 17-20° normally gives the greatest conveying capacity for most granular thermoplastic materials in the feed section of the screw (see Plug Flow Theory, section 3.2.1.). Because of this and the fact that constant pitch screws greatly simplify screw manufacture, the present trend is towards constant pitch screws with the helix angle equal to 17.7° i.e. pitch equal to screw diameter.

7.1.2. The number of starts to the screw

Few multi-start screws are now used because it is extremely difficult to produce screws with identical channels. This has caused serious production problems in the past e.g. surging and lack of uniformity of mixing and plastic melt temperature. In view of their unimportance, multi-start screws will not be considered any further.

7.1.3. Length of screw feed, compression and metering sections

(a) Length of feed section

Compression should be avoided until the plastic feedstock has had time to soften, otherwise excessive frictional work heat may be generated which could cause the plastic to stick to the screw surface near the feed hopper. This in turn interferes with the steady movement of the granules, and could result in a starved or surging throughput. Also if there is a tendency to entrap air, an extension of the feed section often enables the air to escape, this problem is more difficult if fine powders are used as feedstock. For plastics materials such as cellulose acetate and polystyrene in pellet form it was considered normal to have a 1 or 2 D feed section a few years ago with a 15 D screw, but the present trend is towards 4 or 5 D sections with 25 D screws. Longer feed sections are often used with the more crystalline polymers whose temperature needs to be raised almost to the melting point before compression is applied. An increase in the length of the feed section will generally raise the output, the power required by the screw, and the back pressure.

(b) Length of compression section

Relatively little is understood about the compression section, in recent years compression has normally been effected by a constant reduction in the flight depth mainly because of simplicity in the engineering methods of screw production. However, this is not necessarily the best method of producing compression. The general trend for the more crystalline polymers such as nylon and polypropylene has been towards very short compression sections of 0.5 - 1.0 D. In the case of rigid PVC, compression sections extending up to three quarters of the length of the screw are generally recommended, although some vinyl dry blends have been extruded successfully with short compression screws on a $2\frac{1}{2}$ in. diameter extruder (6).

For plastics materials such as cellulose acetate and polystyrene relatively long compression sections of 7 - 15 D have been used for a number of years and this trend appears to be continuing, whereas low density polythene is usually extruded on screws with a 5 - 7 D compression section.

One of the main functions of the compression section is to prevent air from being occluded in the plastic melt. A longer feed section is also recommended for this trouble (but both can be true).

(c) Length of metering section

The usable range of output obtained from a screw with a metering section is usually much wider than that of a screw without metering section, for the extrusion of heat stable melts. Increasing the length of deep metering sections decreases the effect of back pressure and the screw is capable of higher outputs at high back pressures.

With shallow flighted screws, extending the length of the metering section increases the quality of mixing and raises the plastic temperature of the polymer melt, for any given screw speed. In the case of polythene higher outputs as well as a better quality product have been obtained. (6) The length of metering section often used with polystyrene and cellulose acetate materials (excluding torpedo lengths) is usually 4 - 8 D. For heat sensitive materials such as rigid PVC or polyvinylidene chloride the metering section is normally very short, otherwise excessive frictional work heat is generated which can cause degradation of such polymers.

7.1.4. Depth of feed and metering section and screw compression ratio

(a) The depth of the feed section

An obvious limit to the depth of the feed section is the necessity of having sufficient strength in the root of the screw to support the weight of the screw when the extruder is empty and also to transmit the maximum power requirements during extrusion. The feed section must be capable of accepting the feedstock whether it be in powder, pellet, or ground chip form, and conveying it at economic rates without surging. The best feed depth depends on the shape, size and extrusion characteristics of the feedstock.

Increase in flight depth of the feed section will generally raise the output and hence back pressure, and also the power taken by the motor.

(b) The depth of the metering section.

Although deep flighted screws give high outputs at low back pressures, they are more sensitive to an increase in back pressure and therefore often give lower outputs at high back pressures than shallow flighted screws.

Reducing the flight depth in the metering section generally has the effect of increasing the quality of mixing, raising the plastic melt temperature, and reducing the output. For any given output, reducing the flight depth also increases the power required by the drive motor. The increase in quality of the extrudate is often sufficient to permit a greater output of usable product at some higher screw speed, assuming that the additional power required to drive the screw is available.

(c) Screw compression ratio.

The ratio of the volume of the first flight in the feed section to that of the last flight of the metering section is termed the compression ratio of the screw. For constant pitch screws this is the ratio of the flight depths at these two points.

It is possible to have many screws with the same compression ratio but with widely differing extrusion characteristics. Compression can be achieved by several methods e.g. decreasing the pitch, increasing the root diameter, reducing the external screw diameter, or increasing the land width. Screws have often been classified by their compression ratio, although this knowledge without otherscrew dimensions is worthless because other factors of screw design are often more important to quality and output.

The compression ratio of a screw is selected depending on the shape and size of the feedstock particles, the gelling and mixing required, and the thermal stability of the plastic being processed. For most thermoplastic materials the compression ratio is normally between (1.5:1 and 4:1). In the case of polystyrene and cellulose acetate extrusion the screw compression ratio is most likely to be between (2:1 and 3:1) when pelletised feedstocks are used.

Formulations of polymer mixes often greatly effects the optimum screw design and compression ratio. For instance low compression ratio screws of approx. (1.7 to 2:1) are often used for the extrusion of rigid PVC but compression ratios up to 3.5:1 usually give better results with highly plasticised PVC.

7.1.5 Screw length, diameter and L:D ratios

(a) Screw length L

From considerations of engineering construction there are two major limiting factors to maximum screw length when operating in a horizontal plane. The first is the critical figure of the torsional strength of the deep flighted feed section, which is lessened if the barrel is cored for water cooling purposes. The second is the bending of the screw caused by its own weight. Schenkel (6) has calculated that a 12 in. diameter screw 360 in. long and having a total weight of approx. 6 tons will sag by approx. 2 ins. if only supported at one end. Such extruders are currently available, e.g. Barmag-Hartig 30:1 L/D ratio vented extruder with a 12 in. diameter barrel, but little is known about the abrasion of the screw and barrel surfaces or the life of the thrust bearings. To overcome such problems the screw could be mounted vertically but then other problems of gear box and bearing lubrication, strength of thrust bearings, and methods of feeding the plastics material to the screw have to be solved.

(b) Screw diameter D.

The most common screw diameters in use today are $1\frac{1}{2}$, 2, $2\frac{1}{2}$, $3\frac{1}{2}$, $4\frac{1}{2}$ and 6 in. In recent years the general trend in America and Europe has been towards the larger diameter extruders. Extruders with larger screw diameters ranging from 8 - 24 in. have been made and are in use but they are usually designed for special jobs such as preheating, dispersing and/or compounding plastic materials, in the raw polymer state where colour changes etc. are few or absent.

The output capacity of an extruder is dependent to a large extent on the screw diameter. For extruders having similar screw designs and the same L:D ratio a simple rule of thumb relationship predicts that the approx. outputs are proportional to the square of the diameter. Such output ratios for commonly available screws ranging from 1 - 12 in. have been presented in table form by Badonsky (7). Extruder screw diameters up to 6 in. are normally used for the extrusion of plastics sheet, although 8 in. extruders are used in exceptional circumstances.

(c) Screw L/D ratio

Not very long ago extruders with a screw L/D ratio of 15:1 were very popular but the trend in recent years has been towards L/D screw ratios of 20, 25 and 30:1, although many extruder manufacturers insist that intermediate screw sizes such as 22:1 or 24:1 are the optimum screw size. This L/D ratio has been adopted by many as a convenient method of categorising the size of single screw extruders. As we have seen the screw L/D ratio is perhaps now approaching a maximum at 30:1 because of considerations of bending, torsion, processing safety and machine stability.

The major reasons why the screw L/D ratio has increased over the years are:

(i) Longer metering sections give better control at high back pressures with less surging, this in turn gives an extruded product with better dimensional and surface tolerances.

(ii) The shearing and dwell time of plastics materials in the extruder can both be increased, this gives better quality of mixing and a more uniform melt temperature and viscosity distribution.

(iii) The usable range of screw speed is normally much wider and therefore the maximum output potential is often greater, provided enough power can be transmitted to the screw.

(iv) Heat can be introduced to the polymer at a much lower temperature gradient, thus enabling better control while processing heat sensitive materials such as polyvinylidene chloride and PVC.

(v) Well designed extra long screws enable better venting or devolatising to be conducted, thus giving more efficient removal of air, moisture, monomer, plasticiser or solvent from the thermoplastic material. Such screw designs are discussed in detail in section 7.3.

A more recent development in Germany is the extrusion of plastics at very high screw speeds of 600 - 2,000 r.p.m. using extruders with L/D ratio screws of 2 and 3:1. This development is most interesting and is said to have several advantages, but as yet the author has no personal experience with this type of extrusion process.

7.1.6. Effect of screw cooling

If a screw is properly designed for a given plastics feed material there should be no need for screw cooling, particularly if the L/D ratio of the screw is greater than 20:1. However, many screws have to operate with more than one type of feedstock, or have to be used simply because they exist.

The length of the screw cooling channel and the rate and temperature of water flow through the core can greatly affect the screw behaviour. Cooling in the feed section only often overcomes surging and air trapping problems caused by excessive work heat being conducted back to this zone. Some workers have advocated independent water circulation at different temperatures to the feed and metering sections respectively.

In the regions where the material is fluxed the use of cooling water in a screw bore forms a highly viscous layer of plastic along the root of the screw, which has the effect of making the channel depth shallower, and the screw behaves accordingly. Thus, cooling the screw can reduce the output at low back pressures and increase it at high back pressures for any given screw speed. The quality is normally improved provided the water temperature and flow rate are maintained constant, though any fluctuation of either of the latter may cause severe surging troubles.

7.2. Conical Screw Shapes

One type of screw in this category is produced by Battenfeld, the rear two thirds of its length being cylindrical in shape and the front third being conical. Constant pitch screws are made with a L/D ratio ranging between 20 and 26:1 and a compression normally ranging between 2 and 4:1.

The screw acts as a valve and its relative axial position within a smooth bore of similar dimensions determines the back pressure during extrusion. This distance is preset and controlled by a spacing ring at the rear of the screw shaft. This type of screw operates successfully with most plastic materials but it is particularly effective with rigid PVC. In the latter case a larger clearance is left between the screw and barrel surfaces in the conical section to avoid the production of excessive work heat. Unlike conventional single screw extruders, the screw in this case is removed through the back of the extruder for cleaning purposes and the extruder therefore requires extra floor space. However this method does enable quicker cleaning or replacement of the screw because it can be removed before or at the same time as the die is being stripped.

7.3. Single Screw Devolatising Extruders

Development work on this process has been conducted by many companies in the past, e.g. Andouart, Barmag-Hartig, Black-Clawson, Dow, du Pont, Egan, Humbolts, Monsanto, Union Carbide, Welding Engineers and many others.

The most common type of single screw devolatilising extruder is made up of two screws attached in series. The raw material in pellet, powder or flake form is fed to the first screw, and after fluxing in its compression and metering sections passes into the feed section of the second screw. Pressure is released by a sudden increase in flight volume, and frothing occurs as the plastic melt passes into this deep flighted section. The volumetric displacement of this second feed section is normally approx. four times greater than that of the metering section of the first screw. For this reason the feed section of the second screw is often referred to as the decompression or extraction zone. The revolving screw breaks up the froth and releases the gaseous components from the molten polymer. These gaseous components consisting of trapped air, moisture, traces of monomer or solvent etc., are then removed under reduced pressure through a vent opening in the screw or barrel. The plastic melt is conveyed through a second compression and metering section which forces it through a die.

This type of screw is often referred to as a two stage screw but it could equally well be called a six stage screw. To avoid confusion this screw design will be referred to as a two-part screw with each part consisting of three stages or zones. Two to four half inch slots are sometimes cut into each of the last three flights of the first feed zone to prevent the build-up of high back pressures which can cause surging in the first short metering zone.

7.3.1. Extraction via the screw

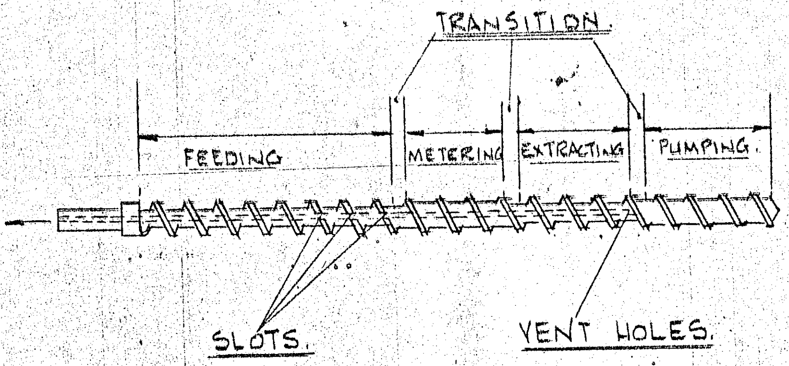
Fig. 23 shows that the extraction hole in the screw is normally situated just behind the trailing edge of the last but one flight of the second feed or extraction zone. It is at this point that the screw is free of polymer melt and froth because of the volumetric relationship referred to earlier. Volatiles are extracted through the hollow core of the screw.

7.3.2. Extraction via the barrel

The best position for the extraction hole in the barrel is likewise situated just before the second compression section, see fig. 24.

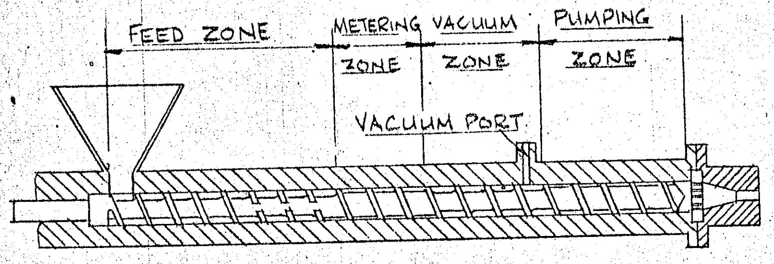
7.3.3. Flow control between the first and second screw parts

Fig. 25 shows an improved devolatising extruder which completely separates the two parts of the screw by the use of a counter thread (i.e. reverse threads of fine pitch and shallow depth). Molten plastic is directed from the metering zone of the first screw, via a by-pass fitted with a valve and pressure guage, and discharged downstream from the vent in the barrel. Another valve and pressure gauge are installed between the end of the



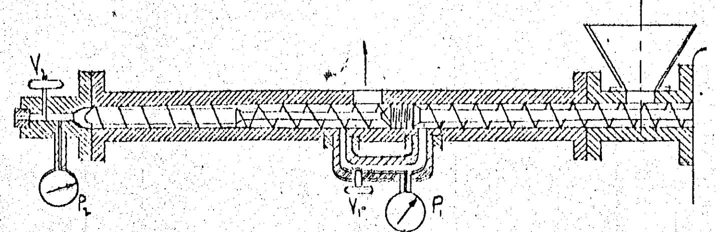
VENTED SCREW

Fig. 23



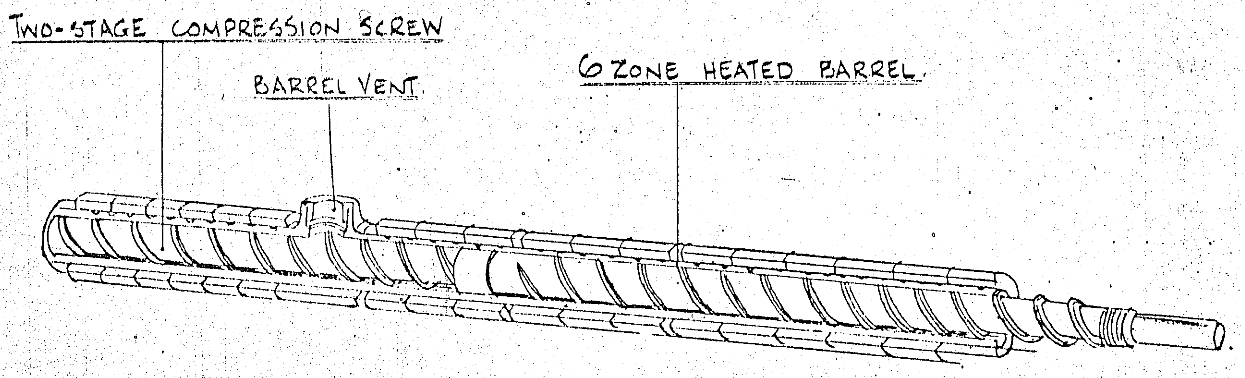
VENTED BARREL EXTRUDER.

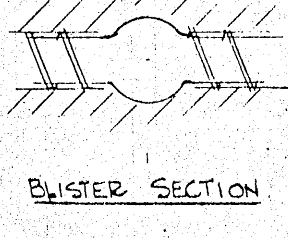
Fig. 24



TWO - STAGE VENT EXTRUDER WITH TWO PRESSURE CONTROLS.

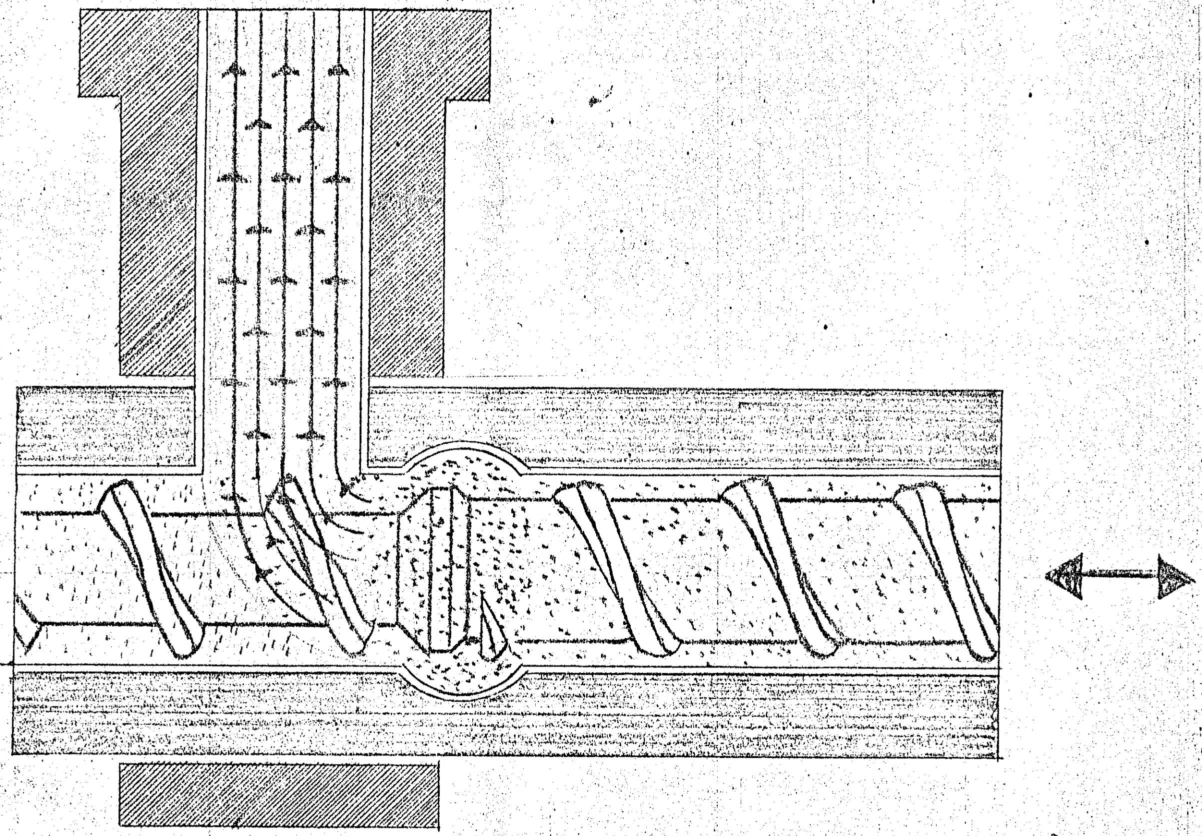
Fig.25





Barmag-Hartig Vent System

Fig.26



Humbolt Vent System

Fig.27

second screw and the die. These valves enable the best pressure balance to be maintained between the two screws, thus giving a better quality product and less chance of material flowing out through the barrel vent. As a result, screw design becomes less critical, and a wider range of materials and dies can be used.

However, this valve assembly is costly and raises many engineering difficulties. It also gives lower maximum outputs because of the increased resistance to plastic flow, and heat sensitive materials are liable to build up and degrade between the first and second screw parts.

7.3.4. Other designs

As shown in fig. 26 Barmag-Hartig force the molten plastic material over a smooth torpedo or blister section approx. one pitch in width, before the extraction zone. Dow have made use of flighted torpedoes leading up to the extraction zone.

Humbolt accomplished a valving effect between the first and second screw parts by using a bulge on the screw which acts in conjunction with a corresponding local increase in the diameter of the barrel as shown in fig. 27. The screw can be moved in an axial direction by manual operation of a hydraulic pump.

Black-Clawson produce a devolatising extruder in which both screws are individually driven, one above the other. The vent in this case is at the point of discharge from the front of the upper screw to the rear of the lower screw.

7.3.5. General discussion on devolatising extruder design

The two part single screw extruder normally has a L/D ratio of approx. 25:1, the L/D ratio of the first stage is often as much as 15:1. Generally barrel extraction systems are more popular than screw extraction systems because they are easier to clean and there is less chance of the vent hole clogging. In cases where the channel depth of the second metering zone is greater than that of the first, it is essential that an adjustable valve control is placed at the outlet end to balance the required flow to the die, otherwise surging will result.

Correct use of devolatising extruders can greatly improve the quality of the product, for instance they reduce the porosity in PVC dry blend extrudates. However, a wide variation of the volatile content in the plastic feedstock can lead to surging troubles.

Two of the best papers on the theory and practice of vented single screw extruders have been presented by Schenkel (8) and Maddock & Matsuk (9).

7.4. Extruder Screw Manufacture

7.4.1. General discussion

The metal workpiece from which a screw is made must be free from flaws, possess suitable machining characteristics and be capable of high surface finish. The screw metal must withstand very high torsional stresses and compressive loads, and must not distort when subjected to normal extrusion temperatures. The screw surface must also be highly resistant to the corrosive by-products sometimes emitted by the plastic mix being extruded. Such a highly polished, non-corrosive finish is usually obtained by depositing a hard chrome layer of approx. 0.001-0.002 in. on the screw surface. It is essential to achieve the highest degree of surface finish by milling alone, as this reduces very considerably the grinding and polishing operations which precede the final chrome depositing process. Parts of the screw which are likely to come into contact with the barrel must possess good non-galling and non-seizing properties.

Constant pitch screws are easier to produce than variable pitch screws, but even so, in the past the initial milling and cutting process was very time consuming because only a thin sliver of metal was removed at each pass. Waldrich Coburg of Germany now produce a milling machine which is capable of cutting a constant pitch variable depth screw in one pass, three such machines are presently being used in West Germany. The author has witnessed this type of milling unit cut a 3 in. diameter screw with a L/D ratio of 26:1 and a compression ratio of 3:1 in 45 minutes.

7.4.2. Common metals used

Usually a screw is made of a heat treated or flame hardened steel alloy e.g. Grades EN19 or SAE 4140, and then chrome plated. Special stainless steels are often used as the base screw metal, some of these are suitable for nitriding, while others have special corrosive resistant, non-galling or non-seizing properties.

Nitralloy is often used to give a screw with an extra hard surface. This metal is an alloy steel containing such elements as aluminium, molybdenum, manganese, chromium, vanadium, titanium, and tungsten which are capable of forming nitrides of extreme hardness when exposed to an atmosphere of nitrogen gas at 510°C for approx. 48 hrs, at which temperature the metal does not distort. All heat treating and milling must be done before nitriding because after this surface hardening process it is very difficult to further modify the screw design. The screw core has a high tensile strength in addition to an extremely hard surface skin approx. 0.010-0.050 in. deep. A recent paper by Naturman (10) discusses fully the techniques used for "Hard Surfacing Extruder Screws".

Stellite is another special metal which is often used for making the flight lands extra hard. Stellite is a cast non-ferrous alloy consisting of approx. 12-17% tungsten, 30-35% chromium, 2.25-2.27% carbon and the balance mainly cobalt. Thin layers of this metal are extremely hard but have practically non extensibility, or flexibility, therefore the base screw must not twist under the operating load or these hardened flight tips will crack.

7.4.3. Special metals used

Special metals such as Z-Nickel or Hastelloy 'B' are usually recommended to avoid catalytic decomposition of polyvinylidene chloride, which will degrade the product at normal processing temperatures. These metals are very resistant to corrosive by-products such as hydrogen chloride.

Z-Nickel or Duranickel as it is sometimes called is a nickel-chromium alloy containing approx. 98% nickel. Screws of this metal even when heat hardened are still liable to twist under normal operating loads but short (12 D) Stellite 'tipped' screws have proved successful at BX Plastics Ltd. under production conditions.

Hastelloy 'B' although more expensive is much tougher than Z-Nickel after heat treatment and gives better resistance to wear, galling, seizing, and hydrogen chloride. Hastelloy 'B' is a nickel, molybdenum, iron alloy containing small portions of carbides.

7.5. Extruder Barrel Manufacture

Extruder cylinders designed for use with thermoplastic materials must be good heat conductors, and from the maintenance point of view they should all be provided with replaceable wear and corrosion resistant liners, which are properly secured against rotational and longitudinal forces. Very close tolerances are required in both the production of screws and liners because the desired radial clearance between the screw flight and the internal barrel surface is often as low as 0.005 in., depending on the screw diameter. Leak flow becomes appreciable for low viscosity melts and fine powdered feedstock materials if the land clearance is greater than 0.010 - 0.015 ins. The liner must expand and contract without losing direct contact with the cylinder, otherwise efficient heat transfer will not be maintained.

Extruder liners are often made of Xaloy 306, the composition of which is approx. 38% nickel; 38% cobalt; 8% chrome; 6% molybdenum; 4% silicon and 3% carbon. The hard Xaloy 306 liners have one disadvantage, they are very brittle and can crack or craze in multiple longitudinal hairfine cracks if strains beyond the limit of elongation are imposed. This can occur if the liners are subjected to very rapid heating and cooling or to excessively high pressures.

Z-Nickel, Hastelloy 'B' and Stoody#6 liners are sometimes used in special cases.

Liners should only be smooth enough to facilitate easy cleaning except in cases where prolonged hold up may cause degradation of the extrudate. Many extruder manufacturers give the extruder bore a mirror-like finish which is not usually desirable because a rougher surface would give higher outputs, see section 3.2., Grooved or rifled bores have been used in the past in attempts to over-come problems in the forward movement of feed materials, but their disadvantage,,such as extra cost of manufacture difficulties of cleaning, hold-up of sensitive materials etc., outweigh any advantages and they are very seldom used in modern extruders. These grooves can vary in size and cross-section and usually run in an axial or helical direction.

Depending on screw design it is sometimes preferable to remove the barrel for cleaning purposes rather than the screw. Some extruder barrels are therefore made in sections which are bolted or clamped together, this sectional method of construction also enable screws of different L/D ratio to be used on the same machine. Heating and cooling methods can greatly influence the engineering construction of extruder barrels.

7.6. Present Trends in Twin and Multi-Screw Design

7.6.1. Present trends in twin screw design

With twin screw extruders it is possible to vary the relative rotation of the screws and the depth of intermeshing as well as the normal single screw variables of thread profile, compression ratio and L/D ratio. The main use of twin screw extruders is the processing of rigid PVC polymers, and this accounts largely for their popularity in Germany and Italy compared to the U.K. and U.S.A., in line with the greater development of this material in these countries.

There are three possible rotational arrangements which two screws may have, these are shown in fig. 28 (a), (b) and (c).

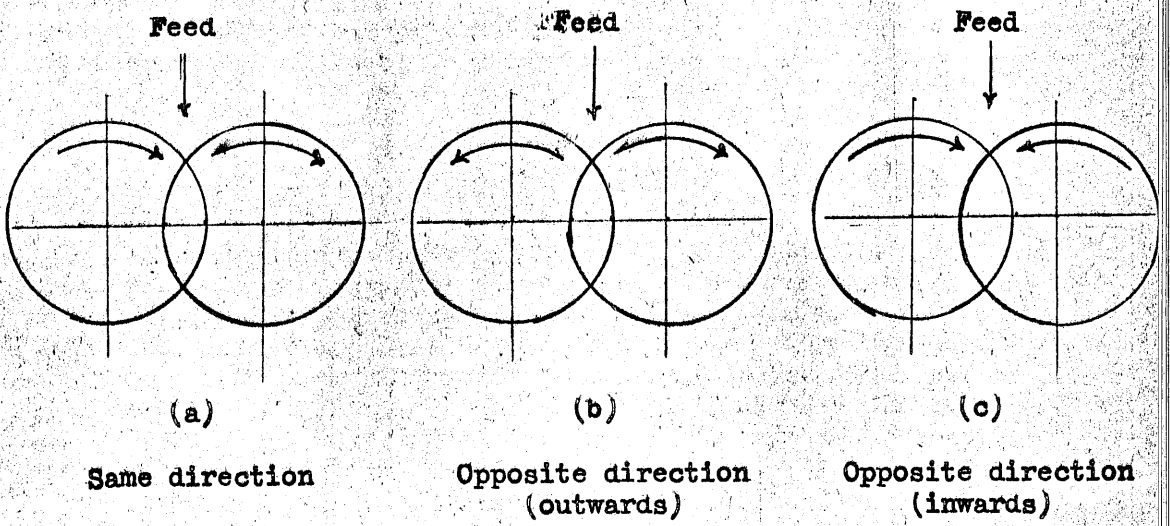


Fig. 28

Prior to 1950, there were eight major twin screw extruder manufacturers, namely Colombo; Boudineuse Universelle; Eckert & Ziegler; Mapré; Pasquetti; Trudex; Welding Engineers, and Windsor. Today there are approx. twenty-one major manufacturers in Europe and the U.S.A.. Listed in alphabetical order these are Alpine; Anger; Barbiera; Bausano; Berstorff; Colombo (L.M.P.); Derex; Kestermann; Kimbell; Krauss Maffei; Leistritz; Mapré; Moi; Negri Bossi; Nothelfer & Söhne; Officiene Dagor; Schloemann; Trudex; Welding Engineers; Werner Pfleiderer; and Windsor. N.B. Derex machines are made by Dusseldorfer Eisenhutten Gesellschaft; Kimbell market a machine developed jointly by Abbegg of Zurich and Zimmer of Frankfurt; Schloemann are now the sole producers of Pasquetti machines; and Windsor produce Colombo type machines under licence. Mapré and Trudex both derive from the same original company.

The majority of present manufacturers prefer the opposite (outward) rotating screw principle shown in fig. 28 (b). but Colombo (L.M.P.), Windsor, Werner Pfliederer and Krauss Maffei use the system shown in fig. 28 (a).

The system of screw rotation given in fig. 28 (c) is not commonly adopted, presumably because intermeshing screws of this type immediately compact the cold plastic feedstock and thus there is a danger of overloading the motor or thrust bearings. Welding Engineers do use this opposite (inward) rotating screw system but their screws do not intermesh.

7.6.1.1. Screw dispositions

Nearly all twin screw extruder manufacturers position their screws side by side in a horizontal plane, but a few are manufactured with the screws positioned one above the other, e.g. Werner Pfleiderer.

7.6.1.2. Feed arrangements

Twin screw extruders are normally run with a starved feed because of the positive forwarding and compressive action of the screws. Hard polymer mixes such as rigid PVC strongly resist the compression enforced by the screws and therefore can cause overloading of the drive motor or thrust bearings. Normally a variable speed single screw conveyor is used to feed plastic materials at a constant but starved rate into the twin screws. The extruder marketed by Kimbell (see section 7.6.1.10 (e)) is an exception and requires a forced but controlled feed screw system.

7.6.1.3. Number of starts

Single start screws are by far the most common, but twin screw extruders with multi starts are used e.g. Schloemann and Werner Pfleiderer.

7.6.1.4. Screw diameters

The diameter of each screw in twin screw machines presently available range mainly between 1 and 6 in., although Moi do produce a machine with 8 in. diameter screws. Normally both screws are the same diameter. A Boudineuse Universelle extruder was made a few years ago which had one screw approx. twice the diameter of the other, the disadvantages of this system (e.g. screw temperature and barrel design problems) were found to outweigh any advantages it may have had.

7.6.1.5. Screw length and L/D ratios

A few twin screw extruders are made which, because of considerations of thrust bearing design, possess one screw longer than the other, e.g. Leistritz, Mapre and Welding Engineers.

L/D ratio is constantly increasing and the majority of twin screw extruders presently available have a L/D ratio ranging between 10 and 15:1. Berstorff; Nothelfer & Sohne; and Leistritz all make production machines with L/D ratios of approx. 20:1 Overall, Welding Engineers make the largest machine with 6 in. diameter screws and a L/D ratio of 25:1.

7.6.1.6. Compression ratio

The compression ratio of twin screw extruders varies considerably from a negative compression 1:3 used in Kimbell extruders to a positive compression of approx. 5:1 used in Colombo machines. As this range indicates, each manufacturer favours a different system (some of which are discussed in section 7.5.1.10), and there is little or no uniformity between manufacturers on twin screw design. The compression ratio can be achieved by a number of methods and a few of these are well illustrated by Fisher (11) and a German publication of B.A.S.F. (12).

7.6.1.7. Maximum screw speeds

Screw speeds of twin screw extruders are relatively low in comparison with most single screw extruders, because of the more positive forward transfer of the polymer. The maximum screw speeds for intermeshing $3\frac{1}{2}$ in. diameter screws varies considerably with each extruder design but it is normally between 20 and 60 r.p.m. Welding Engineers non-intermeshing screw machines, however, have maximum screw speeds of 150 r.p.m.

7.6.1.8. Maximum output of rigid PVC

The maximum output of rigid PVC materials given by 4 - 8 in. diameter twin screw machines varies considerably, depending on screw and extruder design, but it is normally between 300 and 1000 lb/hr. Welding Engineers on the other hand claim outputs of up to 4,000 lb/hr. (13)

7.6.1.9. Polymer path

The polymer path through the screws is normally continuous figure-of-eight, and it is the transference from one screw to the other which gives the polymer its positive forward movement and compounding action, and permits high and uniform die pressures. At the point of intermesh the melt closest to the barrel wall is transferred to a position closest to the screw root, and it is subjected to large shearing forces.

Unlike single screw extruders the barrels of twin screw extruders should be smooth and highly polished to enable more material to be taken round by the screw flights and pass through the points of intermesh. This gives better mixing and compounding of the polymer and lessens the chance of cold material being forwarded to the die. Because of the severe mixing and compounding action between the two screws, devolatising is generally considered to be more complete in twin screw extruders compared to single screw extruders, and it is therefore commonly used.

7.6.1.10 Some twin screw systems

(a) Screws which fully intermesh and rotate in opposite directions must be dimensionally identical to one another in all respects except for the fact that they possess opposite hand threads e.g. Trudex, Pasquetti and Mapré. Screws which fully intermesh and rotate in the same direction must be dimensionally identical to one another in all respects. e.g. Colombo.

(b) Single start, deeply intermeshed, square cut screws with constant pitch but increasing flight land width towards the die. The last few flights have very little clearance at the points of intermesh (Mapré principle).

(c) A number of separate zones with deeply cut, constant depth, multi start intermeshing screws. The screw flights in each separate zone have constant helix angles, but the helix angle of each zone is smaller as the die is approached (Pasquetti principle).

(d) Conical shaped intermeshing screws with flights of constant helix angle, depth and width but with diminishing root diameters. This principle is not presently used although the angular displacement of each screw would enable larger thrust bearings to be fitted than those possible with the parallel system.

Andouart produce an extruder with two converging screws and barrels which is used for two colour extrusion work, but these can be regarded as separate extruders, with separate drives which can be synchonized.

(e) Intermeshing constant depth screws with a constant or progressively increasing pitch in the first zone giving constant decompression. At the outlet end of the screws there are a short series of notched discs which provide the mixing and compounding action (Trudex or Abegg & Zimmer (Kimbell) principles, respectively).

(f) The diameter and helix angle of tapered flights decrease in separate zones between the feed hopper and the die (Colombo principle).

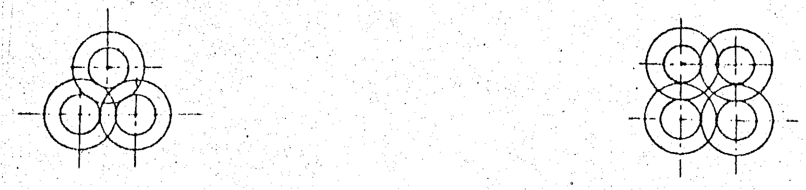
(g) Other positive compression twin screw machines specially designed for good mixing and compounding are discussed in section 7.7.2.

7.6.2. Multi screw extruders

These machines could be useful as high output compounding units starting from powdered feedstock. They are also capable of developing high back pressures and good mixing characteristics with relatively short L/D ratio screws.

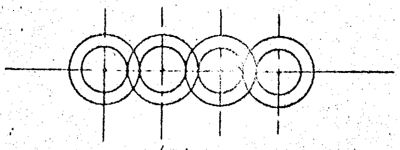
A number of possible designs for multi screw extruders are shown in fig. 29.

Possible Multi Screw Arrangements

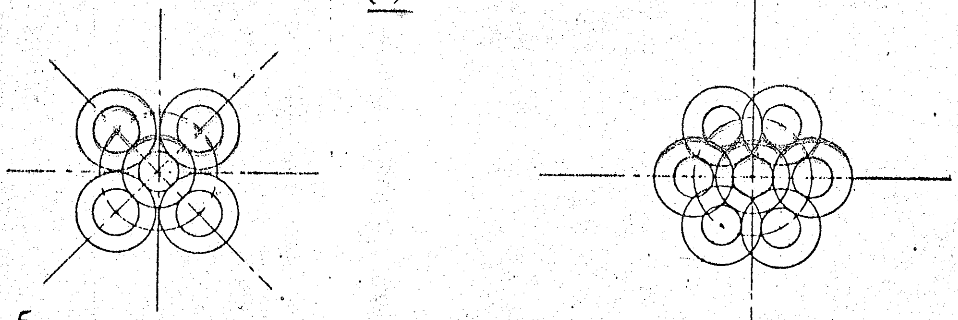


(a) 3- SCREWS.

(b) 4- SCREWS.

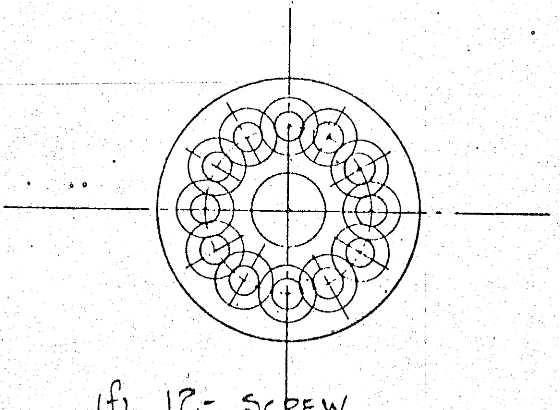


(C)



(d) 5- SCREWS.

(e) 7- SCREWS.



(f) 12- SCREW.

Fig.29

The five screw arrangement shown in fig. 29 (d) has been made in the past. At the present time only two of these designs appear to be available. They are fig. 29 (b) and fig. 29 (e).

7.6.2.1. The four screw extruder fig. 29 (b)

This extruder is manufactured by Jumex, all the screws are of equal length and diameter and posses L/D ratios of 10:1 The screws are fully intermeshing and rotate in opposite directions (I.e. screws situated diagonally opposite to each other rotate in the same direction). Extruders are available with individual screw diameters ranging from approx. 2 - 5 in.

7.6.2.2. The seven screw extruder fig. 29 (e)

This extruder is manufactured by Schalker, in this multi screw arrangement each of the six small planetary screws in the first zone intermeshes only with the seventh small central screw. The six planetary screws form a circle with an outside diameter of approx. $3\frac{1}{2}$ in. In the second zone the small central screw is enlarged to a $3\frac{1}{2}$ in. diameter metering screw. It is claimed that this extruder combines the advantages of multi-screw.positive feed and good mixing with single screw metering characteristics. As yet little has been published about the performance of this extruder.

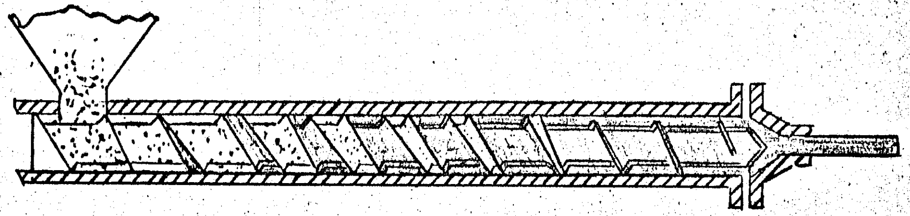
7.7. Recent Trends in Mixing and Compounding Extruder Design.

Two single screw and three twin screw units specially designed to give good mixing and compounding of plastic feedstock are discussed here. Each unit tackles the problem in a different way and yet all five are basically well designed. The working principles of these extruders is shown in figs. 30, 31, 32, 33 and 34. The problem is one of breaking down solid matter in the plastic melt and producing sufficient shear to give good dispersion characteristics, at uniform high output rates, with one pass through the extruder.

7.7.1. Single screw mixing and compounding units

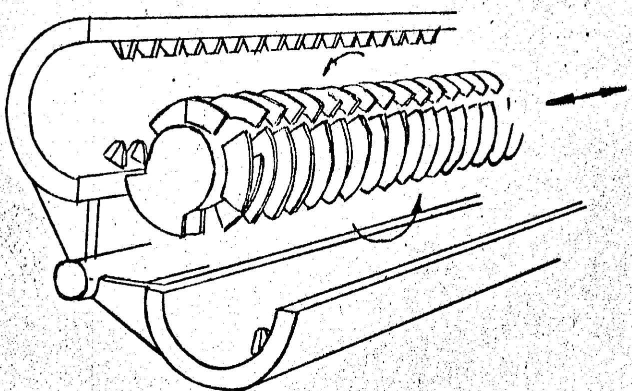
7..7.1.1. The Maillefer extruder

The working principle of this extruder is shown in fig 30 and it has been described fully in a technical journal (14) Briefly the screw has two starts, comprising an entrance channel and separate exit channel. The extrance channel starts at maximum pitch width and depth and reduces to zero. Meanwhile the exit screw channel starts from zero a little way forward of the feed hopper and develops into maximum dimensions as the entrance channel diminishes.



Maillefer Extruder Design

Fig. 30

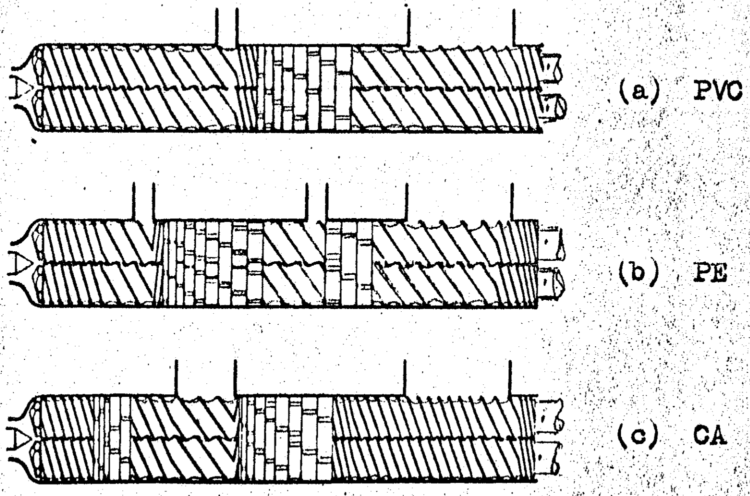


Buss Ko-Kneader Design

Fig. 31

Werner Pfleiderer Compounding Extruder

This diagram shows some of the recommended screw and barrel 'set-ups' for compounding three different plastics materials.



Material flow in kneading sections

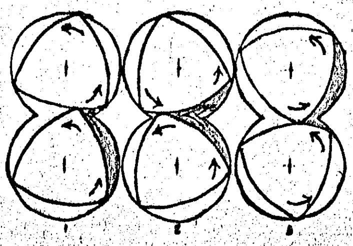


Fig.32.

Krauss-Maffei Compounding Extruder

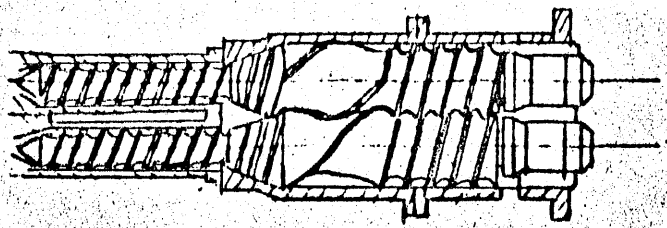
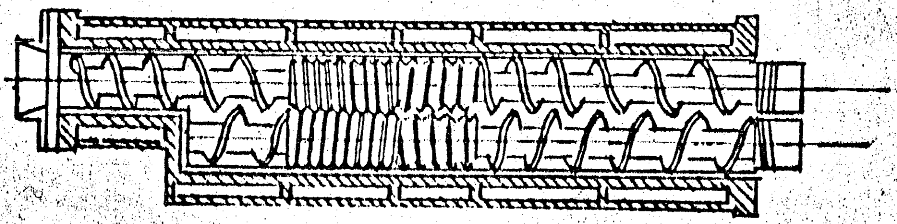
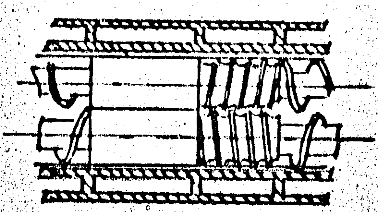


Fig.33.

Welding Engineers Compounding Extruder





Extraction section incorporating a two-roll mill screw design.

Fig.34.

Material can only pass from one channel to the other through the clearance gap between the flights lands and the barrel. These land clearances are slightly greater than those normally used with the conventional screw but are still so narrow that only molten plastics materials can pass through. Cold or imcompletely fluxed materials is thus compelled to remain in the entrance channel until it reaches the necessary fluidity and passes under high shear conditions into the exit channel, thus ensuring that every part of the material has been fluxed and smeared through a narrow gap.

Screw cooling is unnecessary and the percentage of total power lost by heat conduction along the screw is claimed to fall from a figure of approx.44% for the conventional screw to approx. 14% for the Maillefer screw. Better quality of mixing and higher output rates are also claimed.

These extruders are available in screw diameters ranging from $3\frac{1}{4}$ - 6 in. and a L/D ratio of 15:1.

7.7.1.2. The Buss Ko-Kneader

A diagram of the Ko-Kneader in the open position is given in fig. 31 and the working principle has been described in various technical journals (15, 16). Briefly, the flights of the screw are slotted at regular intervals and the screw besides rotating also reciprocates axially. The extruder barrel contains three longitudinal rows of projecting teeth which intermesh with the slots in the screw flights during the reciprocating motion. Plastics material is thus transferred back and forth from one flight to the next and a very intense mixing and kneading action is obtained. The machine is provided with a number of zones which can be separately heated or cooled. The barrel is hinged and can be opened longitudinally, which greatly facilitates cleaning.

One disadvantage of this working principle is that a regularly pulsating or surging output is obtained often containing much trapped air. The problems of volatile extraction and the production of regular sized pellets can be overcome by the use of a short crosshead extruder, fitted with a multi orifice die and face cutting equipment. Volatile gases are extracted through a vent hole in the barrel of the small crosshead extruder, situated on the drive side of the plastic melt intake.

These machines are available with diameters ranging from approx. 2 - 12 in. Outputs of up to 2,000 lb/hr. are claimed for rigid PVC with a 12 in. diameter machine.

7.7.2. Twin screw mixing and compounding units

7.7.2.1. The Werner & Pfleiderer compounding extruder

A diagram of the type of screws used in this machine is given in fig. 32 and the working principle has been described fully in various technical journals (17, 18, 19). This is the most versatile compounding unit manufactured in recent years and consists of two screws positioned vertically one above the other. Both screws rotate in the same direction and are built up from various interchangeable flighted screw bushings of variable pitch, depth and length, and also triangular kneading discs of variable thickness. These sections are keyed on to the screw shafts and thus any conceivable screw design can be used with any sequence of compression and decompression ratios to aid the removal of volatiles.

Screw bushings with narrow channels of reverse pitch are often used to give even better mixing of the plastic melt. The melt is forced through a multi orifice die and is normally pelletised by a face cutting technique. The barrel is removable.

It has been claimed that this machine, with $3\frac{1}{4}$ in. diameter screws and a L/D ratio of approx. 12:1, is capable of screw speeds up to 150 rev/min and outputs of up to 600 lb/hr. for rigid PVC.

7.7.2.2. The Krauss-Maffei compounding extruder

A diagram of the screws used in this machine is given in fig. 33 and the working principle has been described fully in various technical journals (20, 21). This compounding unit combines the functions of a twin-screw extruder, an internal mixer and two single screw extruders. The two screws are parallel and rotate in the same direction. The barrel can be removed in sections.

The diameter of the screws in the feed and compounding sections are approx. 12 in. tapering down and separating into two metering screws each approx. 6 in. diameter. The entire screw assembly may be moved longitudinally so that the volume of the conical compression zone is adjustable and permits control of the compression and shearing forces in the preceding kneading and mixing zones, according to the requirements of the plastic mix being compounded. Volatiles or trapped air can be removed from decompression zones in each of the two separate screws, the vent holes in each barrel are situated just in front of the conical compression sections.

Screw speeds up to 80 rev/min are recommended and outputs of up to 1,000 lb/hr. are claimed for rigid PVC.

7.7.2.3. Welding Engineers compounding extruder

A diagram of the type of screws used in this machine is given in fig 34 and the working principle has been described by Street (22). The two screws rotate in opposite directions (inwards), they do not intermesh but are tangential to each other. Decompression zones where volatiles are removed can be incorporated and the screws frequently include a section designed like a two-roll mill which can produce a rolling bank in an extraction zone.

A reverse multi-flighted compounding section with a L/D ratio of approx. $1\frac{1}{2}$:1 also often gives excellent compounding. To extend the flexibility of these compounding and devolatilising units numerous barrel sections with different vent systems are used.

Screw diameters range from 2 - 6 in., screw speeds up to 150 rev/min are obtainable and outputs of up to 4,000 lb/hr. have been claimed for the 6 in. compounding unit. These units are very long with a L/D ratio of 25:1 thus enabling the use of multi vent zones.

7.8. Advantages of Single and Twin Screw Extruders

7.8.1. Advantages of single screw over twin screw extruders

(a) More flexibility in the choice of drive arrangement, size of driving gears and shaft diameters. Simpler and more robust thrust bearing assemblies.

(b) Barrel is cylindrical and therefore easier to make, also gives more uniform heat transfer.

(c) Normally gives a more uniform flow distribution of material through the adaptor section and into the die head.

(d) Longer screws and barrels permit more heating zones and therefore the temperature gradient along the barrel can be more gradual.

(e) Capital outlay is usually considerably lower.

7.8.2. Advantages of twin screw over single screw extruders

(a) The material flow path in a single flight is longer and therefore L/D ratios can be made comparatively shorter.

(b) Better mixing and compounding if well designed, particularly with incompatible mixes such as wax and resin.

(c) Gives more positive mass transfer

(d) Venting methods are considered to be more effective and multi-stage extraction zones are possible.

(e) Better feed characteristics for fine powders.

(f) Less chance of material sticking to the screw and causing surging.

(g) High back pressures can be built up, if desired, over a short screw length.

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8. DIE DESIGN

8.1. Introduction

The theory of die design was not investigated in detail because of the problems outlined below in sections 8.2 and 8.3 and because altering the dimensions of a large sheeting die to obtain quantitative experimental information is finacially very expensive. Also in the time available it was not possible to conduct a quantitative study. However useful qualitative information was accumulated and a practical problem concerning sheet die widths was successfully solved.

8.2. Theory of Melt Flow Through Sheet Dies

The rate of flow for a Newtonian fluid through a die of any shape is directionally proportional to the pressure drop across the die and inversely proportional to the viscosity of the fluid extruded.

This relationship is given by equ. (20)

$$Q_D = \frac{k \quad \Delta P_D}{\mu} \qquad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (20)$$

Where

Q_D = Output through die in^3/sec

k = Die constant in^{-3}

ΔP_D = Pressure drop across the die lbs/in^2.

μ = Fluid viscosity $lbs\text{-}sec/in^2$.

Values of the die constant for dies of various simple shapes e.g. circular holes, conical sections, parallel slit, tapering slits, any parallel channel, annulus etc. have been calculated by various workers (1,2,3.)

However, plastics materials do not generally behave as Newtonian fluids because their viscosity is not constant for a given temperature, but depends on the rate of shear. The value of the rate of shear varies according to the velocity of plastic flow through the die and the die geometry. Expressions for the rate of shear in dies of various shapes have been calculated (2). The shear rate increases rapidly with a decrease in die opening for a constant output.

Because of this variation of viscosity with shear rate mathematical solutions of the die equation usually make use of either the 'apparent viscosity' concept (1) or the power law (1,2).

The 'apparent viscosity' is defined as the shear to the stress at the die wall divided by the apparent shear rate.

The power law usually takes the form

$$\frac{dv}{dr} = C \, S^n \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots(40)$$

Where

v = the velocity of the fluid in the direction of flow at any point in the moving melt (in/sec).

r = distance perpendicular to the flow direction increasing in the direction of increasing shear stress (in.)

S = the shear stress at any point r in the fluid (lb/in^2).

$\frac{dv}{dr}$ = shear rate (sec^{-1})

n is the slope of log $\frac{dv}{dr}$ plotted against log S, C is the intercept.

The power law usually gives more accurate solutions of the die equation because it involves an empirical relationship which fits the viscosity curve of plastic melts better. Log - log plots of shear stress against shear rate are nearly linear over limited ranges, and the value of n normally lies between 1 & 4, for most thermoplastic materials. For the Newtonian case n = 1, when $c = 1/\mu$.

Several workers (1,2,3,4,5,6) using either of the above concepts of viscosity have attempted to describe the flow of plastics materials through a tube of circular cross section and other simple die profiles, and in some cases quite good agreement between theory and practice has been obtained.

For more complex shapes such as fishtail dies and manifold dies which are used for the production of plastic sheet the mathematical solution becomes very complicated and theoretical results do not usually compare favourably with practice. The plastic melt temperature and die temperature are obviously very important variables when considering the die performance characteristics. In a sheet die the melt temperature, pressure, shear stress and shear rate can vary considerably in different parts of the die e.g. the shear stress in the manifold is small compared to that at the die lips.

In order to keep mathematical derivation for plastic flow through sheeting dies within workable limits it is normally assumed that the plastic melt temperature is constant at all points within the die. This is very nearly so if the die is maintained at the same temperature as the plastic melt. Also it is assumed that the work done in shearing the material as it flows through the die produces a negligible temperature rise.

With centre fed manifold and fishtail sheet dies the extrusion rate at the feed entrance to the die always exceeds that at the far edges. Equations have been derived which attempt to account for the latter effect (1,7), but even in cases where the channel and slit dimensions of a sheet die are fixed the theoretical solution is very complex and often unreliable.

Plastic melts exhibit elastic memory properties in addition to viscous properties and thus the shape of the extrudate emerging from the die is usually distorted.

Consequently the design of sheet dies is based on past experience and trial and error, it is rarely if ever based on theory. For these reasons and those of the next section (8,3) practical solutions based on the theory of sheet die design were not sought.

8.3 Elasticity Properties of Plastic Melts

It is the combination of high velocity and elasticity characteristics which gives polymers unusual flow properties, turbulent flow being observed at low flow rates in comparison to Newtonian fluids. Melt elasticity accounts for the swelling usually observed when the plastic melt leaves the die, and also the memory or weld lines which occur when the melt has been divided and rejoined in the die, e.g. through the use of spiders in film die design.

Problems relating to swelling, melt fracture, streamlining and turbulence, surging and evenness of flow of plastic extrudates have been studied by various workers (8,9,10,11,12). In all cases these problems were found to be related to velocity of extrusion and/or the melt elasticity characteristics. Design of a die limits both the quantity and the quality of commercially usable product obtainable from it. Quality may be limited by surface defects which can be caused by various extrusion problems e.g. volatile constituents, nibs in the polymer, poor die finish, dirt particles, etc. Excluding these troubles from external sources normal extrusion difficulties may be overcome as follows:

8.3.1. Swelling This is sometimes overcome by increasing the dielip land width, which allows certain stresses to relax before the melt leaves the die. Swelling can also be overcome by reducing the output or raising the melt temperature. Often the desired extrudate shape can be achieved by pull-down techniques or by allowing for the distortion effect in the initial die design e.g. presetting the gap between the sheet die lips to a smaller value than the thickness required.

8.3.2. Melt fracture and surface roughness

Roughness is sometimes overcome by increasing the melt and/or die temperature or by reducing the dielip land width. Decreasing the shear experienced in the adaptor between the extruder and the die may also help to overcome surface roughness.

Melt fracture only occurs at or above a critical stress or extrusion pressure, increasing the die opening or reducing the throughput may also help to overcome surface roughness or melt fracture.

8.3.3. Streamlining and turbulence For good quality extruded sheet at high output rates the die design should be as streamlined as possible and all sudden changes of flow path should be avoided. High shear rates usually cause turbulence and result in a wave forming or knobbly output, while still higher rates will result in melt fracture.

Dies should be streamlined in order to avoid the hold up of material in'dead spots'. A colour change followed by dismantling of the extruder die will often show up 'dead spots' which can then be eliminated by the use of metal inserts. Turbulence can sometimes be avoided by increasing the melt temperature, the die opening and the dielip land width, or by decreasing the throughput, and the angle of entry to the die.

Turbulent flow was observed in the case of the extrusion of unmodified PS whenever a screw with a Pineapple Torpedo (see fig. 14.Y) or without a Torpedo (see fig. 14.Z) was used. A Dow Torpedo screw (see fig. 14.X) overcame this turbulent flow effect by raising the plastic melt temperature and reducing the output.

8.3.4. Surging and evenness of flow

Increase in die resistance can either increase or decrease the surging phenomenon depending on the screw and die design but it always reduces the output for any given screw speed. However it must be remembered that an increase in the die resistance may increase the quality so that a greater amount of commercially acceptable material can be produced, although the overall output is lower.

Evenness of plastic flow through a sheet die can be achieved by various methods.

1. Adjusting the die lip gap: this can either be wider in the centre of the die or at the die edges, depending on the size and design of manifold etc.

2. Varying the temperature across the die: this method of checking plastic flow in different sections should be avoided if at all possible, because strains are induced which may lead to uneven surface appearance, uneven reversion characteristics and forming troubles.

3. Adjustable restrictor bar: this method of checking plastic flow is the best method because it enables a constant die temperature and die lip gap to be used and produces a more uniform product.

8.4. Design Of Dies Used For Plastics Sheet Production.

8.4.1. Centre-fed dies

These dies may be sub-divided into two major categories namely:

8.4.1.1. Fishtail dies
8.4.1.2. Manifold dies.

Both types are designed to be used either with or without adjustable restrictorn bars. There are many centre fed dies in existance of widely differing designs. Sketches showing cross sections through the centres of some die designs which incorporate adjustable restrictor bars and which have been used successfully in the past are given in fig. 35, 36 and 37.

Dies which incorporate vertical or Z flow paths (see fig. 36 and 37) possess a shorter die breadth than those with a straight forward horizontal flow path (see fig. 35). Some die designers claim that the former designs enable a more uniform and better temperature control and overcome die and die lip distortion. Reifenhauser recently developed a die with a plastic flow path similar to that shown in fig. 37(b). This die is unusual in that it has a V shaped restrictor bar and not a straight one, and it is claimed to produce extremely accurate thin gauge flexible and rigid PVC sheet over a thickness range of 0.010-0.040 in., but as yet little is known of this new development. However, dies with a horizontal flow path (see fig. 35) are mainly used for the production of TPS and CA thick sheet material because they are usually less expensive to make and maintain and easier to dismantle for cleaning.

8.4.1.1. Fishtail dies

The term fishtail die is often mistaken to mean the same as a "coathanger" manifold die, but for the purpose of this thesis a fishtail die will be regarded as a die not possessing any distinct manifold or deep channel. It is a centre fed die which is precision machined to give greatest resistance to melt flow directly opposite the die entrance, the plastic stream is forced to gradually widen and flatten out in a Y shape as it passes through the die. These dies have proved very successful for the extrusion of PVC sheets.

(a) Without Adjustable restrictor bar

This type of die is very streamlined and possesses a restriction designed to produce uniform plastic flow. The shape and position of this restriction is determined by past experience and trial and error, for any particular plastic material. This type of die is shown in fig.38.

Plastic Flow Paths for Centre Fed Sheet Dies.

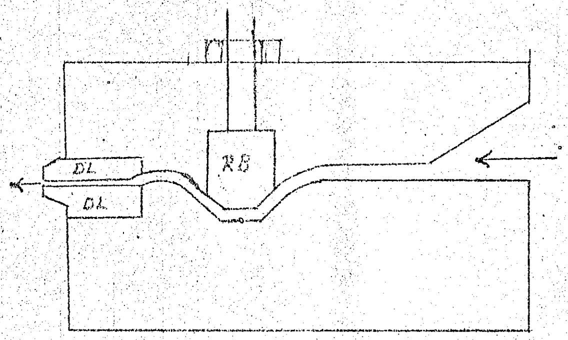


Fig. 35. Horizontal Flow Path .

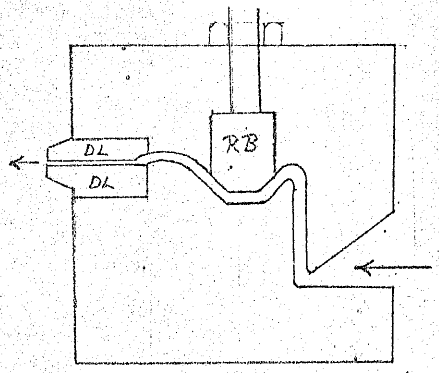
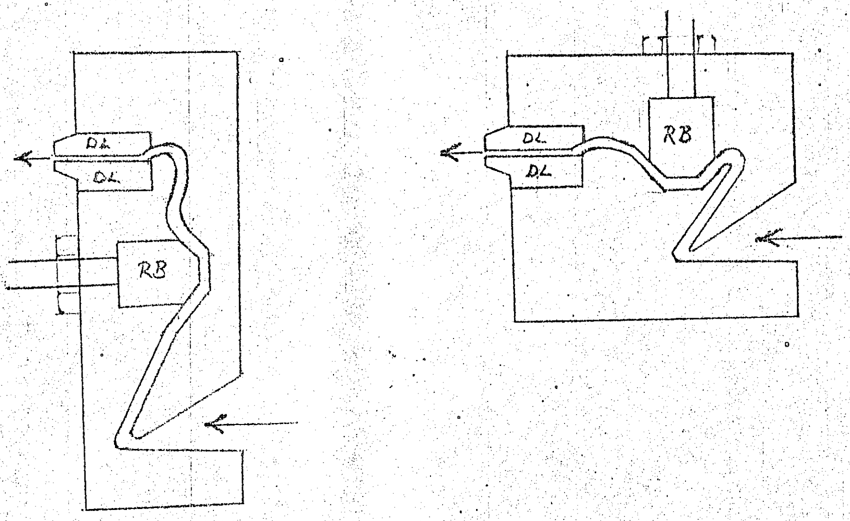


Fig. 36. Vertical Flow Path



(a) (b)

Fig. 37.

Z Flow Path

DL represents die lip

RB represents restrictor bar

Fishtail Die Without Adjustable Restrictor Bar

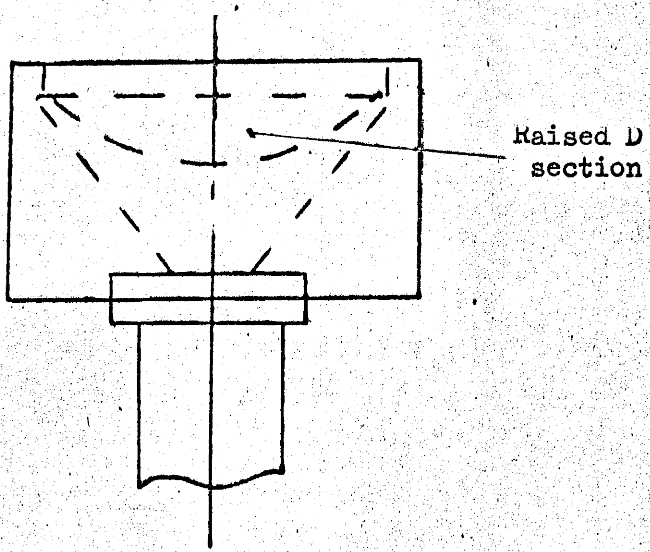


Fig. 38.

Manifold Die Without Adjustable Restrictor Bar

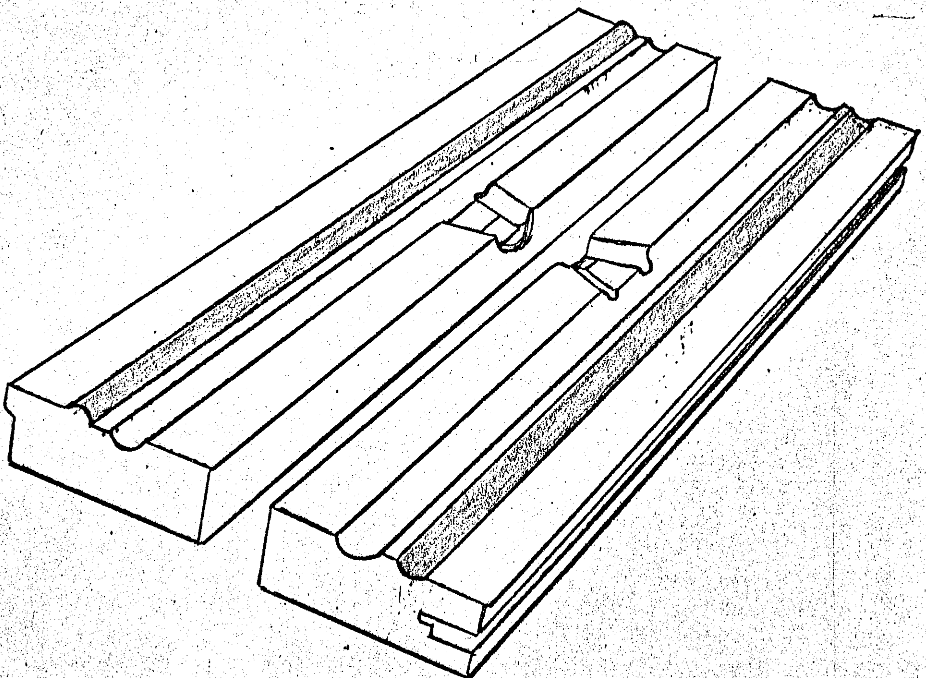


Fig. 39.

Manifold Die Design With Adjustable Restrictor Bar

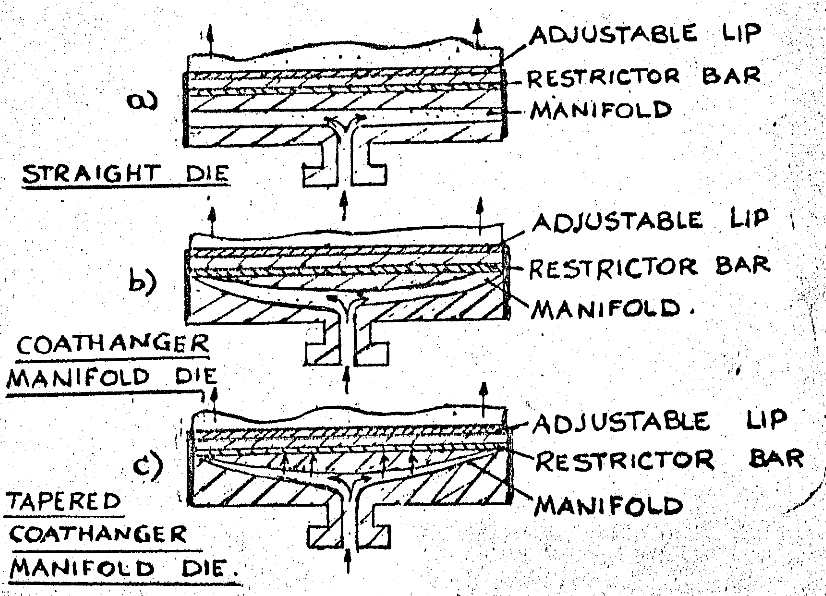
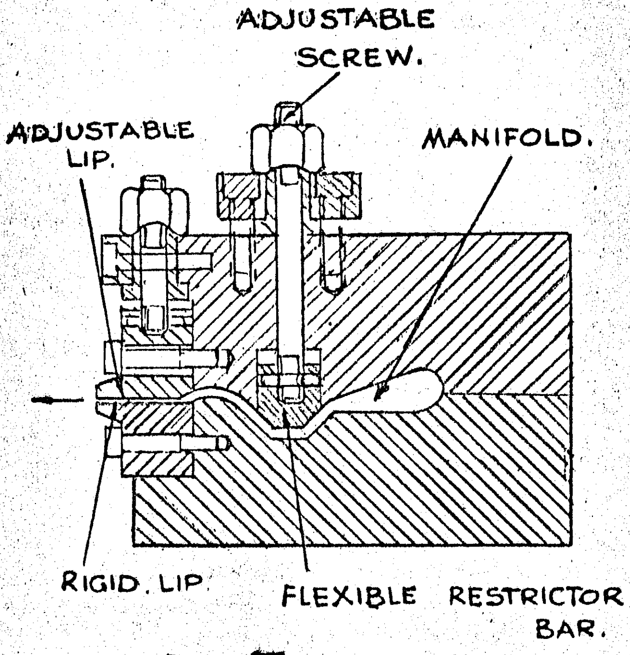


Fig.40.



CENTRE FED TEARDROP MANIFOLD DIE WITH ADJUSTABLE RESTRICTOR BAR

Fig.41.

Such dies are not very versatile because the thickness profile of the extruded sheet can only be controlled by adjusting the die lips and zone temperatures across the width of the die. Often production rates are low in order to obtain good quality sheet. To produce sheets ranging in thickness from 0.040 in. to 0.200 in. it is normally necessary to use two or three different dies and up to 3 sets of die lips are often provided with each die to give better results at varying thicknesses. Valuable time is lost if this type of die is used because they require frequent changing to produce the wide range of sheet sizes normally required by the market.

(b) With adjustable restrictor bar.

An adjustable restrictor bar enables better control of plastic flow before the plastic reaches the die lips (restrictor bar design is discussed more fully in section 8,5,3).

8.4.1.2. Manifold dies.

These fall mainly into two categories:

(1) Straight manifold die

(2) Coathanger die.

Manifold design is discussed in greater detail in section 8.5.2.

(a) Without adjustable restrictor bar.

Fig. 39 shows a sketch of a straight manifold die with a fixed restrictor bar. This type of die is still occasionally used for the production of TPS and polythene sheet, but it is not usually capable of producing a wide thickness range of plastics sheet materials to within the tolerances currently demanded by sheet users.

(b) With adjustable restrictor bar.

These again can either be straight manifold or coathanger type dies plan views of both types are shown in fig.40 (a), (b) and (c).

Both types of die may have a constant cross section manifold but tapered manifolds, although more expensive, give better streamlining and reduce the tendency to form pockets of stagnant material at both ends of the manifold. These dies are frequently used for the production of cellulose acetate and toughened polystyrene sheeting. Such dies are very flexible in the range of sheet widths which they can produce, e.g. single dies capable of producing sheet with a thickness range of 0.060-0.250 in. are common. Well streamlined dies of this type are claimed by Prodex to extrude PVC efficiently.

Centre fed manifold dies have been made which, incorporate a stirrer or screw worm but such dies have not so far been proved to be particularly advantageous and they are not currently available commercially.

8.4.2. End fed dies.

These are manifold dies which usually possess straight manifolds with a reducing circular cross section. They are made with or without restrictor bars and have proved successful in the production of rigid PVC, TPS and CA sheet ranging from 0.010 - 0.040 in. thick.

Recently dies have been produced with rotating stirrers or screws in the side fed manifold to produce a more even flow distribution and to overcome stagnant dead spots. Hagedorn use a tapered stirrer in the manifold with small rounded spikes protruding from it at right angles along its length, driven by a separate motor at the die inlet end. This particular die is set at an angle to the main extruder. Windsor on the other hand have introduced an in-line die with auxiliary rotating screw driven by a separate motor at the far end of the manifold. With both these dies the die lip temperatures can be controlled accurately at values different from the die body. The Windsor die does possess a restrictor bar to control plastic flow but the Hagedorn die relies entirely on the adjustable die lips for good gauge control.

Both dies are claimed to produce high quality rigid PVC sheet but the type of resin extruded in each case is critical. In the case of PVC it is claimed that suspension and bulk polymers are superior to emulsion polymers. The flow properties of the resin should be good and the molecular weight distribution should be within a narrow range and not too high, with K-value not greater than 65. Speciality resins are, therefore, used and the type and quality of stabiliser and lubricant is also critical. Also the scrap to virgin resin ratio often requires careful control.

8.4.3. Circular Dies

This is another method of producing sheet which was initially developed for the production of cellulose acetate and cellulose acetate butyrate sheeting, but is now mainly used for the production of rigid PVC sheet for roofing purposes. The sheet is extruded as a large tube and then slit continuously in a longitudinal direction. This type of die is used in the Columbo process for the extrusion of rigid PVC and a diagram of such a die is given in fig. 42.

This method has the advantage of obtaining symmetrical flow but such dies are normally very heavy and are not designed to withstand pressures greater than 1 ton/in^2. The split tube is usually extruded downwards and flattened with the aid of a spreader bar to the full width of its circumference.

Circular Die Used for the Production of Rigid PVC Sheeting.

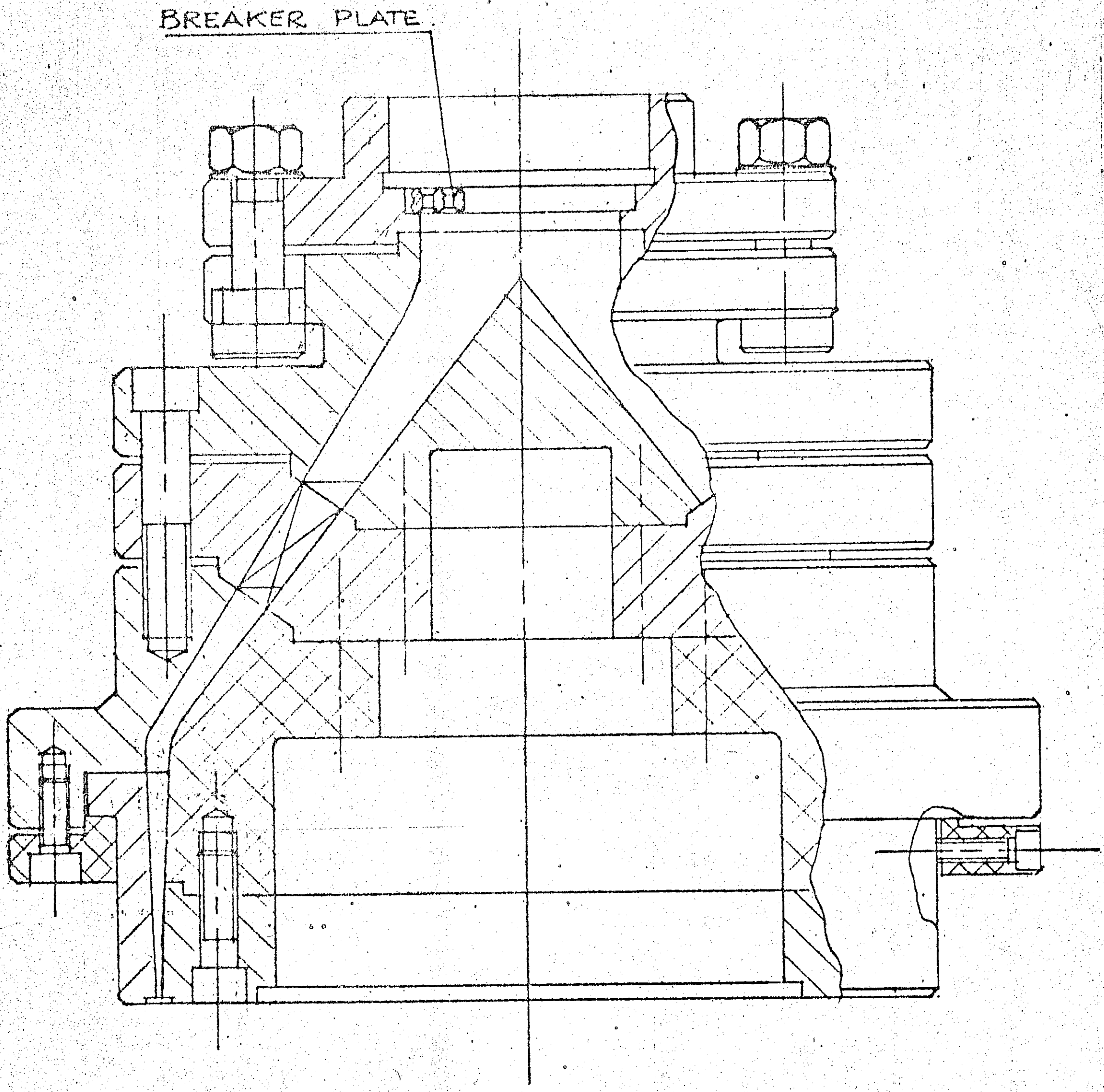


Fig.42.

8.5. Special Features of Die Design.

8.5.1. Design of adaptor section.

The adaptor section connects the die to the extruder. It usually is tapered with an increasing or decreasing circular cross section which matches up to the diameters of the extruder outlet and the die inlet. This type of adaptor is ideal for circular dies but with flat sheet dies better flow conditions are obtained if the adaptor is designed to match up to a die intake channel with an oval cross section.

8.5.2. Manifold design.

The best manifold die designs are highly streamlined. The best flow conditions for toughened polystyrene are obtained with tapering coathanger manifold dies which possess a tear-shaped cross section, see figs. 40 (c) and 41.

These refinements make die manufacture much more costly but they are well worth while particularly if such streamlining overcomes the problem of pockets of stagnating plastic material and increases the tendency to self clean. The latter advantage can produce a great saving both in colour change scrap and in the time otherwise taken to dismantle a sheet die for cleaning purposes.

8.5.3. Restrictor bar design.

An adjustable restrictor bar acts as a type of adjustable valve to obtain an even gauge sheet across the whole width of the die lips, this in turn gives increased flexibility to the sheet extrusion process and makes die design less critical. Such a restrictor bar fits into a specially designed groove as shown in fig. 41 and it is designed to minimize stagnation problems.

Restrictor bars are normally bent into the required shape by alternating push and pull bolts or by one series of bolts which both push and pull depending on the direction in which they are turned. Bolts with finer threads give better control of adjustment but if the thread is too fine there is a serious danger of the bolts seizing. It is usual to situate the adjusting bolts on top of the die assembly though some companies, e.g. Reifenhauser, position these underneath, both arrangements have their disadvantages, in the former case there is a danger of adjustment tools slipping onto the sheet and damaging the polished take off calendering rolls, in the latter case restrictor bar adjustment is difficult and sometimes dangerous for the operator.

Sheet dies which possess wide and shallow restrictor bars rather than narrow and deep ones are known to give better results for the extrusion and toughened PS materials. Some restrictor bars have been divided into a number of separate sections but they were liable to break down and do not help to improve the flow behavour of plastic melt through the die.

Rigid non flexible restrictor bars have been used in conjunction with end fed tapered straight manifold dies, the idea being to control the flow of plastic melt by canting the restrictor bar rather than bending it. Such dies have been used successfully for the extrusion of thin polystyrene sheet but flexible restrictor bars are generally preferred because they give better gauge control over a wide range of sheet thicknesses.

8.5.4. Die lip design.

Die lips are normally cut square or slightly recessed to the face of the die to ensure maximum rigidity and protection from knocks which may seriously damage them. Protruding die lips are occasionally used because they lessen the bulk of the die. The lower lip is usually fixed and the upper one adjustable, if possible it is better to maintain the two lips parallel at a preset value rather than control the flow of plastic melt by adjusting one of them. The adjusting bolts operate in a similar manner to those decribed for the restrictor bar in the previous section.

The combined effect of the restrictor bar and die lips forces the plastic melt to the far ends of the manifold and ensures that the die channels are full of plastic material. The die lip land is normally between 1-2 in. wide. Theoretically, die lip land widths should be long enough to give sufficient back pressure and even flow characteristics, produce good gloss and eliminate swelling phenomena. On the other hand too wide a land will apprecialy reduce the output and may lead to surface roughness problems. Die lips are occasionally heated separately to the main body of the die to produce a better surface finish on the extruded sheet.

Many companies including Bone Bros. and Troester incorporate an expansion chamber between the restrictor bar and the die lips, this is commonly thought to improve melt flow behavour and ensure that the die lips are always full even when thick sheet in the order of 0.25 in. is extruded. The author seriously doubts if this expansion chamber gives any improvement to a well designed streamlined die.

8.5.5. End plates and blanking off plates.

(a) End Plates

It is at the ends of the manifold channel next to the end plates where pockets of stagnating or decomposing plastic melt are most commonly found. This problem is sometimes overcome by attaching permanent streamlined metal inserts to the end plates so that they locate in the dead spots at the end of the manifold channel, or by drilling bleeder holes in the end plates. Flow of plastic melt through these bleeder holes is usually controlled by turning a small bolt acting as a valve. This technique is often used with end fed dies because it helps to reduce the time and plastic scrap involved in colour changes.

(b) Blanking off plates.

These are normally simple brass dams which are bolted flush against the die lips, at each edge, to reduce the width of the plastic sheet being extruded and thus avoid large quantities of edge trim scrap. Plastic behind the blanking off plates will stagnate but provided no decomposition takes place the die may only need purging when a wider sheet is required. Bleeding off the plastic melt behind the dams through the end plates will prevent much of the stagnation.

Plugs which fit inside the die are occasionally used but the die must be dismantled every time the die width is altered and this is a time consuming and therefore costly business.

Both methods of reducing the die width greatly upset the evenness of flow of plastic melt through the die, also neither are satisfactory with PVC because the stagnation problem usually leads to granually increasing degradation throughout the sheet width.

8.5.6. Size of sheeting dies and channel dimensions.

Present markets for plastics sheet normally require sheet which is between 0.01 and 0.25 in. thick. Lately there has been an increasing demand for extruded plastic sheet up to 0.500 in. thick but such markets are as yet small.

Existing markets normally require sheet which is anything up to 60 in. wide, although thick sheets up to 72 in. wide have been made. The four most popular die width sizes are approx. 24 in., 36 in., 48 in. and 60 in.

Large sheet dies can weigh up to 1 ton and they are often supported by trolleys on the same rails as the sheet take-off equipment, or suspended from an overhead hoist. The latter is often useful in assisting the dismantling of large sheet dies. Some dies are housed in large surrounding pressure plates to reduce to a minimum any tendency for the die to bulge under internal pressure of the plastics.

Die channel dimensions depend greatly on the size of sheet required and the plastics material extruded. Most companies have their own ideas on channel dimensions and there is very little standardisation or co-ordination between them on this point. In general it is true to say that the more thought and streamlining which goes into the die design the more expensive it becomes, e.g. a well designed 60 in. wide sheeting die can cost well over £3,000

8.6. Materials Used For Sheet Die Construction.

The die body is normally made from fine grain tool steel and it is often chrome plated on the internal surfaces in contact with the plastic melt. Die lips are usually highly polished and often chrome plated, although sand blasted and smooth hardened steel surfaces are sometimes used. The adhesion of each particular polymer to metal surfaces and the resistance of those surfaces to degradation products plays a large part in die material selection.

The steel chosen for the die lips and restrictor bar must be flexible to withstand the bending induced in it by the push-pull bolts. Very hard grades of steel alloys such as Nitralloy are not recommended because after undergoing the special hardening processes they are very difficult to modify and dies often require modification.

A very soft metal such as brass should be used for gauging and cleaning purposes so that the highly polished metal surfaces are not scratched or damaged in any way. Scratches or tool marks on the die lips and/or restrictor bar can cause surface lines on the extruded sheet, particularly if they run in a longitudinal direction to plastic flow. Tool marks in the transverse direction usually have little effect on the extrudate surface unless they are deep.

8.7. Summary of Die Design.

Die design is dependent to a large extent on screw design and therefore alterations made to die channel dimensions should be conducted on the machine with which it is used. Ideally a die should be designed specifically for one plastic material extruding at a constant die temperature and one thickness and width, so that when the restrictor bar is in the fully open position an approx. even gauge sheet is obtained with parallel die lips. The control of the die temperature is extremely important because different temperatures produce different flow characteristics with any given plastic material.

If at all possible an adjustable restrictor bar should be incorporated into the die to give greater versatility to the process by producing even plastic flow over a wide sheet thickness range. The plastic flow path should gradually decrease in volume and the greatest throttling action should be caused by the restrictor bar.

Well streamlined coathanger manifold dies with adjustable restrictor bars are the most successful for the extrusion of toughened polystyrene and cellulose acetate sheets. Prodex claim that this type of die is also suitable for the extrusion of PVC sheet. A wide range of designs are usually used for the extrusion of rigid PVC sheet, all are designed to give maximum streamlining, avoid stagnation, and obtain evenness of flow. Among the more successful designs have been the fishtail dies with adjustable restrictor bars and the large circular dies.

8.8. Calculation of Optimum Die Width Combination to Suit Sheet Size Requirements

When sheet of one plastics material is to be manufactured in quantities requiring the use of more than one extrusion unit, problems arise in the selection of combinations of die widths which will lead to the most economic operation of the plant as a whole. Each die necessarily produces sheet of a particular width, although with heat-stable polymers it is possible to use adjustable blanking-off plates and thus alter the effective die width. However, each change of die width still involves a production stoppage and much time is lost in making a mechanical adjustment, restarting the extruder and getting "on-gauge". Each start-up is often accompanied by the production of large quantities of off-size scrap and although this is mainly reusable the percentage which can be reworked with virgin material is sometimes limited.

Commercial demand frequently requires a wide range of sheet sizes and these are usually made on a small number of machines whose die widths are either fixed or comparatively costly to adjust. Under these circumstances one of the most important aspects of production efficiency is the amount of edge-trim scrap arising when dies are used to make sheet which is narrower than that for which they were designed or adjusted.

If the pattern of sheet size requirements is known in advance or can be reasonably predicted on the basis of experience, it should be possible to calculate the combination of widths of any number of dies which reduce the number of machine start-ups and keep the quantity of edge trim scrap to a minimum. It should therefore be possible to predict the best width of die for a new extruder if it is necessary to expand existing plant, or to select the optimum settings of blanking-off plates for a forthcoming scheduled period of production. This type of planning would be particularly profitable if the die lips did not need adjusting to alter sheet thickness or if they could be adjusted accurately and quickly without stopping the machine.

In order to show that methods are available for solving such problems, the relationship between edge trim scrap and sheet die widths was studied by the author and a colleague Mr. P. Chandler for a simplified case based on orders for TPS sheet received by BX Plastics Ltd. during 1960. In order to obtain the results in a reasonable period of time and overcome the tedious task of manually conducting many repetitive calculations it was necessary to programme on an electronic digital computer for part of this work and the Distillers Co. Ltd. kindly made available their Elliott 803 machine at Great Burgh.

The 1960 data was condensed to some 77 sheet sizes where,

Length varied between	16 - 83 in.
Width varied between	12 - 45 in.
Thickness varied between	0.040 - 0.240 in.
Annual tonnage of each size varied between	0.75 - 44.25 tons.

The best die width combinations which gave the least edge trim scrap were determined for one, two, three, four and six dies for two sets of product requirements.

(a) when the direction of extrusion was fixed relative to the longer dimension of the sheet sold: a graphical solution was used.

(b) when the sheet could be cut either longitudinally or transversely to the direction of extrusion, depending on which gave the least trim scrap; the computer was used.

The assumptions made, outlines of the methods used for solving both series of problems and the results obtained are all given in a BX Plastics Ltd. Confidential Report No. 629, issued on the 30th January, 1962. Other cases where edge trim scrap may become a serious problem also are outlined in the report.

The results show that the choice of die widths selected had considerable influence on the total annual trim scrap to be reprocessed and hence the first time yield. Although a number of simplifying assumptions were made in this work, it was considered that the methods could be extended so as to take into account the important factors which had been ignored. This might require considerable further effort, but it is felt that interesting possibilities for future study have been opened up by this novel approach.

8.9. References

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9. GENERAL EXTRUDER DESIGN AND OTHER COMPONENTS

9.1. Drive Motors

Ideally an extruder drive should

(a) maintain a constant set screw speed without drifting
(b) operate over a wide speed range from near zero to maximum
(c) be remote controlled
(d) have low initial cost
(e) require low maintenance cost.

The extruder drive H.P. is often a better indication of extruder production capacity than the screw diameter and in the past many extruders have been sold which were greatly underpowered. Although the modern trend is towards higher outputs, extruder manufacturers still differ widely in their views of drive motor H.P. Consequently extruders can be bought equipped with drive motors varying over a wide H.P. range as shown in table 49.

Diameter of Extruder in.	Drive Motor Range H.P.
2	6 - 20
3	10 - 50
4½	40 - 125
6	80 - 250

Table 49

The most popular H.P. ratings for these extruder sizes in the U.K. are approx. midway between the max. and min. values given in table 49. The drive motor determines the screw speed and the torque supplied to the screw. Occasionally the screw speeds have to be limited because of the danger of excessive work heat which causes the decomposition of certain plastic melts. The H.P. which can be supplied to a screw is also limited by the maximum torque which the screw will withstand before shearing off in the feed section, the value of this maximum torque is proportional to the cube of the root diameter of the screw. The torque limitation has been expressed in a technical bulletin (1) in terms of the approx. permissible H.P./100 r.p.m. of the screw, and the results are reproduced in table 50.

Screw diameter in.	1¾	2	2½	3½	4½	6	8	10
Approx. permissible drive H.P./100 r.p.m.	5	10	20	60	120	210	550	800

Table 50

To prevent serious damage to an extruder caused by excessive torque some safety overload device should be built into the drive mechanism, such as an electrical overload cut-out or a shear pin in the drive assembly.

The speed regulation of a motor is defined as the percentage drop in speed from no load to full load. A well regulated drive motor will compensate for a surging load and maintain a nearly constant screw speed. Load variations during steady running are usually less than 10% and therefore an extruder drive with a speed regulation of 10% will give an approx. total variation of output of 1% which is well within normal commercial limits.

The major types of drive motor system currently used on extrusion machines are

9.1.1. The mechanical drive: this consists of a standard A.C. motor driving a pulley and belt system. Pulleys of different diameters may be manually interchanged to give different drive ratios. This type of drive motor is usually used in conjunction with a P.I.V. gearbox or some other speed regulating device.

Power transmission efficiency is approx. 70%.

Mechanical drives are of simple and robust construction and they are often used with extruders requiring less than 40 H.P.

9.1.2. The Electro-Magnetic eddy current clutch drive: this unit is driven by a standard A.C. motor and the clutch drive operates on a magnetic slip principle which varies speed, depending on the field strength. These systems give a constant torque and therefore H.P. is proportional to screw r.p.m. At half speed only half of the input H.P. is available at the output shaft, the other half of the motor energy being dissipated as eddy currents in the magnetic coupling. These currents generate heat which must be removed by air or water cooling.

Efficiency is approx. 90% at top speed but at lower speeds falls in proportion to the speed.

Units up to 1,000 H.P. or more are available, smaller units of this type being widely used with medium and large size extruders. Their efficiency is very low at low screw speeds.

9.1.3. Variable voltage drives (A.C. - D.C. motors): this system (often known as Ward-Leonard sets) usually consists of a standard A.C. motor driving a D.C. generator which powers a D.C. drive motor. These motors are normally designed to give either a constant torque output over their whole speed range or a constant H.P. drive over the top half of their speed range.

Power transmission efficiency is approx. 75% at top speed but reduces at lower speeds.

Units up to 1,000 H.P. or more are available.

9.1.4. A.C. Commutator Motors: these consist of two types

(i) the motor speed is varied by rotating the commutator brushes.

(ii) the brushes are fixed and the motor speed is varied by a separate induction - voltage regulator which controls the voltage applied across the motor winding.

Power transmission efficiency of approx. 80% can be obtained at top speed but reduces at lower speeds.

These motors operate better at high speeds and at less than half the maximum speed they tend to overheat.

9.1.5. Hydraulic drive motors: these normally consist of a constant speed A.C. motor which drives a variable displacement oil pump, the oil being piped at high pressure via a non return valve to a hydraulic motor which turns the screw. The hydraulic pump is a swashplate unit and by altering the angle of the swashplate screw speeds varying between less than 5 r.p.m. and maximum can be obtained depending on the rate of oil flow.

These motors provide maximum torque over their whole working range and their power transmission efficiency is high at top speed.

9.1.6. General discussion

Speed regulation of all these motors is normally within 5%. The mechanical drive motor is the least versatile as far as screw speed is concerned, because of the limited ratio available from a P.I.V. gearbox, to widen the speed range the ratio of the drive pulleys is usually manually changed.

Mechanical drives are not usually operated by remote control but the other drives are, also the drives discussed in sections 9.1.2., 9.1.3. and 9.1.4. can be electrically synchronised with ancillary take-off equipment.

The variable voltage drives and hydraulic drives are the most expensive motors to buy and the mechanical drive motor is the cheapest, the latter is also one of the easiest to maintain.

Whether a constant torque or a constant H.P. drive motor is required depends largely on the type of material to be extruded.

The power requirement of a conventional single screw extruder is approx. 0.1 - 0.2 H.P. per lb/hr., it is difficult to be more specific than this on the question of extruder power requirements because it is dependent on many factors e.g. screw speed, output, screw design, material extruded, plastic melt temperature, extruder size, die design etc.

9.2. Gear Boxes

Because electric motors normally operate at high speeds (e.g. 1,000 - 2,500 rev/min.) it is necessary to use some form of speed reducer to obtain extruder screw speeds over a useable range. For this purpose gear boxes are designed to give speed reductions ranging from 6:1 to 30:1. One of the practical limitations in selecting the ratio of a gear box is the maximum torque which the screw will withstand, which in turn is dependent on screw design, the type of plastic material to be extruded and processing conditions.

The transmission of the drive power to the extruder screw in smaller extruders is often via belt or chain couplings, but in the larger extruders it is usually direct via the reduction gears, to overcome problems of belt slip etc.

The design of a gear box is very important and the gears in particular must be very robust and capable of transmitting the maximum drive power to the screw for lengthy periods. This may involve gear box oil filtration and extreme pressure lubrication problems. Speed reduction gears are either splash lubricated or pressure lubricated, and the gear box must therefore be fitted with good oil seals.

Some of the more commonly used gears are discussed below.

9.2.1. Spur gears: this type of gear has teeth which are cut square to the gear wheel itself, they are cheap to make and are often used in the cheaper extrusion machines. Spur gears suffer from the disadvantage of straight line contact between inter-locating teeth, each tooth in turn is therefore called upon to transmit the full motor loading. Consequently they possess poor wear and power transmission characteristics (90% or less) and often give rise to excessive noise.

9.2.2. Helical gears: all the teeth of this type of gear are cut slantwise and possess the same helix angle. These gears overcome some of the disadvantages of spur gears in that the full motor loading is passed on and shared by more than one tooth at a time. They possess better power transmission (94-97%) and wear characteristics but suffer from the disadvantage of developing an axial thrust on the gear shaft. To overcome the latter problem a second helical gear with teeth of opposite helix angle can be mounted on the same shaft, but this often results in shorter gear life because the loading is not always equally balanced on all gears. They are quieter running than spur gears, and although more expensive they are frequently used in extruders.

9.2.3. Herringbone gears: each herringbone gear wheel can be regarded as two helical gears with teeth of symmetrically opposed helix angles joined together. They are made to extremely close tolerances and overcome the disadvantages of the helical gear system. Although herringbone gears are more expensive than helical gears they possess better drive power transmission characteristics (approx. 98%) and are often used in the more expensive extrusion machines.

9.2.4. Worm gears: the power input shaft of this gear system consists of a screw worm, which drives a worm gear wheel with specially designed teeth, of constant helix angle. These worm gear systems are cheap and give a high ratio speed reduction in a relatively small gear box. However, they suffer from the disadvantages of producing large axial thrusts, particularly on the screw worm shaft, and possess poor wear and power transmission characteristics (80-90%).

9.2.5. Change gear systems: these systems are often advocated in cases where plastic materials of widely differing extrusion characteristics are used with the same extruder. For instance polythene will give a good quality product at low torque and high screw speeds whereas PVC usually requires high torque and relatively low screw speeds. The drive motor would have to be considerably overpowered to meet both these requirements unless full benefit of the drive H.P. is derived by altering the gear ratios to give the appropriate screw speeds.

9.3. Thrust Bearings

Thrust bearings in extruders are designed and positioned to absorb the axial rearward thrust of the screw. Radial thrust bearings keep the rotating screw concentric within the extruder barrel. To obtain a long thrust bearing life it is necessary to distribute the thrust load evenly over the greatest possible contact area. Thrust bearings must be well lubricated with oil and for this reason they are often pressure lubricated. For obvious reasons the thrust housings must have an efficient oil seal separating them from the screw.

The major types of thrust bearings used in extruders are listed below.

9.3.1. Ball bearings: these bearings are the cheapest to make but have two major disadvantages, they have a low contact area and require perfectly flat parallel seats which are not always easy to achieve, particularly on large extruders.

9.3.2. <u>Cylindrical roller bearings</u>: these have a higher contact area than ball bearings and are situated in slots in the bearing plate. To overcome the peripheral speed differential of one long roller, two or three shorter rollers are often used in each slot. With this system uniform loading of the rollers is extremely difficult to achieve and any slight misalignment causes severe excessive loading of individual rollers.

9.3.3. <u>Tapered roller bearings</u>: these overcome the peripheral speed differential problem but slight misalignment still reduces their working life appreciably.

9.3.4. <u>Spherical roller bearings</u>: these bearings have spherical curvature which makes them self aligning and overcomes the peripheral speed differential problem. Although more expensive than the bearings previously mentioned they are capable of absorbing greater thrusts and possess longer working lives.

9.3.5. <u>Michell bearings</u>: with this type of thrust bearing, the axial and radial thrusts are carried and absorbed by a number of thrust pads which are carefully designed and pivoted underneath so that they will float the load on their own oil film. The surface of these axial or radial thrust pads is either flat or curved respectively. The pivots are not normally centrally situated but are positioned towards the trailing edge of the thrust pads for reasons of maximum efficiency (i.e. minimum friction).

As the extruder screw shaft rotates in its oil bath the oil adhering to its surface is carried round and lifts all the pads at their leading edge, thus each of the pads generates a wedge shaped pressure-oil-film which carries the thrust. The thickness of this oil film is not dependent on the pressure exerted by an external oil pump but is dependent on the load, screw speed, and viscosity of the lubricating medium.

These bearings were developed for marine purposes but they are still used in some medium and large sized extruders, they are relatively cheap and possess long lifetimes if set up correctly.

9.3.6. <u>Multi-plate bearings</u>: these are often used in twin screw extruders (e.g. Moi & Windsor) where the largest diameter of the thrust race is limited by the distance between the screws. The axial thrust load of each screw is taken up by two sets of flat or conical plates which are staggered between the shafts of the two screws. These plates run in two series of correspondingly staggered multi-grooved seatings. The idea being to distribute the axial thrust over a large area, but usually only one plate bearing per shaft carries the thrust continuously because it is difficult to align these bearings accurately. The second bearing gets axial thrust only after the first has reached its load limit and therefore only gives protection from severe overloading. Forced oil lubrication and manufacturing costs are high.

9.4. Plastics Feed Section

It is essential to have a constant and regular feed of plastics material to the extruder otherwise the extrudate quality will vary considerably.

9.4.1. Feed hopper design: all sides of an extruder hopper should be set at steep angles to reduce the tendency of the plastic feedstock to bridge across the entrance port to the extruder. The entrance port is usually square, circular or rectangular.

It is important to maintain a constant head of material above the entrance port if at all possible, as the feed intake and hence the extruder output may vary with slight changes in the head pressure. To overcome such pressure changes a small ancillary constant head hopper is often situated between the main hopper and the extruder. Alternatively sensitive height measuring probes are often used which when co-ordinated with some form of feeding device (e.g. vibrating platform or feedscrew) control the feeding of free flowing granules to within fine limits. The rate of feed can be adjusted by varying the amplitude of the vibrating platform or the speed of the feedscrew and/or by opening or closing a shut off gate which is usually situated at the base of the main hopper.

9.4.2. Feed box design: as already mentioned the feed port is normally square, circular or rectangular in cross section and it is usually designed to expose 1 - 1.5 screw flights. The larger feed ports reduce the tendency for the plastic feedstock to bridge across the extruder hopper. The feed box design usually allows the plastic granules to flow vertically down on to the extruder screw but occasionally the feed box is cut away on one side to give either a tangential feed or undercut feed. The latter design is preferred for strip feeding of plastic materials but little quantitative data is available concerning the relative merits of each of these three basic designs when granular or powdered feedstocks are used.

The feedbox is often provided with a water cooled jacket to prevent conducted heat, from other parts of the extruder barrel, overheating the thrust bearings or the feed zone. Water cooling the feed box often gives more uniform and higher outputs because it prevents the materials from melting and sticking to the barrel in the feed zone, which would interfere with feeding.

9.4.3. Bridging of feedstock in the hopper: this can cause low or fluctuating outputs which considerably affect the product quality. Bridging of the feedstock in the hopper must be avoided and methods of achieving this have already been mentioned (see section 9.4.1.).

In the past few years this problem has received much more attention because of the tendency to improve compounding units or to avoid the intermediate compounding and granulating costs altogether by converting dry-blend and powder feedstocks directly into sheet or profile form.

Large rotating agitators in the hopper have been designed to overcome this bridging problem, also rotating propellers have been developed which fit into the feed box at the side of the screw to encourage better feeding of fine powders or dry blends (e.g. Werner Pfleiderer compounding unit).

Crammer - feeders such as ribbon, single screw or twin screw devices which deliver pre-compacted feed materials under pressure to the extruder screw have also proved successful with fine powder, bead, flake or plasticised feedstocks (e.g. Prodex & Barmag-Harting crammer feeders).

9.4.4. Other feed hopper equipment: preheating and drying equipment is sometimes used in the feed system but care must be taken to ensure that the feedstock temperature is uniform and controlled to within fine limits, otherwise fluctuating extruder outputs may result. Preheating the feedstock can give rise to higher extruder outputs at fixed screw speeds, but systems in which hot air is blown through the feedstock at the base of the main hopper are not recommended because sometimes air channelling effects are observed which give a non uniform temperature distribution.

Vacuum hopper systems have been shown to give such advantages as reduced porosity, better surface quality and improved physical properties of PVC dry blend extrudates (2). Such systems remove air, released gases and small amounts of moisture via the rear of the extruder screw, they also densify powdered and dry blend feedstock materials which may give better feed characteristics. However they do give rise to problems of constant flow and head of feed materials in the feed hopper, also these systems are known to cause surging troubles in the extruder if the volatile content of the feed material varies greatly.

9.5. Breaker Plate Design

9.5.1. Basic design: a breaker plate is located between the screw and the die, its purpose normally being to support screen packs. The latter are used in the extrusion process for a number of purposes e.g. adjust the back pressure to give improved mixing and/or quality, or to filter out dirt or gel particles.

Breaker plates usually consist of a large number of $\frac{1}{8}$ - $\frac{1}{4}$ in. diameter holes bored through a metal plate and counter sunk for streamlining purposes. Slotted hole breaker plates consisting of one, two or three sections are sometimes used because they are easier to clean. Ideally breaker plates should be carefully streamlined but this is not always necessary with heat stable polymers.

9.5.2. Replacement of breaker plates and screen packs

Breaker plates are usually designed to locate in a recess at the end of the extruder barrel and then the adapter section and die are attached. To replace contaminated screen packs thus involves stopping the extruder, dismantling the adapter and die sections and then refitting after changing the screen pack. This causes a serious and costly loss of production, in addition to a high scrap rate with each renewed start up.

Recently equipment has been developed which overcomes most of these difficulties.

9.5.2.1. Screen-Clean systems (e.g. Welding Engineers & N.R.M.)

These units enable screen packs to be cleaned in situ by a back flushing process, they also enable screen packs to be inspected and changed one at a time without disturbing the extrusion operation. Each unit contains two sets of screen packs and the plastic flow can be directed through either one, both, or a by pass. The latter is useful for purging purposes because start up material can be purged without passing plastic melt through the screen pack or the die.

These units are however both bulky and expensive and it is doubtful if they could be used with heat sensitive materials such as PVC because of material hold up and decomposition problems.

9.5.2.2. Quick change systems (e.g. Black-Clawson, Waldron-Hartig & Sterling)

These units only necessitate the slowing down of a sheet extrusion plant for a few seconds. All three units consist of a slide which carries two breaker plate and screen pack assemblies, and while one is in the operating position the other can be changed and cleaned. The slide can be hydraulically or pneumatically operated so that a clean breaker plate and screen pack locates in the working position and displaces the contaminated set at regular intervals of time or when required.

9.6. Valves

Part II section 6 of this work has shown the importance of controlling the back pressure in an extruder in order to achieve good mixing and/or overcome surging. This can be achieved by varying the resistance of the screen pack, but a better and more versatile method is to use a valve.

9.6.1. Static valve systems: these are positioned between the extruder and the die, they are very efficient and cheap but usually possess poor streamlining characteristics and give rise to dead spots in which heat sensitive materials such as PVC can stagnate and degrade. These are often referred to as plug valves and basically they consist of an adjustable plug or needle which can be manually operated to control the gap through which the plastic melt is able to flow. This type of valve system was used in the experimental work of Part II section 4 and it is well illustrated in a number of technical papers (3,4,5).

9.6.2. Dynamic valve systems: in this system the extruder screw acts as a working part of the valve. Although more expensive than the static valve system it avoids all dead spot and hold up areas in which heat sensitive materials would degrade.

The forerunner of this type of valve system was the "Hankey Pressure Head" which was developed by Monsanto in 1952 (6,7). There are two basic designs of dynamic valve systems.

(a) The end of the screw is tapered to form a cone which seats into a ring with a similar taper. Valving is accomplished by making an axial adjustment to either the valve seat or the extruder screw. The screw can be adjusted by either a mechanical or hydraulic mechanism which is located at the rear of the screw, or by the use of spacing rings e.g. Battenfeld extruders. Both these two possibilities are well illustrated in technical papers (5,6).

(b) A special nose piece is attached to the screw equipped with multiple tapered rings, these rings intermesh with similarly designed rings on the valve seat. Valving is accomplished by making an axial adjustment to the screw using the method described above. This system was developed by Black-Clawson and is well illustrated in technical papers (5,8).

9.6.3. Comparison of dynamic and static valve systems

As already mentioned the streamlining characteristics of a dynamic valve system are better than those of the static valve system but the former are much more expensive to make. With nearly all the valve systems mentioned above it is the last $\frac{1}{8}$ in. adjustment which produces any great change in pressure.

The dynamic valve system generally gives increased shear, higher plastic melt temperatures, better mixing and requires more drive H.P. than static valve systems operating at similar screw speeds or output rates.

9.7. Temperature Control

Heating an extruder barrel or die can be achieved by the use of electrical heaters, steam, or various fluids at controlled temperatures. Ideally temperature control equipment should operate by regulating both an efficient heating system and an efficient cooling system.

Steam is a good heating medium for jacketed barrels or cored dies and gives fast heat-up. However it is not always available and suffers from a maximum safety temperature and pressure limitation of approx. 205°C at 200 lb/in^2.

Circulation of hot fluids such as oil has the advantage of supplying an even heat without hot spots and temperatures approaching 300°C can be used if desired. Cooling is available when used in conjunction with a heat exchanger. However, despite its low operating pressure this system suffers from fluid leaks which give rise to high maintenance costs and cleanliness problems.

Electrical heating systems are generally preferred because of flexibility and cleanliness. For brevity future discussion will be confined mainly to these more popular heating systems.

9.7.1. Electrical heating systems

The object of heating an extruder barrel and sheet die is to lower the viscosity of the plastics material in the screw and produce a product of desired shape. It must be remembered that in most cases the heat supplied to 'melt' the plastic is mainly provided by the drive motor and not the barrel heaters, the final plastic melt temperature is therefore mainly dependent on screw geometry, screw speed and back pressure. Other factors affecting melt temperature are feedstock temperature and screw cooling. Autogeneous extrusion systems only require barrel heaters to initially heat up the extruder.

It is normally necessary to have a temperature gradient along the extruder barrel which rises from the feed section to the metering section of the screw. For this reason extruder barrels usually possess a number of separate heating zones. Large extruders with L:D ratios of 25:1 normally possess up to five such zones.

Large sheeting dies may also contain as many as six separate heating zones.

9.7.1.1. Barrel heating systems

The capacity of barrel heating systems varies considerably from one extruder manufacturer to another, depending on the type of heaters used, the plastic material to be extruded and the specified initial heat up period. The range of heater capacities used on the most popular sizes of 20:1 L:D ratio single screw extruders is summarised in table 51 below.

Diameter of extruder inches	Range of max. heater capacity K.W.
2	8 - 20
3½	15 - 50
4½	30 - 80
6	40 - 120

Table 51

Table 51 shows that there is little standardisation between extruder manufacturers on this subject of maximum heater capacity, although the most common values used are approx. midway in the above range.

Some of the major heating systems used on extruder barrels are summarised below:-

(a) Band resistance heaters are commonly used for heating extruder barrels because they are relatively cheap. They are in direct contact with the extruder barrel and are held at high temps. for extended periods of time, the insulated wire elements are at even higher operating temps. This together with the fact that they are not airtight and therefore susceptible to moisture condensation within the heater leading to arcing when initially switched on, account for frequent burn outs, and high maintenance costs.

(b) Tubular resistance heating elements are also used, these are frequently cast into aluminium blocks which are bored to fit the extruder barrel. Alternatively they can be machined to fit into circular grooves cut in the surface of the aluminium block. Segmented curved aluminium channels with small expansion gaps are sometimes used to eliminate air gaps and heat transfer problems, because aluminium expands twice as much as steel for the same temp. change.

(c) Induction heating systems were developed to give a faster response to temp. corrections and to decrease the "bandwidth" of temperature cycling, and are claimed by many to be more efficient than resistance heater systems.

The barrel is surrounded by copper induction coils although the two surfaces are separated by an air gap. Mains frequency alternating current flows through the induction coils and sets up an alternating magnetic field which generates eddy currents in the steel casing of the extruder barrel, thus causing a temperature rise. The air gap between the induction coils and the barrel surface facilitates faster air cooling than resistance heating systems, and a longer heating life is claimed.

However, they are much more costly than the previously mentioned systems and are subject to switching problems. Also the thermocouples in the barrel have to be well shielded from the alternating magnetic field otherwise they will give artificially high temperature readings.

(d) Pressurised liquid vapour systems surround the barrel with jackets which are connected via a condenser to a valve, each jacket contains a liquid and resistance heaters are situated on the outside of each of these units. When the barrel is heated the fluid acts as a temperature equaliser around the barrel. To extract excess work heat from the system the pressure is reduced and some of the liquid boils off and rises to the condenser which is air or water cooled, here it again becomes liquid losing the latent heat of vaporisation and drains back by gravity into the jacket. These systems are claimed to be very efficient and since the cooling takes place in the condenser which is situated above the barrel there is no thermal shock to the system.

Distilled water is used in the jackets for polythene and PVC but several other liquids may be used if a wider operating temp. range is required. This system therefore employs the advantages of steam and oil but overcomes their disadvantages.

9.7.1.2. Die heating systems

These are mainly confined to strip and cartridge resistance heaters, the former are bolted to the surface of the die and the latter fit into cylindrical holes. It is often difficult to achieve a uniform temp. across the width of a large sheet die with these heaters, because both systems normally incorporate air gaps which partially insulate the heater from the die. Also parts in contact with the die are liable to form hot spots. Fluid heating systems are seldom used because they require bulky heat-exchange equipment and give rise to fluid leakage problems. However, if well designed they have the advantage of giving a uniform die temperature.

9.7.2. Heater control

There should be at least one temperature transducer for each zone of the barrel or die which is separately heated. It is particularly important to maintain a sheet die at a constant temperature over the whole of its length and width for the reasons stated in part II section 8. Whatever type of heating system is used the accuracy of control is dependent to a large extent on the sensitivity and position of the temp. transducers.

Ideally it is the temperature of the internal barrel and die surfaces in contact with the flowing plastic which need to be controlled, but all too often the wrong temperature is measured. It should be remembered that deeply set thermocouples which are near to these internal surfaces do not necessarily record and control these surface temperatures accurately, because the thickness of the steel between the thermocouples and the heater and cooling system causes a response lag which may lead to overshooting and a bad 'hunting' situation. On the other hand if the thermocouples are placed too near the heaters they merely control the heater temps. and this may mean that the recorded temps. are not directly related to the plastic melt temp. in a particular zone. To overcome this problem West Instrument Ltd. developed a dual thermocouple recording system which averages out the difference in temp. between the internal and external surfaces of the barrel and minimizes temperature 'hunting'. The deeply set thermocouple may be regarded as establishing the temperature control point on the control instrument and the shallow one establishes the on-off control required for operation.

Some of the more commonly used systems for controlling barrel and die resistance heaters are listed below.

9.7.2.1. Manual controller

This system uses manually adjusted rheostats or variable ratio transformers.

9.7.2.2. On-off controller

A certain amount of 'hunting' invariably results with this system because the heaters are switched on or off as the indicated temperature sensed by a thermocouple falls below or rises above the set value.

9.7.2.3. Sensing on-off controllers

This is an improved form of on-off controller which senses the amount of time during which the heater should be switched on as the recorded temp. nears the set temp. This system reduces the difference between the maximum and minimum temp. of the barrel or die zones.

9.7.2.4. Stepless or proportioning controllers

These systems can greatly minimise hunting phenomenon by controlling the heat input smoothly in a manner related to the process requirements. The heat lost by such a system may be closely matched by the heat input, the power input is in direct proportion to the temp. deviation at any given time and as the required temp. is approached the power correction is automatically reduced.

Stepless controllers should be used if possible for die temperature control and sensing on-off controllers are usually adequate for barrel temperature control. The former suffer the serious disadvantage of being very expensive.

9.7.3. Cooling systems

Cooling is usually necessary to remove excessive conducted or frictional work heat from some part of the extruder system, and is seldom if ever used in connection with sheeting dies.

9.7.3.1. Barrel cooling

The incorporation of cooling systems on the barrel normally insulates parts of the heating system from the barrel and makes the latter less efficient. Some commonly used barrel cooling systems are listed below.

(a) Pressurised liquid vapour heating/cooling systems have already been discussed in detail (see section 9.7.1.1. (d)).

(b) Flattened copper cooling coils which may locate in helical grooves on the barrel surface and enable heaters to be fitted over them are sometimes used. A cooling medium such as air, water, steam or oil may be circulated through them depending upon the heat removal requirements. The circulating of warm liquids avoids the thermal shocks experienced by the use of cold water. Cooling coils are sometimes incorporated into cast-in aluminium heaters.

(c) Cool air is blown through grooves or gaps between the heaters and barrel surfaces e.g. induction heating system.

In general air cooling systems remove heat more slowly and are less effective than water cooling systems but the latter suffer from the disadvantage of scale being deposited in the cooling channel and they require regular cleaning or replacement unless specially softened water is used. Cooling systems which are separated from the barrel surface by the heating system are much less efficient than those mentioned above because the heaters must first be cooled. Cast aluminium fins are sometimes used to help dissipate heat.

9.7.3.2. Screw and feed-box cooling

The effect of passing cooling water through a cored screw or jacketed feed box has been discussed at length earlier (see part II sections 7.1.6. and 5.4.2.). However, it is worth remembering here that cooling the screw reduces the effective depth of the screw channel in the melt section which results in increased shear of the plastic melt and lower outputs at low back pressures, this in turn may give rise to higher, rather than lower plastic melt temps.

9.8. Extruder Instrumentation

The main purpose for instrumenting an extruder is to assess what is happening inside the extruder and die, and to measure the extrudate profile accurately. Such information greatly helps the extruder operator to maintain close control over the dimensions, quality and properties of the extrudate.

Many extruders in use in the plastics industry today are under instrumented, the extra initial capital cost of a well instrumented extruder should be quickly repayed by the increase in the efficiency of the process (e.g. by decreasing the start up time and the total percentage of scrap produced). Good recording and controlling instruments may cost as much as the basic extruder. There are many systems available for amplifying and recording electrical signals from all types of transducers, but the instruments which give greater accuracy and quicker feed back of information are usually more expensive to buy and maintain.

Instruments which are frequently used to measure, control and record the more common extruder processing variables are listed below.

9.8.1. Barrel and die temperatures

Methods for controlling the barrel and die temperatures have already been discussed (see sections 9.7.1. and 9.7.2.).

9.8 2. Plastic melt temperature

This should be measured at a point between the extruder and the die, so that the die may be maintained at or near to the temp. of the entering melt. Unshielded melt thermocouples have now been developed (9) which are much more sensitive than conventional shielded ones by virtue of their direct contact with the melt, and which indicate temperature variations at least three times as large as those indicated by the latter.

Radiation pyrometers can be used to measure the extrudate temp. as it leaves the die but these are much more expensive than melt thermocouples and they are generally not as accurate or useful.

9.8.3. Plastic melt pressure

If the extruder is equipped with a valve, and only one pressure transducer is used this should be situated on the extruder side of the valve. If a second pressure transducer is available this should be situated at some point in the main die channel. If a valve is not used the pressure transducer should be positioned on the extruder side of the breaker plate.

Bourdon-gauges were and still are, popular as pressure transducers but since 1959 advances in the design, reliability and accuracy of strain gauge pressure transducers (9,10) make them well worth the extra initial cost because they overcome many of the disadvantages of the former such as the use of silicone grease which may contaminate the plastic melt and inefficiency due to plastics material clogging the instrument.

9.8.4. Height measuring probes

These have been mentioned briefly in section 9.4.1. and employ the principles of electrical capacity to detect the approach to, or recession from, the probe of any free flowing granules in an extruder hopper. Thus changes in capacitance can be made to actuate a relay which will stop or start a feeding mechanism to the extruder hopper and thus maintain the pre-determined level of feedstock material in the hopper.

9.8.5. Drive power input

A torque measuring device fitted to the screw drive makes it possible to calculate accurately the total power delivered to the screw without taking into account motor and drive efficiencies. Such a torque measuring device is well illustrated and described in a paper by Kennaway (11).

Failing this, a wattmeter or ammeter in the motor circuit gives a continuous indication of the motor load, so that excessive loads can be avoided and correct operating conditions can be established. Wattmeters are more expensive and more accurate than ammeters for indicating the motor load, but either are useful in solving feed problems.

9.8.6. Integrated heat power input

Wattmeters can also be used to give a continuous reading of the total heat input at any given time if this information is considered necessary.

9.8.7. Screw speed

Some form of tachometer should be attached to the screw drive to continuously indicate and/or record screw speed.

9.8.8. Extrudate dimensions

The thickness of extruded plastics sheet is usually measured with a manually operated micrometer screw gauge.

Non contacting thickness control systems have been developed in recent years and their application will no doubt continue to expand. Many papers have been published on the subject of beta ray thickness measuring devices (12,13,14). These units can be made to traverse across the sheet width and can give continuous data concerning thickness variation. The radio active source of these detectors can be varied to give accurate thickness measurements of extruded film approx. 0.00025 in. thick and extruded sheet up to 0.150 in. thick. Stronger radioactive sources are now available for measuring even thicker gauge plastic sheet, although the safety aspects of using such sources must be watched to prevent harmful irradiation of personnel.

Electrical capacitance methods are also being developed to measure sheet thickness variations and they have already proved successful in measuring accurately the wall thickness of plastic tubes.

Pneumatic gauging devices for measuring differences in sheet thickness have also been developed but they do not appear to be used commercially at the present time.

9.8.9. Discussion on instrumentation

9.8.9.1. Accuracy

All instruments should be regularly checked for accuracy and recalibrated if necessary. It is now possible to obtain temperature and pressure transducers and rapid response recorders which give readings accurate to within 1°F and 1 lb/in.^{2} respectively in less than a second. Such accuracy is not necessary on extrusion equipment, and readings which are accurate to within $\pm 2^{\circ}$C or $\pm$ 10 lb/in.^{2} are adequate.

9.8.9.2. Automation

It would be a great advantage to the extruder operator if all extruder variables were recorded and controlled at a central control point but this would prove very expensive and is not essential. The most important variables which a sheet extrusion operator must watch carefully, apart from take-off and ancillary equipment, are die and barrel temperatures, back pressure, plastic melt temp., output, screw speed, drive power and sheet thickness. The extruder operator should also be able to identify and rectify certain surface defects. The responsibilities of a sheet extrusion operator are therefore considerable and over the years his task has become more difficult as the quality of the sheet has been improved and standards have tightened to the present day high level.

There is little automation in a sheet extrusion line to help the operator after the initial sheet extruder variables have been manually preset, only the barrel and die temperatures and the sensitive feed probe in the hopper are self adjusting. Recently thickness measuring equipment for extruded tubes and pipes has been successfully co-ordinated with the take-off equipment, so that the latter will automatically speed up or slow down to give pipe of the correct dimensions. A recent paper by Sponaugle (15) shows how beta gauge equipment can be used to give automatic profile gauge control of film and extrusion coatings.

The present obstacle towards increased automation on a sheet extrusion line is its very high initial capital cost, and many technical difficulties have yet to be overcome. However the idea of a sheet extrusion line controlled by an analogue computer may well become fact at some time in the future.

9.9. Maintenance

9.9.1. Object of maintenance

Maintenance of extrusion equipment may be split into two categories 'preventive' and 'breakdown'. The purpose of a maintenance department is to assist production personnel to keep production machines in good running order so that they suffer the minimum inconvenience from breakdowns and lost production efficiency. Also, they must carry out modifications to existing plant, services and buildings.

9.9.2. Company maintenance policy

Small companies which use only a few extrusion machines and instruments may not be able to support a large maintenance staff and therefore often rely almost entirely on breakdown maintenance; they continue until breakdown occurs and then attend to it or call in outside help. This system when applied to extrusion machinery is not very satisfactory because if only the actual component causing breakdown is repaired other components which have not been checked have been known on many occasions to cause further breakdown. Such repetitive breakdowns not only considerably reduce the production efficiency of an extrusion plant because of the downtime and loss of production, but they may also greatly inconvenience the customer if his orders are not met on time.

Larger companies may find it economically necessary to organise preventive maintenance where vulnerable components are inspected and/or replaced at regular intervals regardless of whether they show any sign of trouble, and they often support large maintenance units. The latter may be subdivided into mechanical, electrical and instrument maintenance units. For these separate units to work efficiently they must be well co-ordinated otherwise valuable production time will be lost.

The immediate repair service which a maintenance unit is capable of giving depends to a large extent on the store of replacement parts which the company holds. Much less capital is tied up in a maintenance store which provides for a production department containing similar extrusion machines in comparison to one with many dissimilar machines. It is therefore important, whenever possible to standardise on extrusion equipment, particularly instrumentation where spares are often expensive.

Breakdown maintenance is seldom predictable and therefore companies working twenty four hours a day should provide stand-by maintenance staff at all times.

9.9.3. Extruder maintenance

9.9.3.1. Drive, bearings and gear box

These working parts should always be protected against severe overloading by an electrical overload cut-out system or a shear pin in the drive assembly. The correct grades of oils and greases should be used for lubrication purposes and lubrication systems should be regularly checked to ensure that there are no oil leaks due to faulty seals, gaskets or clogged breather units. If these working parts are continuously operated 24 hours/day they should be given an oil change and closely inspected for wear approx. 2-4 times per year. Chain or belt drives should also be regularly checked to make sure they are not excessively tight or loose.

9.9.3.2. Screw and barrel

The clearance between the screw flights and the barrel is an important factor in production efficiency and excessive wear results in a considerable decrease of extruder output for any given screw speed. Maddocks (16) has shown that for efficient running a screw should be repaired or replaced when the flight land clearance between the screw and the barrel surfaces has increased to approx. 15% of the metering channel depth. Leakage flow is considered negligible if the flight land clearance does not exceed 5% of the metering channel depth, this value is recommended by Maddocks as the maximum for original screw design. When it is necessary to replace a screw or build-up the screw flights the barrel should also be inspected and measured, if excessive wear has taken place the barrel or barrel liner should be replaced.

The useful lifetime of extruder screws and barrels is usually lengthened if care is taken to ensure that no foreign matter such as thin slivers of metal are allowed to pass into the extruder with the feedstock material. Such metal components are often present in rechip material and are normally removed by magnets as the feedstock is metered into the extruder hopper.

9.9.3.3. Die, adapter, valve etc.

Care must be taken at all times to ensure that no surface of these parts which comes into contact with the plastic melt is knocked, scratched or damaged in any way. When not in use these parts should be carefully greased and stored. All nuts, bolts, and threaded parts of these sections should be coated with a release compound before use, to prevent parts seizing due to heat.

9.9.3.4. Cleaning methods.

If heat-unstable plastic materials are being processed, the die and screw should be removed and cleaned immediately after shut-down, and the barrel must also be cleaned as quickly as possible.

Steel tools should never be used to clean the extruder barrel, screw or die parts as they are too hard and can easily damage the surfaces. Wood, or brass levers can be used to remove screw and die parts. The die and screw are normally cleaned with brass scrapers and/or wire brushes and the barrel is normally cleaned with copper or brass mesh pads.

Occasionally plastic materials are removed by burning off techniques but this should be done by experienced personnel otherwise die parts etc. may become warped or annealed. It is usual to clean nuts and bolts etc. in a solvent for the plastics material.

9.9.4. Instrument maintenance

All heating, cooling and screw temp. control systems must be readily accessible for periodic inspection, repairs and replacement. Band heaters should be periodically tightened to ensure a tight fit, as any dead spots insulated from the barrel by air will decrease the life of the heater. Also thermocouple seatings and connections in all parts of the extruder 'set-up' should be checked regularly.

All recording and control instruments should be protected from unnecessary vibration and for this reason it is advisable to mount such instruments in a control panel isolated from the extruder.

All instruments should be checked and recalibrated as often as is practicable; on no account must they be knocked if they are not working correctly because this is likely to cause further damage.

9.9.5. Major causes of maintenance problems

Care must be taken at the outset to ensure that all extrusion equipment is installed correctly as this can lead to serious maintenance troubles. Cold starts, the use of the wrong grade of oil or grease and/or incomplete lubrication, and careless or accidental damage to the die or screw are some of the major causes of many maintenance problems. Also many extruder operators do not take the trouble to ensure that whenever possible the extruder is clean and free from water, oil and plastics leaks.

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10. COLOURING SHEET MATERIALS BY THE EXTRUSION PROCESS

10.1. Introduction

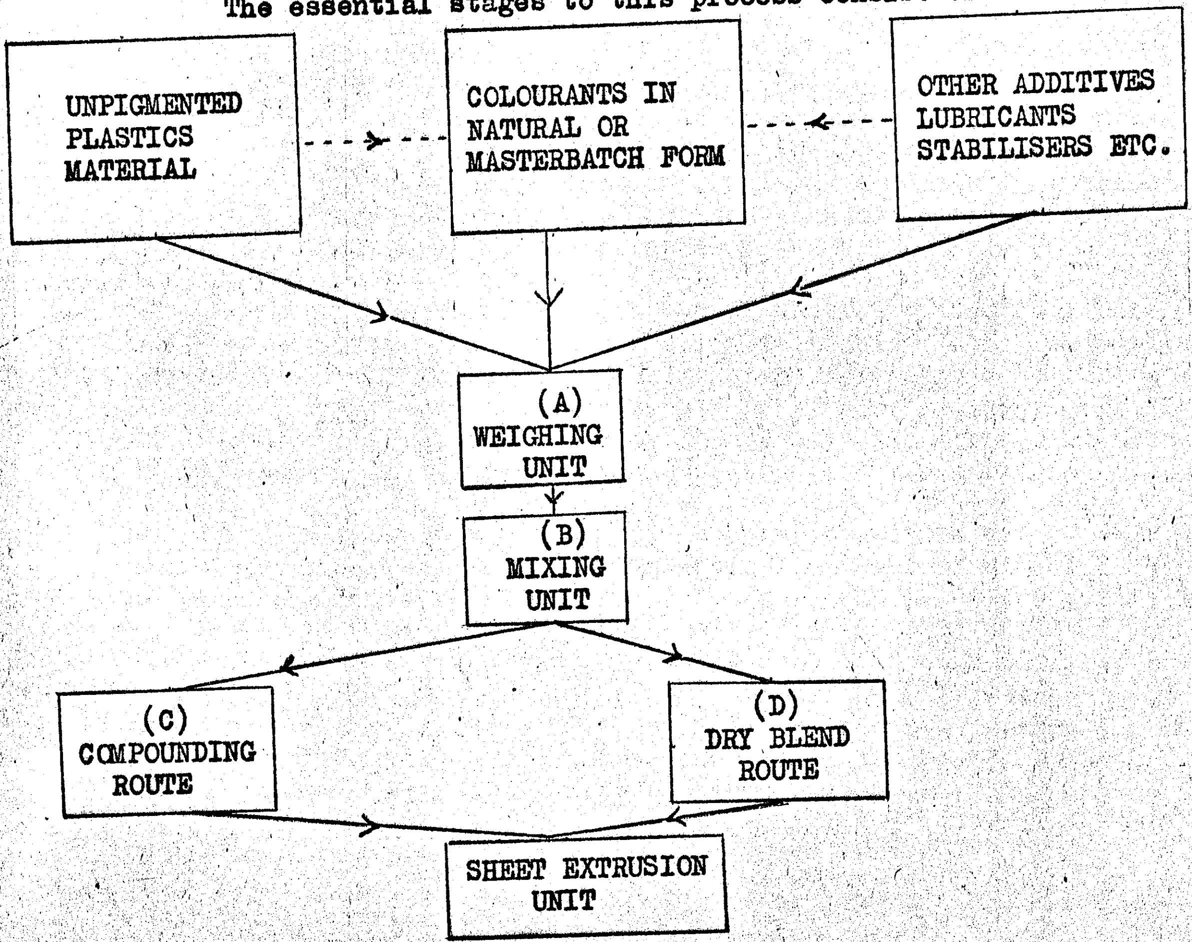
Colouring by the extrusion process is primarily a method by which pigments and/or dyes are uniformly mixed and dispersed in a plastics material as they pass through an extruder. A secondary function of this process is to uniformly mix and disperse other important additives such as lubricants, stabilizers, fillers, plasticisers, wetting agents and antioxidants etc.

The term dispersion is used here in connection with the extent to which small agglomerates of different components have been reduced in size and separated by a shearing action.

The term mixing is used here in connection with the statistical distribution in space of two or more different components without affecting their degree of dispersion.

10.2. The Extrusion of Coloured Sheet Materials

The essential stages to this process consist of:-



A. The weighing stage consists of either

(1) Weighing the polymer, colourants and other additives in their natural form.

or (2) Weighing the colourants in masterbatch form and the polymer and other additives in their natural form.

A masterbatch consists of a uniform mix of polymer and pigment and normally contains approx. 10 - 50% by weight of pigment, and it may also incorporate all the other additives in the correct proportions. Usually up to 10% by weight of the masterbatch is mixed with the natural (unpigmented) plastics material to produce the final colour blend. Hot compounding equipment such as banbury mixers, rolls, compounding extruders etc., is often used to produce uniformly mixed and dispersed masterbatch materials in compound form, but masterbatches may also consist merely of an intimate powder blend.

The initial weighing and mixing are usually batchwise processes but continuous weighing machines are available and sections 10.3 and 10.4 outline experiments which were conducted to make these two processes fully continuous. It is essential that the initial weighing process is fully reliable and accurate to within ± 2% if normal production colour tolerance standards are to be maintained. This applies for both continuous or non continuous and constant weight or constant volume metering techniques.

The colourants used in the production of thermoplastics sheet by the extrusion process should be capable of being readily dispersed without forming agglomerates and they should possess good heat and light stability properties. They should not promote oxidation or degradation or react in any way with the plastics material under normal processing conditions, or on ageing.

B. The mixing stage consists of producing a uniform mix of all the preweighed ingredients. The theoretical concept of mixing is by no means clearly understood, and it is not easy to define quantitively the degree of mixing or dispersion obtained in solid/solid or solid/liquid systems. Mixing efficiency is dependant on such factors as processing techniques, mixing time to reach equilibrium, size and structure of the ingredients, local viscosity, rate of shear, specific gravity, electric charge induced by the mixing action, surface tension and wettability of liquids etc.

Because of the complexity of the problem, arbitrary standards of good and bad mixing and dispersion are usually set by the compound or sheet manufacturing company in terms of visual appearance of product. For this reason the theory of mixing and statistical description of equilibrium or random mixes of two or more components will not be discussed in detail, although this problem has been studied by many workers, (1.2.3.4.5.6.7.8.9.)

However, in general it can be stated that as the ratio of concentrates of the components of a perfect mix of two or more solid constituents increases, the chance of two samples from the same mix taken at different points being identical is greatly reduced. Also, the chance of obtaining a representative sample increases as the particle size of the constituents is reduced, but decreases as the size of the sample taken is reduced.

Mixing is normally achieved by using revolving tumblers or drums of various shapes and designs, ribbon mixers, conical blenders, air fluidising mixers or mechanical high speed mixers etc. The latter (e.g. Henschel and Papenmeier) have the advantage of raising the temp. of the mix appreciably because of the work heat evolved. However, these high speed mixers must be carefully controlled to ensure that the plastics material is not fluxed or gelled and if discharged directly into the extruder hopper all batches must be at the same mix temperature, otherwise extruder surging and non-uniformity of product may result.

(C) The compounding route consists of passing the premixed ingredients through a compounding extruder, the latter further gels, mixes and disperses the ingredients and face cutting, pelleting or dicing equipment is used to produce granules of standard size. Torpedoes are frequently used in compounding and/or sheet extruders to encourage mixing and dispersion of the separate ingredients, and the effect torpedo design can have on the degree of mixing of the constituents has been discussed in PART II section 6. The most efficient types of compounding extruders have been discussed in detail in section 7.7.

The compound so produced is re-extruded into the required sheet form using the sheet dies and take-off equipment described in sections 8 and 11 respectively.

(D) The dry blend route consists of feeding the pre-weighed and pre-mixed ingredients direct to the sheet extruder, thus omitting the compounding stage.

This means that the sheet extruder must supply the extra mixing and dispersing action normally performed by the compounding extruder, and yet be free of surging so that good quality sheet may be made in one pass through the extruder. Therefore, dry blend extrusion normally requires longer extruders, higher compression ratio screws with shallower metering sections, and different extrusion conditions e.g. barrel temps., and screw speed etc. Increased back pressures are usually necessary in order to obtain equivalent melt temperatures and quality of extruded product.

The major advantages of dry blend extrusion are lower conversion and capital costs because the intermediate compounding process is eliminated, and a reduced heat history which may produce a better quality product with heat unstable materials such as PVC. The major disadvantage of this process is that the degree of mixing and dispersion is not usually as good as that achieved via the compounding route. PVC, CA, and TPS sheet materials are all produced via both the compounding and dry blend processes, the former method usually produces a better quality sheet, with heat stable materials, but often neither of these processes give the standard of colour mixing or dispersion normally achieved by internal mixers and heated rolls.

Problems such as formulation changes, blending and mixing methods and the effect of vacuum hopper, screw design and vented extruder systems on the product quality of dry blend extrusions have all been studied by other workers (10.11. 12. 13. 14. 15.) and therefore will not be discussed here.

10.3. Fully Continuous Weighing, Mixing and Extrusion Colouring Techniques

As previously stated the extrusion process is a fully continuous process, but the initial weighing and mixing processes currently used in the plastics industry are essentially batchwise processes. The problem of making the initial weighing and mixing processes fully continuous was investigated and experiments were conducted with recently developed constant weighing and metering equipment with this object in mind. It was thought that such a continuous process would be labour saving, reduce the risk of human error and ensure a high standard of quality.

10.3.1 Testing continuous weighing and/or metering machines

The following machines were investigated and represent a good cross section of the constant weighing and metering machines which are capable of handling plastics materials in granule and/or powder form and which are currently available to the plastics industry.

None of the machines investigated were capable of handling dry pigments in powder form or dry blend PVC masterbatches containing a high proportion of plasticiser.

These continuous metering machines fall naturally into two categories, namely constant weight displacement machines and constant volume displacement machines (see table 52). It is essential that the accuracy of a constant weighing and metering device is fully reliable during extended production runs otherwise large quantities of off-colour scrap will be produced.

TYPE		MODEL	MANUFACTURING CO.
Const. Wt. Displacement	(a)	Select-o-Weigh	Richardson Scale Co. Ltd
	(b)	Englehardt Feeder	Industrial Plastics Ltd.
	(c)	Constant Weight Feeder	Sinex Engineering Co.Ltd.
	(d)	Pneumatic Merchen Feeder	Wallace & Tiernan Ltd.
Const. Vol. Displacement	(e)	Color-o-Meter	Wilson Products Ltd.
	(f)	Vibra Screw Feeder	Simon Handling Engineering Ltd.

Table 52

Each of these machines operates by a different technique, briefly these are as follows:-

(a) Programmed additive weight dialling technique

(b) Synchronised small batch weighing and tipping devices

(c) Vibratory feeder synchronised to a continuous weigh belt suspended on one arm of a high precision balance

(d) Continuous flexible belt system with a pneumatically controlled mechanism for sensing the weight of material on a fixed length of belt

(e) Constant volumetric discharge from rotating platforms

(f) Constant volumetric discharge from rotating/vibrating feed screw.

Each of the above machines was investigated from the point of view of size, accuracy, mechanical and electrical reliability, ease of cleaning and versatility. Four of these machines were assembled in line with $3\frac{3}{4}$ in. production extruders in the BXP Ltd. Compounding Dept. and extended colour dispersion, mixing, and colour toning trials were conducted.

In some cases more than one size of the same model was investigated.

The object of these trials was to meter polystyrene masterbatch materials accurately over a range of 5 - 35 lbs/hr and natural rough cut polystyrene (BXP grade 234) over a range of 150 - 350 lbs/hr. Masterbatch materials contained approx. 10% by weight of pigment.

Only a brief summary of the results of this work are given here: the full results can be found in a series of BXP Internal Reports (16).

10.3.2 Summary of results

The mechanical and/or electrical reliability of most of the machines tried was suspect but if working correctly all but one of the systems investigated were capable of continuously supplying polystyrene masterbatch and/or rough cut natural polystyrene at the required feed rates. Further, provided a baffled funnel was used between the weighing and/or metering equipment and the main extruder hopper, production colouring and toning trials showed that the colour and quality of mixing obtained during extended production runs was comparable to the results obtained by the standard batch weighing and tumbling techniques.

The models which operate on a constant volume displacement principle suffer from the disadvantage of having to be recalibrated with each feed material, because the output from these models depended on the bulk density and feed characteristics of the feed material. However, the most versatile system was the rotating/vibrating screw model because it could continuously feed a wider range of materials accurately, these included powdered feedstock materials e.g. powdered PVC (grade Geon 112) and powdered ABS (grade BXP Abstrene). None of these systems was capable of metering dry finely powdered pigments or plasticised PVC cold blends accurately.

Apart from the production of masterbatch materials the polystyrene extrusion compounding and dry blend sheet extrusion processes could be made fully continuous. Further, these experiments indicated that a rotating/vibrating screw metering machine was the best method for feeding bulk natural rough cut polystyrene materials and a constant weigh belt feeder the best method for feeding masterbatch materials.

The cost of these fully continuous metering units varies considerably depending on the model, but generally the constant volume displacement machines appear to be cheaper and more robust.

The cost of replacing conventional batch weighing and tumbling processes with fully continuous plant is dependant upon the existing plant layout (e.g. position, size and number of extruders etc.), the size of customers' orders, and the range of coloured masterbatches and compound used to produce the required sheet. Fully continuous weighing and metering equipment is ideally suited to large compounding plants which produce large orders over a small colour range.

On the other hand installation of fully continuous weighing and metering equipment in a company producing a large number of small orders over a wide colour range may lead to high initial capital cost, high maintenance costs and increased breakdown and off colour scrap problems compared to conventional batch weighing and tumbling techniques, without giving any increase in extruder output or quality. Also it may involve a complicated chute and/or belt network feed system without making any appreciable saving in manpower because the metering equipment may require constant cleaning, resetting and recalibrating.

Companies producing plastics compound and/or sheet are often faced with the problem of matching non standard colours. Large orders are economical but small orders are often unprofitable because of the high cost of colour matching, and the relatively high scrap rate and cleaning costs involved with the extrusion of small orders. Manufacturers usually rely upon their standard range of colours and/or large orders for profit and undertake production of small quantities in the hope that more profitable business will follow.

10.4. Continuous Weighing and/or Metering of Dry Pigments

As stated previously none of the above machines were found to be suitable for metering dry pigments in powder form. In an attempt to eliminate the masterbatch compounding process, trials were conducted with an Alexanderwerk Pigment Compounder. This machine consists of two hollow cylinders rotating in mill fashion, one cylinder having holes drilled through it, while the other cylinder has a serrated surface and rotates at an approx. 20% faster rate. The shear between the two cylinders compresses the pigment, forcing it through the drilled roll whence it is broken into pellets by a scraper blade mounted on the inside.

The aim of these experiments was to obtain uniform pellets of pigment and lubricant etc., which would keep their shape while being fed to a compound or sheet extruder by one of the methods indicated in section 10.3.1 and yet would break down in the extruder to give uniform mixing and pigment dispersion.

Only a brief summary of these results will be given here, although full details of the work conducted can be found in a BXP Ltd. Internal Report (17).

10.4.1 Summary of results

Early trials with dry pigments indicated that a binding medium would be required to obtain satisfactory pellets. Several binding media were tried and partial success was achieved with a few selected pigments. The results of this work showed that it was very difficult to obtain more than an 80% conversion of powder into pellets with one pass through the machine. Also in the few small scale extrusion trials conducted the pigment dispersion and mixing efficiency of these pellets was not as good as that obtained via the dry blend tumbled mix, or the masterbatch route. Much more development work is necessary before this method of extrusion colouring could be used commercially.

10.5. Conclusions

Methods are available for making the initial weighing and mixing techniques of the extrusion colouring process fully continuous. However, these methods are costly and are not worthwhile unless the colour range, size of orders and plant layout is such that a few weighing and/or metering devices require very little attention to produce large quantities of product to within the required commercial colour mixing and dispersion tolerances. Although these systems reduce the human error factor they increase the risk of mechanical breakdown or faulty operation which causes the production of off-colour scrap.

These continuous metering machines can be used to feed masterbatch and bulk natural materials in pellet form direct to a sheet extruder and thus eliminate the intermediate compounding costs, provided there is no significant lowering of sheet quality.

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11. THE DESIGN OF ANCILLARY SHEET TAKE-OFF EQUIPMENT

11.1. Introduction

The functions of extruded sheet take-off equipment are to receive the semi-molten stream of plastics materials from the die, support it fully while still hot and flexible, and cool and convey it in a controlled manner so as to deliver a cool solid flat sheet which may be cut into panels or wound into rolls as required. The degree of flatness required depends on the subsequent treatment on application: for example sheet to be rolled up need not be free from longitudinal curl, and sheet for corrugating (involving re-heating) need not be perfectly flat.

In addition, the take-off equipment is usually required to impart a better surface finish than would be available direct from the die, to assist in getting a uniform thickness sheet, to trim to the correct width, and to accommodate other ancillary processes such as continuous lamination of foil to the sheet surface.

The ancillary sheet take-off equipment used for the production of TPS, CA and rigid and flexible PVC sheet is usually similar in all cases. However, it can vary depending on the purpose and required physical properties of the product and for this reason the ancillary sheet take-off equipment used for each of the plastic sheet materials mentioned above will be discussed separately.

11.2. Toughened Polystyrene sheet take-off equipment

Three types of TPS sheet are normally produced for refrigerator, packaging and display market outlets. These are:-

11.2.1. Normal extruded surface sheet.

11.2.2. Foil coated sheet.

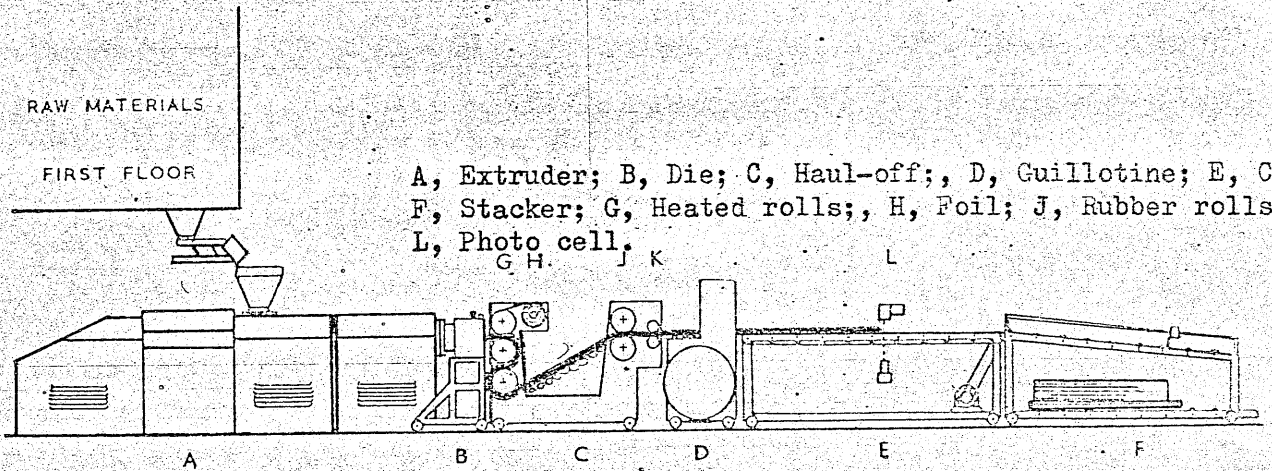
11.2.3. Heat polished sheet.

11.2.1. Normal extruded surface sheet

11.2.1.1. The extrusion of thick sheet

The ancillary take-off equipment used for the production of thick toughened polystyrene sheet (i.e. between 0.040 - 0.350 inches thick) normally consists of (a) a calender roll stack unit
(b) a roll conveyor unit

Diagram Of Typical Extrusion Plant Used For The Extrusion Of TPS Sheet



A, Extruder; B, Die; C, Haul-off;, D, Guillotine; E, Conveyor table; F, Stacker; G, Heated rolls;, H, Foil; J, Rubber rolls; K, Trim wheels; L, Photo cell.

Fig. 43

(c) a roll pull-off unit
(d) edge trim units
(e) a sheet guillotine unit
(f) a sheet delivery and automatic stacking unit.

A typical TPS thick sheet production line is shown in fig. 43 and a detailed description of each of the six units (a) - (f) mentioned above is given below.

(a) <u>The calender roll stack unit</u>

The sheet is extruded through a slot die and then picked up by either the middle or bottom roll of a vertical three roll calender stack. If the sheet is threaded through the top nip, round the middle roll, through the bottom nip and under the bottom roll it is said to move 'down the stack'. If on the other hand it is threaded through the bottom nip, round the middle roll, through the top nip and over the top roll it is said to move 'up the stack'. Both techniques are claimed to have marginal advantages when the sheet is used for different purposes but the most widely used technique is 'down the stack'. Some manufacturers prefer to use vertical four roll calender stacks which by slight height adjustment enable them to use either technique.

All the rolls of the calender stack usually have the same diameter, although some units are specially built with one or two of the rolls of larger diameter so that a better surface finish may be imparted to the extruded sheet. The most popular roll diameters presently used on the larger sheet take-off calender stacks are 6, 8, 10 and 12 ins, the larger diameter rolls being also much more expensive to buy. All calender rolls should be highly polished and are usually chrome plated to give a durable mirror like finish. They are cored and baffled or of a double shell design to permit uniform heating or cooling by high pressure hot water. The surface temperature across the rolls should be uniform and controlled to within ± 2°C, using a separate closed circuit heat exchange system for each roll in the calender stack. For good temperature control the heat exchange units should be situated as near to the roll stack as possible, and for this reason one manufacturer places them horizontally beneath the roller conveyor unit. The water used by these heat exchangers should be chemically treated to prevent scale build up which leads to inaccurate control of roll temperatures and other heat transfer problems. Hot oil circulation units are sometimes installed by European machinery manufacturers to enable higher roll temperature to be used if these are necessary, but these units are much more expensive.

Both the top and bottom rolls of a three roll stack should be capable of movement in the vertical plane which is usually achieved by pneumatic pistons which are connected to slides in the end frames of the unit. The slides in turn control the movement of the rollers and permit a maximum daylight of approximately six inches between the top or bottom roll and the fixed middle roll when in the fully open position. Each nip separation should be known and finely adjustable by means of manually operated mechanical gauging devices. The calender roll stack should possess permanent safety stops to ensure that no two rolls can ever make accidental contact with each other.

The peripheral speed of the rolls is usually controlled by a variable speed drive acting through a worm gear or chain drive system. It is most important that the surface speed of all the calender rolls should be synchronised otherwise a shear will be imparted to the plastic sheet as it is pulled through one or both of the roll nips which may give rise to undesirable reversion or sheet surface defects. Also all the rolls should be concentric to within 0.001 in. otherwise if one or both nips are closed the thickness of the extruded sheet will fluctuate.

The rolls must be capable of rotating at constant slow speeds when thick sheet (e.g. 0.150 - 0.350 in.) is being extruded. The speed range of the rolls should be wide to cope with both thick and thin sheet and it should be carefully specified when ordering new calender roll stack equipment. For instance a 6 in. extruder producing 60 in. wide toughened polystyrene sheet at a rate of 600 lbs per hour should be capable of maintaining accurate roll surface speeds of approx. 1 - 10 ft./min if sheet with a thickness range of 0.040 - 0.350 in. is required. Obviously for lower outputs the surface speed of the rolls should be slower and for thinner or narrower sheet it should be faster. The surface speed of some roll stacks is as low as 6 ins./min. when extra thick sheet is being produced, but it may be as high as 60 ft./min. for very thin sheet.

(b) Roll conveyor unit

This usually consists of small (approx. 1 - 2 in) diameter parallel steel rollers which are not driven but support the plastic sheet as it travels from the three roll calender stack to the pull off rolls. The space between the latter two units is usually approx. 3 - 10 feet, and the sheet is gradually air cooled as it is conveyed over this distance. Forced air cooling by compressed air or cooling fans can be used if necessary.

(c) Roll pull off unit

This consists of a pair of rubber pull off rolls driven by a variable drive which is capable of a slightly wider speed range than the calender roll stack unit. These rubber pull-off rolls are usually 6 or 8 in. diameter and the pressure applied at each side of the top roll should be manually adjustable to ensure that the sheet is square (i.e. lined up correctly) as it approaches the guillotine or cutting device.

Although a separate drive is used for the pull-off rolls it should be an integral part of the calender roll stack unit. The surface speed of both units should be synchronised by mechanical, electrical or electronic means, because the difference between the surface speed of these two sets or rolls controls the strain or reversion properties of the plastics sheet. The surface speed of both roll units should be capable of separate adjustment, but a third control which speeds up or slows down both units simultaneously should also be available. Many combined units of this type incorporate a gauge which shows the percentage difference between the surface speed of the two roll units when not synchronised.

(d) Edge trim units

Each edge trim or longitudinal cutting unit consists of a pair of driven cutting wheels. The wheels possess bevelled cutting edges and are offset in the vertical plane so as to produce a shearing action. They can be situated on either side of the rubber pull down rolls and as the extruded sheet passes between them it is sheared almost through.

The horizontal distance between two edge trim units should be mechanically adjustable to the sheet width required and the vertical clearance between a pair of cutting wheels must be adjustable whenever it is necessary to alter the sheet thickness. A third pair of cutting wheels may be used if it is necessary to further subdivide the extruded sheet width (e.g. produce two sheets of half width). The drive for the edge trim units is usually an integral part of the rubber pull off roll unit.

(e) Sheet guillotine unit

This is often referred to as a shear unit and consists of a clamp device and a cutting blade which are automatically activated when the extruded sheet reaches an electric eye, microswitch or mechanical trip mechanism. The sheet length depends on the distance of such a mechanism from the far side of the cutting unit, this distance being adjustable. The purpose for which a guillotine is to be used should be carefully specified when buying new equipment because some models will not cut thick plastics sheeting. Guillotines can either be mechanically, hydraulically or pneumatically operated, the latter method is often preferred by plastics sheet manufacturers because it is the cheapest way of obtaining quick acting units. The normal cutting tolerance required by sheet fabrications is ± 1/16 in, although some fabricators now demand tolerances of ± 1/32 in, and this is difficult to achieve even with rapid response equipment.

(f) <u>Sheet delivery and automatic stacking units</u>

The sheet delivery unit consists of a system of driven continuous narrow bands which convey the sheet after the trimming and cutting processes to the end of the production line. An automatic stacking unit is sometimes used, consisting of a collapsible table which is activated when the sheet hits a stop at the end of the stacker. In many plants each sheet is inspected and packed as it comes off the production line and therefore an automatic stacking unit is not used.

11.2.1.2. The extrusion of thin sheet

Thin sheet (i.e. 0.010 - 0.040 in. thick) is either produced and cut into individual sheets by the process outlined in section 11.2.1.1., or the guillotine process is omitted and it is wound into rolls of continuous sheeting by conventional duel-turret constant tension winders which automatically adjust to the extruded sheet speed.

This TPS sheet is also produced by a process which was first developed by Plax in the U.S.A. and consists of stretching the sheet simultaneously in the longitudinal and transverse directions. In place of the calender roll stack, the take-off equipment is made up of two rows of water cooled chain driven metal nip rollers which grip the edges of the extruded sheet. These rollers can be carefully lined up so that they direct the edges of the sheet outwards from the die face in a parabolic manner. The positions of the small nip rolls govern the transverse stretch and the speed of the small nip rolls and a large pair of rubber covered pull off rolls (as described in section 11.2.1.1. (c)) govern the longitudinal stretch. To ensure that the sheet is stretched to a uniform thickness heating units positioned above and/or below the sheet are sometimes used. The biaxially orientated sheet produced by this method may be trimmed and cut into individual sheets of required size as described in section 11.2.1.1. (c), (d) and (e) or rolled into long lengths of continuous sheeting.

11.2.1.3. General design features

All the ancillary sheet take-off units mentioned in section 11.2.1.1. should be lined up on tracks running perpendicular to the sheet die. This enables them to be rolled back and allows easy access to the die face, and also helps to prevent misalignment of the take off units. All units should be provided with safety stops e.g. to prevent the calender roll stack or die lips being damaged by contact with each other. Large sheet take-off equipment is heavy and therefore a simple mechanical ratchet and lever device is often used to move each take-off unit along the guide tracks.

The optimum horizontal clearance between the die lips and the calender roll stack depends on the extrusion conditions, but it may be less than 1 in. The height of the first nip of the calender roll stack should be marginally below that of the die slot if the 'down stack' technique is used so that the sheet runs straight into the nip, for otherwise air trapping between the extruded sheet and the middle roll may be observed.

11.2.2. Methods of obtaining a surface gloss

Foil coating and heat polishing techniques are used to produce an extruded TPS sheet surface with a high gloss so that it is more eye-appealing and easier to clean. The problem is one of producing a cheap method for obtaining a high and uniform gloss across the full width of the sheet surface without seriously reducing the physical properties of the base sheet. Also the sheet must retain a high percentage of this gloss if it has to undergo subsequent reheating and thermo-forming (e.g. the production of refrigerator liners).

Foil coated TPS sheet normally retains a better gloss after subsequent reheating and thermo-forming than heat polished sheet, but it is usually more expensive and often possesses a lower impact strength than the latter.

Methods of heat polishing TPS sheet and the effect of different variables such as rubber content, melt temperature, and heat polishing conditions on the gloss and other physical properties of the extruded sheet have been studied by various workers (1.2.3.4.).

Most sheet producers differ in their views regarding the best polymer formulations and methods for producing extruded sheet materials which will meet the customers requirements. Much development work is still being conducted in an attempt to find economic formulations and processing techniques which will give better all round physical properties including surface gloss retention and increased impact strength to extruded sheet materials.

11.2.2.1. Ancillary take-off equipment for the production of foil coated sheet

A thin film of biaxially-orientated unmodified polystyrene is sometimes laminated to the upper surface of the extruded TPS sheet to give it higher gloss and greater sales appeal. A roll of this film which is usually 0.001 - 0.002 in. thick is mounted on a spindle above the top roll of the calender roll stack unit. The film is preheated by passage round the top roll on its way into the first nip of the roll stack where it is laminated to the base polystyrene sheet. The preheating temperature is critical because the film will shrink and pucker if too hot or fail to laminate if too cool.

11.2.2.2. Ancillary take-off equipment used for the production of heat polished sheet

There are three major methods for securing a high gloss on extruded toughened polystyrene sheet by heat polishing or heat glazing techniques, these are

(a) Continuous band technique

The sheet is extruded directly onto a highly polished flat stainless steel or chromed steel continuous belt which travels around two large cored drums. The drum nearest the die is oil heated to a temperature which is approximately equal to that of the plastic melt as it leaves the die, the other drum is water cooled. The plastics sheet produced takes the surface finish of the band, and possesses very low reversion which is often desirable for vacuum forming purposes etc.

(b) Non contact surface heating technique

The extruded sheet is passed through a heat polishing unit which is placed between the die and the calender roll stack. This type of heat polishing unit consists of a series of small diameter driven rollers which are water cooled to prevent the plastics sheet sticking to their surface. These rollers convey the plastics sheet under a number of banks of infra-red heaters which improve the gloss of the upper surface of the sheet.

The position of the infra-red heaters relative to the sheet should be adjustable so that the number of heaters switched on and their height above the sheet can be varied to give the best results without scorching the sheet.

(c) Calender stack technique

The sheet is extruded down a three roll calender stack in the usual way and a curved bank of infra-red heaters is positioned on the take-off side of the second roll. The distance of these heaters from the surface of the plastic sheet, as it passes from the first to the second nip of the three roll stack can be adjusted to give the best results for gloss.

Provided the lower roll is highly polished with no surface blemishes, the gloss imparted to the upper surface of the sheet by this method may be further improved by gradually raising the temperature of the lower roll until the sheet just adheres to its surface. Care must be taken to ensure that this lower roll is not overheated otherwise the plastic sheet will stick too firmly to its surface and will be carried back into the bottom nip.

11.2.2.3.Other methods for producing sheet materials with a high surface gloss

None of the processes outlined below are currently being used commercially but they have at various times been considered by the TPS sheet producing companies and are therefore included for the sake of completeness.

(a) Application of surface coatings using paint spraying techniques

(b) Pressing a polyester film against the hot sheet using the calender roll stack laminating technique, and then peeling it off

(c) Extruding a thin layer of crystal polystyrene onto the surface of the base toughened polystyrene sheet

Of these three methods the last probably has the greatest commercial potential, this process is currently being developed by BXP Ltd. (5).

The first two methods are not used commercially because of higher processing costs, poor retention of gloss with subsequent reheating and vacuum forming, or the adverse effect they have on the physical properties of the final product. The third method also makes the process more complicated to operate and is liable to lead to more scrap at least until it is fully developed.

11.3.Cellulose Acetate Sheet Take-Off Equipment

Wide cellulose acetate sheet take-off and laminating equipment is basically similar in design to that described for thick and thin toughened polystyrene sheet in sections 11.2.1.1. and 11.2.1.3.

11.4. Polyvinyl Chloride Sheet Take-Off Equipment

The take-off and ancillary equipment for both rigid and flexible PVC sheet extrusion lines is outlined below.

11.4.1. Slot die method

The ancillary sheet take-off equipment used for the extrusion of thick and thin rigid or flexible PVC sheet from a wide slot die is similar to that outlined in section 11.2.1.1.

11.4.2. Circular die method

This type of die is mainly used for the extrusion of rigid PVC sheet approximately 0.040 - 0.120 in. thick. The large diameter PVC tube is extruded vertically downwards and it is slit longitudinally by a knife blade at the die face. As the tube moves vertically downwards over a distance of approximately 3 - 4 feet it is gradually spread into flat sheet form by a spreader bar. Infra-red heaters are carefully positioned before the spreader bar to prevent the sheet from cooling too rapidly. The sheet is then turned through an angle of 90° and passes over a metal table, a second stationary metal sheet, which is sometimes weighted, is placed on top of the plastic sheet as it passes over the table to make sure that it is flattened. The plastic sheet then passes through a pair of driven rubber covered pull-off rolls the speed of which can be varied. This process does not give an extremely flat sheet but is of adequate standard for corrugating.

If long lengths of thick sheeting are required the sheet is directed onto two horizontal idling rolls which are separated by a gap of approximately 1 ft. and wound into large diameter rolls convenient for handling.

Thin PVC film and sheet may also be produced from a circular die by the layflat method e.g. as outlined by A.L. Griff (1).

11.5. Other In-Line Extrusion Take-Off Equipment

Equipment designed to manufacture specific end products (e.g. packaging film, roofing, partitioning, and floor tiles etc.) is often used in-line with the sheet extrusion take-off equipment. Sources of information relating to some of the more commonly used processes are given below.

11.5.1. In-line equipment for biaxially orientating extruded sheet

A full description of the in-line plant available for the biaxial orientation of thin sheet and film extruded from wide slot dies has been given in a recent review article (6). The packaging industry is the main commercial outlet for the toughened plastics sheet and film produced by these techniques.

11.5.2. In-line equipment for corrugating extruded sheet

A full description of the in-line corrugating equipment available to the plastics industry has been given in recent review articles (7,8,9). Equipment is available for corrugating thick rigid PVC extruded sheet in either the longitudinal or transverse direction. The building industry is the main commercial outlet for corrugated plastics sheet produced by these techniques.

11.5.3. In-line equipment for the production of floor tiles

The in-line take-off equipment available for the production of extruded PVC floor tiles has been reported in various technical journals (9,10,11). In this case particular care should be taken by an annealing process to ensure minimum residual shrinkage.

11.5.4. In-line equipment for embossing extruded sheet

Embossing units consisting of a water cooled embossing roll and rubber back up roll may be used in place of the calender roll stack to impart a three dimensional design to the surface of the sheet. The top roll of a calender roll stack is sometimes replaced with an embossing roll for this same purpose.

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12. SHEET EXTRUSION FAULTS CAUSES AND REMEDIES

12.1. Difficulties In Production

It is often difficult to obtain good quality sheet quickly after starting up a particular plastics sheet extrusion production line. This applies even if the extruder operator is provided with a complete set of previously successful running conditions for the production of sheet of similar size, surface quality and physical properties. Sheet extrusion operators are faced with many problems and sheet faults during a normal production run, and they must as far as possible be able to understand the cause of such faults and know how to overcome them.

It would be impracticable to describe in detail every known fault, cause and remedy, so to keep within reasonable bounds the most important ones which arise in producing thick TPS sheet (approx. 0.080 - 0.200 in) are outlined in chart form (see section 12.2.). For the purposes of this chart it is assumed that the extruder operator uses a three roll calender stack and the 'down stack' sheet production technique, also that he has had sufficient experience to enable him to make the necessary temperature setting, restrictor bar, die lip, screw speed and take-off adjustments etc. when a fault arises. It is further assumed that support facilities exist for testing raw materials and finished products, and that where appropriate, mechanical, electrical and instrument maintenance personnel can be called upon to find and correct faulty extruder and take-off components.

Because of the number of different foil coating and heat polishing techniques used for securing a surface gloss to extruded TPS sheet materials (see section 11.2.2.) many of the faults met by production personnel are not given. However, some of the more common surface defects which affect the gloss of sheet produced by these methods, together with an outline of the best methods for obtaining good gloss retention are given.

12.2. Chart of Faults, Possible Causes and Remedies

Fault	Possible Causes	Remedies
1. Extruder output surging badly	a) Feedstock materials give an uneven feed to the extruder screw if the individual particles are the wrong shape or size. This fault may occur when reground chips from off size scrap are being reworked. Also a fluctuating feedstock temperature may cause surging if preheated feedstock is used.	a) If reground chips from off size scrap are being reworked, reduce the proportion of reground to virgin chips being fed to the extruder. If necessary extrude all feedstock materials through a compounding extruder to produce granules of standard size. If preheated feedstock is used ensure that the feed temperature is constant.
	b) Bridging of feedstock in the extruder hopper. This fault may occur if the lubricant has been accidentally omitted from the feedstock.	b) Normally tapping the hopper or disturbing the granules in the base of the hopper will overcome isolated bridging problems. Check all lubricants etc. have been added in the required proportions to the feedstock material. Hopper attachments to overcome bridging problems may be necessary in extreme cases (e.g. powder feed materials).
	c) Wet feedstock material (see also 12.2.3.(b))	c) Make sure the feedstock is dry to within the normal commercial standards necessary for good quality sheet production.
	d) Solid plug of plastics material sticking to the screw, this fault sometimes occurs when the feed zone of an extruder is overheated. Also this difficulty may arise when a machine full of plastics is started up after standing idle for some time.	d) Check if the cooling facilities of the feedbox, barrel and screw are functioning correctly, extra cooling on the screw and/or barrel may overcome this problem. Occasionally a different feedstock material is used to purge the extruder of a solid plug of plastic, often feedstocks with a high bulk density or heated feed materials will overcome this problem. If necessary remove and clean the screw.
	e) Insufficient back pressure in the metering section of the screw.	e) Check the barrel temperatures and screw speed, make sure that the correct breaker plate and screen pack are in use. If the pressure at the front of the metering section can be measured and a valve is used on the extruder, the valve should be gradually closed until a steady back pressure similar to the value used on previous runs is obtained.
	f) Temperature recording instruments or heating and cooling systems incorrectly set or not functioning properly.	f) Check settings of all temperature controllers, make sure all thermocouples are seated correctly and all parts of the heating and cooling systems are functioning properly.

Fault	Possible Causes	Remedies
2. Poor gauge holding characteristics	a) Extruder output surging badly	a) See above in 12.2.1.
	b) Restrictor bar or die lips incorrectly set or speed of synchronised take-off equipment too fast or too slow.	b) Adjust the restrictor bar, die lips or speed of take-off equipment accordingly.
	c) Temperature of different die zones fluctuating badly or incorrectly set.	c) Check settings of die temperature controllers, ensure the thermocouples are correctly seated and the heating system is working correctly.
	d) Misalignment of one or more of the rolls of the calender stack.	d) Ensure all rolls of the calender roll stack are parallel and concentric about their own axis.
3. Longitudinal lines down the sheet as it leaves the die.	a) Continuous lines are formed by nicks in the die lips and dirt or decomposed particles adhering to the die lips.	a) Clean the die and die lips with brass tools and solvent. Die lips can be lightly stoned and polished to remove small surface scratches and blemishes, but if they are badly nicked they must be reground, polished and chromed. For this reason it is advisable to keep a second set of die lips in good operating condition.
	b) Continuous or non-continuous lines occur if there is excess moisture or other volatile material in the feedstock.	b) Dry the feedstock in ovens or use hopper drying equipment to ensure that the feedstock is dry to within the normal commercial standards for good quality sheet production.
	c) Dark lines on the surface of the sheet may be caused by decomposition products forming in the barrel and/or die. Dark lines or blemishes are also caused by plastic leaking from the die, discolouring and flowing on to the extruded sheet surface, leakages of this type suggest that the die has not been reassembled with sufficient care before use.	c) Reduce the die and plastics melt temperatures if decomposition products are not overcome by purging the extruder with a cleaning mix then both the extruder and die should be dismantled and cleaned. All plastic leaking from the die should be removed before it has time to flow on to the sheet surface.
	d) Continuous lines are sometimes observed if the die pressure is too low.	d) This fault is sometimes overcome by reducing the die temperature, or by treating as for moisture etc. (see 12.2.3.(b)).

Fault.	Possible Causes	Remedies
4. Longitudinal lines down the sheet as it leaves the calender roll stack, not present at the die.	a) Uneven thickness across the sheet, the thinner sections give a strip which is duller than the rest of the sheet.	a) Check the extruder is not surging (see section 12.2.1.). Adjust the restrictor bar accordingly to produce an extruded sheet of even thickness across its width. Check die zone temperature settings, thermocouples, and instruments. Only in extreme cases when the basic die design is at fault should the die zone temperatures or the die lips be adjusted to compensate for uneven plastic melt flow.
	b) Calender rolls are not parallel and therefore one side of the sheet will appear shiny while the opposite side is dull.	b) Adjust the stops on the calender stack nips until the rolls are parallel and correctly set. Maintaining the nip openings of the roll stack at slightly lower thickness values than the extruded sheet thickness often helps to remove surface defects and gives a smooth sheet with better appearance if the sheet is extruded at a high plastic melt temperature. Make sure the air pressures acting on the cylinders of both sides of the adjustable rolls of the calender stack are equal.
	c) One or more of the polished calender rolls may contain a band of deposit around its circumference or it may have been accidentally scored.	c) Clean the roll with solvent or replace it if badly damaged with a new roll.
	d) Shallow troughs are sometimes observed, particularly on the underside of the extruded sheet, caused by air being trapped between the sheet and the middle polished roll at the first nip of the calender roll stack.	d) Reducing the plastics melt temperature slightly by reducing the die temperature may overcome this problem. Make sure that the level of the die lips is slightly above that of the first nip of the calender roll stack, also ensure that the roll stack is as near to the die lips as possible to prevent the extruded sheet drooping on to the middle roll before it enters the first nip. If necessary lower the temperature of the middle roll.
	e) Lines can occasionally be seen when a thin foil of biaxially orientated unmodified crystal PS is laminated to the base sheet if the laminating conditions are either too hot or too cold, or if the foil is creased.	e) Heat or cool the top polished roll of a three roll calender stack accordingly, until the lines disappear. Make sure there are no creases in the polystyrene film either before or during the laminating process, if necessary change the foil.

Fault	Possible Causes	Remedies
	f) Longitudinal lines separating dull and shiny sections of the sheet can occur if parts of the sheet are sticking to one of the polished rolls. These may be accompanied by transverse lines if the sheet is not sticking evenly to the roll (see 12.2.6.(b)).	f) If it is intended to make the sheet surface adhere to a polished roll to impart a better gloss on the sheet the temperature of the roll should be gradually raised until this effect is achieved; external I.R. heaters on the roll stack may also require adjusting. If on the other hand it is not intended to make the sheet adhere to a polished roll then the temperature of the roll causing sticking should be lowered slightly. Ensure that all the rolls are parallel.
5. Transverse lines across the sheet as it leaves the die.	a) May be caused if the extruder is vibrating badly.	a) Make sure the extruder barrel and die are correctly supported to overcome vibration troubles.
	b) Curved flow patterns across the sheet caused by poor mixing.	b) Increase the back pressure at the front of the screw by using a valve or finer screen pack, this will also raise the plastic melt temperature but reduce the output, if the necessary drive power is available it may be possible to raise the screw speed to make up for this loss in output and still obtain a better degree of mixing. Lowering the screw speed or cooling the screw will also give better mixing characteristics, but these are also accompanied by loss of output.
6. Transverse lines across the sheet as it leaves the calender roll stack, not present at the die.	a) May be caused by the sheet take-off equipment vibrating or jerking badly.	a) Check take-off equipment is secured in position and make sure driving chains and belts etc. are not too tight or too slack.
	b) Intermittent sticking of plastic sheet to one of the calender rolls leaves a mark across the sheet where it has detached itself. Such marks do not occur if the sheet peels off smoothly.	b) Reduce the temperature of the polished roll causing intermittent sticking, or reduce the heat intensity on the sheet surface if external infra-red heaters are being used.
	c) Curved lines are caused when a large rolling bank of plastics material builds up in a nip because of a thick spot in the sheet.	c) Adjust the restrictor bar until the sheet is an even thickness, and make sure all the polished rolls are parallel to each other with the correct nip settings.
7. Pock marks.	a) These may be caused by air being trapped in the extruder because feedstock particles are the wrong shape or size.	a) Check size of feedstock particles, if necessary re-extrude through a compounding machine to produce particles of standard shape and size. Reducing the screw speed, raising the back pressure at the front of the extruder or lowering the rear barrel temperatures may also help to overcome this problem.

Faults	Possible Causes	Remedies
	b) Excess volatile material in the plastic sheet may also cause this fault. This fault is not always immediately apparent but may show up when the sheet is reheated for further fabrication e.g. thermoforming.	b) Make sure the volatile components of the feed material are within the normal commercial standards necessary for the production of good quality sheet. Some sheet manufacturers overcome this fault by using hopper driers, vented extruders or vacuum hopper equipment.
8. Silver spots	This is caused by poor lamination of biaxially oriented unmodified crystal PS foil to the upper surface of the base TPS sheet.	Raising the temperature of the top roll of the calender stack or drying the compound will usually overcome this fault.
9. Orange peel effect.	This surface fault usually arises when extra high impact polystyrene grades are extruded or when the plastic melt temperature is too low.	Raising the plastic melt temperature by raising the barrel and die temperatures or the back pressure at the front of the screw will help to overcome this problem.
10. Other surface blemishes.	a) Poor mixing in the extruder.	a) see 12.2.5.(b)).
	b) Foreign matter in the original feed e.g. decomposed material, coloured pellets, paper, string, dirt particles etc.	b) Check feed materials for contamination and remove all foreign matter if possible, otherwise scrap the faulty batch. If necessary clean extruder and die and replace contaminated screen pack.
	c) Nibs or dirt in crystal PS foil.	c) If crystal PS foil is being laminated to TPS sheet and this is giving nib or dirt trouble, replace with a better quality film.
	d) Regularly spaced surface blemishes may be caused by small particles sticking to one of the polished or rubber rolls.	d) Remove source of troubles with solvent or brass tool etc.
	e) Damage to the surface of the polished rolls of the calender stack.	e) Remove fault by stoning if damage is slight, otherwise replace the damaged rolls.
	f) Leakage from the die may cause dark surface blemishes (see 12.2.3.(c)). Also fume condensate dripping on to the sheet surface may produce brown spots on the sheet surface.	f) Make sure the die is reassembled correctly and periodically remove all plastics leaking from the die (see 12.2.3.(c)). Prevent all fume condensate from dripping on to the sheet surface.

Fault	Possible Causes	Remedies
11. Poor retention of surface gloss after reheating.	Loss of a high proportion of gloss when foil coated TPS sheet is thermo-formed may occur because the plastics melt temperature was too high during the sheet extrusion process and this in turn enforces a lowering of the calender roll temperatures, (N.B. the latter have a greater effect on gloss retention properties.)	Obtain the correct balance between the plastic melt temperature and the calender roll stack temperature. The optimum combination of plastic melt temperature and calender roll stack temperature depends on the grade of TPS material extruded and the method of applying a gloss to the extruded sheet surface. Usually the plastics melt temperature greatly effects the surface gloss as extruded but calender roll temperatures may have a greater effect on gloss retention.
12. Sheet discolouration.	a) Plastics melt temperature too high. b) Heat polishing unit too intense. c) Poor compound stability. d) Concentration of reground scrap in final mix is too high.	a) Check temperature instrument settings, thermocouple seatings and heating systems, reduce the barrel and die temperatures. b) Reduce the heat energy supplied to the heat polishing unit or increase the height of the unit above the sheet. c) This is seldom a problem with TPS although it often arises with PVC materials; check operating temperatures and make sure all stabilizers and lubricants have been added to the feed material. d) Reduce the percentage of reground scrap to virgin material. Reformulate if scrap has been reworked too many times and if necessary use for the production of darker coloured sheet.
13. Warped sheet.	This may be caused by low calender roll stack temperatures or uneven cooling of the sheet. It is also caused by excessive pull down between the calender stack and the rubber pull-off rolls.	Raise the roll temperatures of the calender stack and avoid uneven cooling of the extruded sheet. Adjust the tension on the sheet between the calender roll stack and the rubber pull-off rolls.
14. High reversion.	a) This fault in the base TPS sheet may be caused by a low plastics melt temperature and excessive pull down between the calender stack and the rubber pull-off rolls.	a) Raise the plastics melt temperature by raising the barrel and die temperatures or increasing the back pressure etc. Reduce the distance between the die lips and the first nip of the calender stack to a minimum. Make sure the peripheral speed of all the rolls of the calender stack are synchronised to each other and to the speed of the rubber pull-off rolls.

Fault	Possible Causes	Remedies
	b) This fault occurs in the foil layer which is laminated to the base TPS sheet, if the foil temperature has not been raised sufficiently during application.	b) This fault is detected by a curling of the sheet when heated. Raise the temperature of the top roll of the calender stack or raise the foil temperature after lamination by the use of external infra-red heaters positioned between the first and second nip of the calender stack. Synchronise the speed of the calender rolls and the rubber pull-off rolls.
15. Uneven reversion	a) Temperature variation across the die giving widely differing flow characteristics of the plastic melt through the die. b) Uneven sheet thickness. c) Excessive calender roll stack nip pressures. d) Take-off equipment out of line.	a) Make sure the die temperature is constant across its width. b) Adjust the restrictor bar until the sheet is a constant thickness across the width of the die. c) Reduce the calender stack nip pressures and make sure the pressures are equal on either side of the rolls. d) Make sure all take-off equipment is square to the extruder die.
16. Sheet not square.	a) Excessive pressure on one side of the rubber pull-off rolls. b) Guillotine out of line.	a) Adjust the pressure applied to each end of the rubber pull-off rolls until the sheet is square to the guillotine. b) Make sure the guillotine and other take-off equipment is square to the extruder die.

SUMMARY OF ANTICIPATED FUTURE DEVELOPMENTS PREVIOUSLY OUTLINED IN PARTS I AND II

The anticipated future developments for TPS, PVC and CA extruded sheet materials are individually summarised in this section. Trends towards the integration of sheet extrusion equipment with other ancillary processes are also discussed and anticipated future trends in the design and operation of extrusion equipment are outlined.

Toughened Polystyrene Extruded Sheet Materials

If the present growth rate for the consumption of PS and TPS materials throughout the World is maintained, and as yet there is no sign of it lessening, the annual tonnage consumed by the major polymer producing countries in the World will more than double between 1963 and 1970. In recent years the annual tonnage of TPS grades produced for extrusion and moulding purposes in the U.K. and the U.S.A. has gradually overtaken that of untoughened PS grades, this trend is likely to continue and by 1970 it is anticipated that approximately 60% of the combined tonnage of these two materials will be made up of TPS grades.

Almost all extruded polystyrene sheet materials over 0.010 inches thick currently being manufactured are based on TPS grades because of the poor strength characteristics of untoughened grades.

The selling price of both PS and TPS materials has been falling gradually over the years and this trend is likely to continue although no significant reduction in the manufacturing costs of styrene monomer is anticipated in the immediate future. The collapse of the market price of PS and TPS materials since 1958 has not deterred the larger oil and chemical companies from making large capital investments in new production plants and expansion programmes to existing plants. These plastics materials will therefore remain highly competitive because of the excess production capacity which exists in the World and this will result in an increased incentive to develop the more profitable specialised TPS resins.

New grades of TPS sheet materials with improved physical properties are continually coming on to the market e.g. better low temperature toughness, greater resistance to environmental stress cracking, improved surface appearance and thermo-forming properties etc. Anti-static and non-taint grades will also rapidly increase in popularity over the next few years. It is mainly the excellent thermo-forming properties and low cost which will ensure the continual expansion of existing and newly developed TPS sheet materials and applications.

The production of refrigerator liners is still the largest single market outlet for extruded TPS sheet materials in the U.K. and the U.S.A. although the overall annual percentage of TPS sheet consumed by this application has been gradually falling. In the past the limiting load bearing properties of thermo-formed TPS refrigerator liners have tended to restrict the use of this extruded sheet material to refrigerators with a capacity less than 6 cu. ft. Recently, however, new methods of insulation based on modern polyurethane foaming in-situ techniques have revived interest in this application particularly in the U.S.A. where large domestic refrigerators and deep freeze units are more popular. Most of the initial problems encountered with the polyurethane forming in-situ process have now been solved.

The annual sales of refrigerators in both the U.K. and the U.S.A. has been static for the past 3 to 4 years, however an extended hot dry summer and the current low competitive prices of refrigerators could easily cause a boom in this market, particularly in the U.K. where only 30% of homes possess a refrigerator. There is likely to be greater competition in the refrigerator market from extruded ABS sheet as the price of this material falls because ABS possesses better low temperature strength properties than normal rubber toughened PS grades and it also gives an improved surface gloss in natural extruded sheet form.

Because of the seasonal nature of the refrigerator industry there has in recent years been a strong tendency for extruded TPS sheet producers to manufacture thin sheet for the packaging industry since this business is not seasonal and helps to keep the extrusion machines fully employed at all times of the year. The possibilities in this field of packaging are almost inexhaustible and consequently extruded TPS sheet for this application is expanding rapidly. For instance the recent development in the U.S.A. of mass produced thermo-formed cigarette packets from thin TPS sheet has caused great excitment in the plastics sheet extrusion industry because the potential annual tonnage absorbed by this application alone could be greater than that currently used to manufacture refrigerator liners. A list of other packaging fields which thermo-formed TPS sheet is likely to break into within the next 2 to 3 years is given in section 2.4.4.1.

New and cheaper methods of improving the surface appearance of extruded TPS sheet materials are constantly being sought and new outlets are being developed for printed film laminates and embossed plastics sheet materials.

Polyvinyl Chloride Extruded Sheet Materials

The excess World production capacity and low profit margins which have existed for PVC resins over the past few years have not deterred major PVC resin producers from investing large sums of money in expansion programmes and the building of entirely new PVC polymer producing plants. However in spite of this very rapid growth a temporary World shortage, particularly of speciality PVC resins, arose towards the end of 1963 and this has helped to stabilise and even raise the selling price of these resins. Prior to this the very low selling price of general purpose resins has encouraged polymer manufacturers to develop and produce the more profitable speciality grades.

If the present expansion rate for the production of PVC resins is maintained, and as yet there is no sign of it slowing down, the U.K. and the U.S.A. will double their 1963 capacity by 1970 and the Common Market countries and Japan are likely to treble their capacity, over the next seven years. New PVC resin grades with improved toughness, weathering and heat and light stability properties etc. will be developed and introduced on to World markets.

Extruded PVC sheet materials have been produced for a number of years in Italy and Germany and they have now established a firm foothold in the U.K., U.S.A., and Japan. The packaging and building industries form the major outlet for extruded PVC sheet materials and it is mainly from these sources that future expansion can be anticipated.

The trend in recent years indicates that thin PVC sheet markets in the U.K. and the U.S.A. is only expanding slowly whereas thick sheet markets are growing rapidly. There are a number of reasons for this such as the wider use of PVC flooring materials for office and domestic purposes. The greatest immediate growth potential in the U.S.A. and the U.K. for thick extruded rigid PVC sheet is in the field of roofing materials for the building industry. This application will continue to expand and increase in popularity because rigid PVC sheet is a strong, relatively cheap, lightweight material which is capable of transmitting light. Competition in this market is already very keen and thinner gauge sheets (e.g. 0.035 in. to 0.040 in) are now being marketed in an attempt to lower selling prices. A corrugated PVC sheet which incorporates a mesh of fine wire has been introduced on to the market by the Japanese in an attempt to overcome restrictive building regulations, this type of sheet will soon be mass produced in the U.K. and the U.S.A.

There is also a large potential market for thinner gauge (0.010 in. to 0.040 in) extruded rigid PVC sheet particularly in food packaging applications, see section 3.4.4. (a), but as yet little inroad has been made into this field.

Applications for extruded plasticised PVC sheet materials greater than 0.040 in. thick are growing rapidly, particularly in Germany and the U.S.A., see section 3.4.4. (b). There is little chance that extruded flexible PVC sheet less than 0.020 in. thick will compete directly against calendered sheet before 1970, however, development work is now in progress and these two methods of thin PVC sheet production may compete directly against each other at some time in the future.

Already the extrusion of thick rigid and flexible PVC sheets is competing successfully against calendered and pressed sheet. The trend towards extrusion as a method of sheet production will continue in cases where there are large orders and a good commercial outlet for particular grades and sizes of sheet materials. Sheet extrusion lines have certain advantages over calendering equipment, for instance they are more versatile in that they can produce continuous sheet over a much wider thickness range. Extrusion equipment is much cheaper than calendering equipment, and it also has the advantages of handling plastic melts better and reducing degradation of the sheet material by oxidation products.

Cellulose Acetate Extruded Sheet Materials

The CA plastics industry has expanded gradually over the years but the annual tonnage currently being produced is small in comparison to PVC and TPS materials. Very few new applications are being developed and the future of CA extruded sheet materials lies mainly in the production of spectacle frames and packaging materials. The spectacle frame market is the largest single outlet for extruded CA sheet materials and this will continue to expand slowly. Newer and more ingenious methods of producing a wide variety of coloured configerations in extruded CA sheet for the spectacle frame industry using dual extrusion techniques etc. will continue to be developed.

Because of the interest shown in extrusion grades of CA there will be more development work conducted by CA manufacturers in an attempt to improve the appearance of extruded CA sheet materials by reducing the gel and nib content etc. Although CA extruded sheet materials are likely to have a firm future in the spectacle frame market, there is a strong case for the production of ABS, TPS or PP spectacle frames for under-developed countries. Injection moulding techniques have long been used for the commercial production of cheap sunglasses and there is no reason why plain coloured spectacle frames (e.g. black) should not be manufactured on a large scale for countries such as Africa, India and China where greater consideration is given to the numbers produced rather than to fashion trends.

Integrated Plant

In modern industry there is a marked tendency towards the integration of production plant, if this is at all possible, in order to reduce manufacturing costs.

Depending on the final application there may be a good case for the partial integration of a number of separate units, e.g. polymerisation plant and colouring, sheet extrusion and further fabrication equipment. For instance it may be beneficial to have two or more machines in close proximity to one another. Often the conversion of existing plants to fully integrated systems does not warrant large scale investigation because of the large number of technical and breakdown problems which have to be solved. One of the major disadvantages associated with fully integrating sheet extrusion equipment with other ancillary production processes is that all parts of the plant must be geared down to work at an output rate similar to that of the slowest component. Also if one unit breaks down the entire line will be out of action, the flow of plastics material must therefore be stopped or restricted and this may give rise to considerable difficulties and wastage of materials, particularly if continuous polymerising methods are being used.

As yet no company uses fully integrated polymerisation and sheet extrusion equipment presumably because only relatively small proportions of plastics materials are consumed by the sheet extrusion process and also because of the numerous technical difficulties e.g. colouring and breakdown problems. For economic running polymer plants rely mainly on numerous other moulding and extrusion outlets for the sale of their produce.

Although some companies find it worthwhile to position sheet extrusion equipment in-line with further fabrication equipment these are not usually fully integrated. For instance some of the larger refrigerator manufacturers in the U.S.A. have found it worthwhile to extrude their own TPS sheet before vacuum forming this into refrigerator liners for final assembly. This equipment is not however fully continuous because the vacuum forming stage is a batchwise process. Currently there is only one refrigerator manufacturer in the U.K. seriously thinking of installing a TPS sheet extrusion line. The anticipated trend by the larger refrigerator manufacturers to extrude their own sheet materials has not taken effect presumably because the relatively large capital expenditure involved is not worthwhile in view of the current low competitive prices of extruded TPS sheet materials. However, in recent years most refrigerator manufacturers have installed vacuum forming equipment for the production of refrigerator liners and this has had the advantage of eliminating the profit made by custom moulders and considerably lowering packaging and transport costs.

In the packaging industry machines are being developed which will take hot sheet from the extruder and thermoform it, using continuous in-line equipment. Although in theory, continuous operation has the advantages of producing savings in labour costs, heating and forming time cycles, and storage space, it sacrifices versatility and independence of machine operation. For instance manufacturers of disposable containers (e.g. cups, cartons etc) usually produce rolls of extruded plastics sheet and then feed individual rolls to the thermo-forming machine. In this way stoppage of the entire production line is avoided if either the sheet extruder or the thermo-forming machine break down, provided a surplus stock of rolls is kept. Such a process may be more efficient if it is not fully integrated because one thermo-forming unit may cope with the output of sheet from more than one extruder.

The same advantages and disadvantages apply to companies who use both sheet extrusion machines and thermo-forming equipment to produce corrugated rigid PVC sheet for the building industry. Only in a few instances are these two in-line machines fully integrated.

The patent situation and the relatively large capital expenditure involved in producing extruded CA sheet will probably continue to restrain spectacle frame manufacturers from purchasing CA sheet extrusion equipment.

The general public are gradually being educated to regard domestic appliances etc. as disposable units and this together with the trend by some final product manufacturers to extrude their own sheet materials may result in a lowering of standards (e.g. the surface appearance of extruded sheet materials) over the next few years.

The Extrusion Process

Derived single screw extruder flow equations have in the past been subjected to certain restrictive assumptions to keep mathematical equations within workable limits (e.g. Newtonian fluids, isothermal or adiabatic operating conditions etc.). Because of these restrictions modern theories cannot generally be applied quantitatively to the extrusion of most plastics materials, however, they do give rise to a better qualitative understanding of the extrusion process.

A working knowledge of what goes on inside a single screw extruder has gradually been built up in recent years by observation and theoretical derivation and this work will continue. More attention will be given to the derivation of theories for the extrusion of non-Newtonian melts and more development work will be conducted on feed particle shape and size, and the effect of lubricants in the feed materials. Theories which derive expressions for the flow of plastics materials through complicated die structures will become more advanced and greater emphasis will be placed on experimentally determined extrusion equations for large scale production plant.

There are still large gaps in our understanding of extrusion scale up work and further attempts will be made to fill these gaps. The science of extrusion has been studied in great detail over the past ten years and this work will provide a firm foundation for future studies but it is doubtful if such great strides will be made in the advancement of extrusion theories over the next decade. More work will be done to determine the coefficient of friction of different plastics compositions on a large variety of metal surfaces over a wide temperature range, and this will help future workers to derive quantitative theoretical flow equations for non-Newtonian plastics materials.

The length to diameter ratio (L/D) of single screw extruders has been gradually increasing over the years but because of considerations of bending, torsion, and machine reliability this is now approaching a maximum value of 30/1 for screws which are only supported at one end. If in the future longer screws are required, attempts may be made to support the screw at both ends to prevent excessive wear occurring. Although the maximum sizes of extruders currently available in terms of screw length and diameter are not likely to be increased over the next decade, the maximum screw speeds at which these machines will be operated almost certainly will be increased to enable higher output rates and hence lower conversion costs with heat stable polymer mixes. The most popular range of extruder diameters used for large scale sheet extrusion plant will remain at 4.5 to 6 inches. In the past there has been a tendency for single screw extruders to be underpowered but this is now being corrected and modern equipment e.g. drive motors, gear boxes, thrust bearings etc. are adequate to take the extra loadings necessary to give increased outputs at higher screw speeds.

The constant pitch single start screw will continue to gain in popularity but greater attention will be given to screws which are specially designed to improve the mixing efficiency of the single screw extrusion process (e.g. the Maillefer screw). There is little hope of any standardisation with twin screw designs and these will continue to be used mainly for rigid PVC materials.

Although many new designs of compounding extruders giving improved mixing and compounding characteristics will continue to come on to the market there will be a natural tendency to decrease production costs by omitting the intermediate compounding costs and converting coloured dry blend mixes containing fine powders, beads, flake or plasticised feedstock materials, directly into sheet form.

Thus wider use will be made of the more positive methods of regulating the feed of plastics materials to the extruder (e.g. Crammer Feeders etc.)

Other equipment which is designed to ensure better quality control such as drying systems and devolatilising extruders will continue to gain popularity, in the latter case barrel extraction systems are usually preferred in comparison to screw extraction systems. Better quality control will also be achieved by the careful use of extruder valving systems and fuller instrumentation (e.g. greater use of equipment which will accurately measure and record plastic melt pressures and temperatures), there is little doubt that the use of these components will become more widespread over the next few years particularly in the case of plastics sheet extrusion equipment.

At present there is little automation on a sheet extrusion line to help the operator, usually only such items as the barrel and die temperatures, and a sensitive feed probe which is used for maintaining a constant head of material in the feed hopper, are self adjusting. However, recently attempts have been made to develop equipment which will self correct for any unwanted variation in the thickness of extruded plastics sheet. The big obstacle to increased automation on a sheet extrusion line is its very high initial capital cost and the many technical difficulties which have yet to be overcome. The sheet extrusion process will therefore be manually controlled for many years to come but the idea of a sheet extrusion line being controlled by a computer may well become a reality at some time in the future.

No new important breakthrough in the design of large sheeting dies is anticipated in the next few years, die manufacturers are likely to follow established and proven trends. There will be a tendency for die design to become more streamlined and as stated above greater efforts are being made to produce a self correcting device which will be used on sheet extrusion lines to ensure automatic rapid response thickness control to within the normal commercial limits. A few years ago some die manufacturers developed presses and clamping devices which surrounded the die, these clamps were designed to replace the large number of bolts which are usually used in die assembly and thus allow quick and easy dismantling of large sheet dies for cleaning purposes. However, these devices either for reasons of bulk or problems associated with plastic leakage from joints do not appear to have been fully accepted by the plastics sheet extrusion industry.

Only in a few isolated cases is there likely to be a demand for sheet greater than 60 in. wide, the majority of extruded plastics sheet is less than 48 in. wide and this trend is likely to continue. In many cases (e.g. CA for spectacle frames and plasticised PVC for floor tiles) the width of sheet produced has been reduced to a narrow band in order to maintain greater control over the process and/or simplify further fabrication techniques.

APPENDIX 1.

A CRITICAL STUDY OF PLASTICS

SHEET EXTRUSION PROCESSES

The thesis will deal with the technical and administrative factors involved in the extrusion of Polystyrene, Polyvinyl Chloride, and Cellulose Acetate sheet on a commercial scale, from the buying of the raw materials to the sale of the finished products.

The general scheme will be as follows:-

1. An Historical Introduction

The early development of polymers and plant which led to the extrusion processes up to the present day.

2. The Market Requirements

Customer requirements for extruded plastic sheet and competitive materials.

3. The Raw Materials

A broad outline of the economic and technical factors involved in the production of the three polymers.

4. The Extrusion Process

a) Essential features of the process including mass and heat transfer, and the behaviour of plastics melts.

b) The design of extruders involving considerations of engineering construction, screw and die design, instrumentation, properties of plastics materials to be processed and product requirements.

c) Design of ancillary equipment.

d) Operating techniques.

e) Polymer storage and handling.

f) Comparison with other competitive sheet manufacturing processes including capital and operating costs.

5. Further Fabrication

The equipment necessary for fabricating extruded plastics sheet into finished products for specialised uses.

6. Integrated Plant

The advantages and disadvantages of a single plant in which the processes of polymerisation, extrusion and further fabrication are integrated as one unit.

7. Future Development

Assessment of likely future developments in both the manufacture and marketing of extruded plastic sheet.

APPENDIX 2.

GLOSSARY OF TERMS

A full definition of most of the terms in this report will be found in "Nomenclature of the Screw Extruder" SPE Journal p.295 March 1962. Other terms used are listed below.

Adiabatic extrusion	The ideal case in which no heat is transferred either in or our of the plastics material but heat is generated by mechanical working of the plastic which raised the melt temp.
Autogenous extrusion	This is the practical case of "adiabatic" extrusion in which there is no deliberate heating or cooling as such. Heat is generated by mechanical working of the plastics material and part of this heat is lost naturally from the system by conduction etc.
Isothermal extrusion	The ideal case in which the melt temp. of the plastic is uniform as it passes through the extruder or a section of it.
Polytropic extrusion	The practical case where heat is generated by mechanical working of the plastics material and heat is also supplied to or removed from the system by deliberately heating or cooling the screw or barrel.
Newtonian fluid	A fluid in which the shear stress between two adjacent layers in relative motion is proportional to the rate of shear (i.e. the viscosity is independent of the rate of shear).
Non-Newtonian fluid	A fluid whose viscosity varies with the rate of shear. Most plastics melts and solutions are non-Newtonian and the viscosity usually increases with increasing shear rate.
Isovels	Lines passing through points of equal velocity of flow.

Drag flow	Drag flow is the major component of forward flow in an extruder; it is the forward conveying action produced by the relative motion between the screw and the barrel and it is theoretically equivalent to the maximum output of an extruder operating at open discharge.
Pressure flow	Pressure flow is a component which opposes the forward flow of material in an extruder. It arises when a restriction such as a die or breaker plate is attached to the front of an extruder so causing a build-up of pressure, and hence giving rise to a pressure gradient in the screw channel. Pressure increases towards the restriction and acts to push material back along the screw channel down the gradient. Note that pressure flow is only a component, and no net backward flow actually occurs.
Leakage flow	Leakage flow arises when material tends to flow back over the top of the screw lands.
Transverse flow	Transverse flow does not contribute to the net forward movement of material in an extruder; it is a circulating flow which takes place across the screw channel and contributes to the mixing efficiency of the process.

APPENDIX 3.

MATHEMATICAL NOMENCLATURE

All units are expressed in terms of the following units and they remain unaltered in the plural.

Length - inch (in)
Area - square inches (in^2)
Volume - cubic inches (in^3)
Time - second (sec)
Mass - pound (lb)
Temperature - degrees Celsius (^oC)
Angle - degrees (deg or ^{o})

The mathematical nomenclature employed in this report is listed below.

C	Constant $\frac{(D - 2h)}{(D)}$
C_a	Average specific heat of plastic melt ($in - lb/in^3/^oC$)
D	Screw diameter (in)
E	Constant $(D - h) D$
F_{CD}	Channel curvature factor for drag flow (dimensionless)
F_D	Drag flow shape factor (dimensionless)
F_P	Pressure flow shape factor (dimensionless)
F_Z	Power shape factor (dimensionless)
H	Fractional distance from screw surface
K	Constant $\left[\frac{E(\tan \phi + f_s)}{1 - f_s \tan \phi}\right]$
L	Length of screw full of plastic material (in)
L_a	Axial flighted length of screw (in)
M	Heat energy lost by conduction and radiation from the extruder (in - lb/sec)
N	Screw speed (rev/sec)
P	Rise in pressure along the screw (ΔP) (lb/in^2)
Q	Volumetric rate of discharge from screw (in^3/sec)
Q_D	Volumetric drag flow along screw channel (in^3/sec)
Q_L	Volumetric leakage flow across screw flights (in^3/sec)

Q_P	Volumetric pressure flow along screw channel (in ^{3}/sec)
R_1	Radius of screw root (in)
R_2	Radius of screw (in)
S	Velocity perpendicular to screw flight (transverse flow velocity) (in/sec)
T	Plastic melt temp. (°C)
T_i	Initial plastic temp. (°C)
ΔT	Temp. rise in plastic material ($T-T_i$) (°C)
U	Velocity of barrel surface (in/sec)
V	Velocity in down channel direction (in/sec)
Z	Power required to turn the screw (in-lb/sec)
a	Ratio of pressure to drag flow (for maximum output $a = 0$, for no output $a = 1.0$)
b	Constant
e	Axial flight width (in)
e	Base of natural logarithms
f_b	Coefficient of friction between the plastic and the barrel surface
f_s	Coefficient of friction between the plastic and the screw surface
h	Depth of screw flight (in)
k	Die constant
m	Resultant velocity of a particle derived from down channel (V) and transverse channel (S) flow (in/sec)
n	Number of starts to the screw
p_1	Pressure at inlet end of the feed section (lb/in^2)
p_2	Pressure at discharge end of the feed section (lb/in^2)
r_1	Radius of curvature of flight at trailing edge (in)
r_2	Radius of curvature of flight at leading edge (in)
t	Flight lead (in)
v	Local velocity of particle at point (x.y.) (in/sec)
w	Channel width, measured at right angles to the screw helix (in)

x	Distance measured along the x axis (x varies 0 - w) (in)
y	Distance measured along the y axis (y varies 0 - h) (in)
z	Distance measured along the helical z axis (in)
α	Constant depending on screw geometry
β	Constant depending on screw geometry
γ	Constant depending on screw geometry
δ	Radial clearance between top of screw flight and barrel (in)
θ	Angle of direction of solids movement, measured from a reference plane perpendicular to the screw axis (deg)
μ	Viscosity of plastic melt (lb/in^2 /sec)
$\bar{\mu}$	Average viscosity of plastic in screw channel (lb/in^2/sec)
μ_i	Initial viscosity (lb/in^2/sec)
μ_L	Viscosity of plastic in radial clearance between top of screw flight and barrel (lb/in^2/sec)
ρ	Density lb/in^3
τ	Shear rate sec^{-1}
ϕ	Helix angle of screw, measured from reference plane perpendicular to the screw axis (deg)

APPENDIX 4

The Quality Of Mixing Achieved With Screw Y. (Pineapple Torpedo)

Breaker Plate Assembly
Type C (60 mesh)

Breaker Plate Assembly
Type D (120 mesh)

Breaker Plate Assembly
Type E (200 mesh)

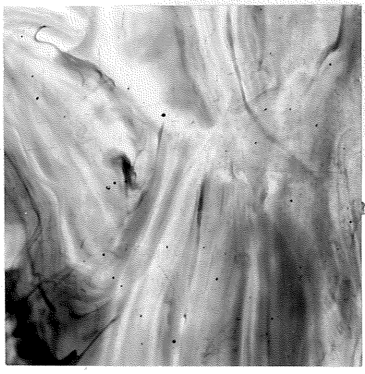
215 lb/hr.

30
rev/min.

205 lb/hr.

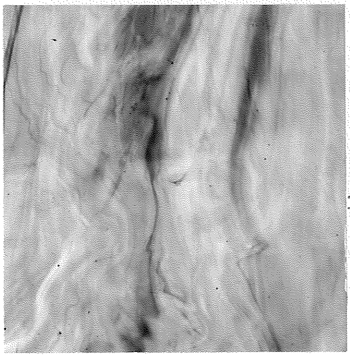
30
rev/min.

200 lb/hr.



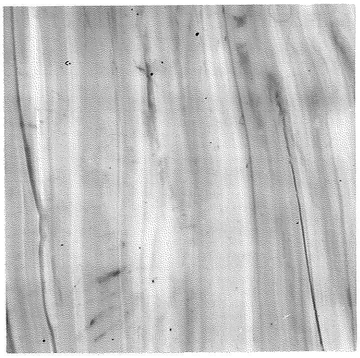
330 lb/hr.

60
rev/min.

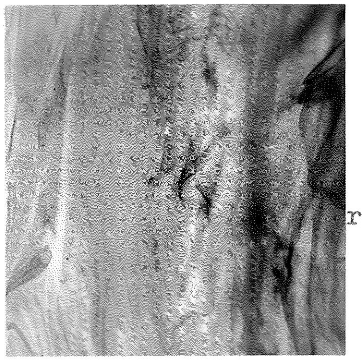


315 lb/hr.

60
rev/min.

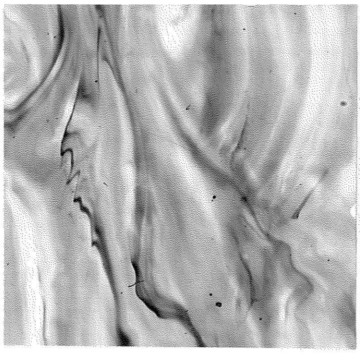


310 lb/hr.



415 lb /hr.

90
rev/min



400 lb/hr.

90
rev/min.

395 lb/hr.

APPENDIX 5

Comparison Of The Quality Of Mixing Achieved With All Three Screws At Maximum And Minimum Values Of Output

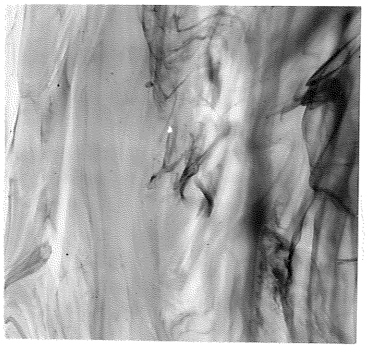
Screw X
(Dow Torpedo)

Screw Y
(Pineapple Torpedo)

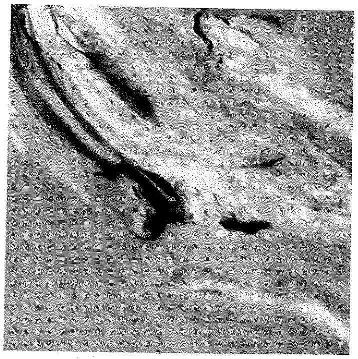
Screw Z
(No Torpedo)

Extruded At A Screw Speed Of 90 rev/min. Using Breaker Plate Assembly Type C. (60 mesh)

350 lb/hr.

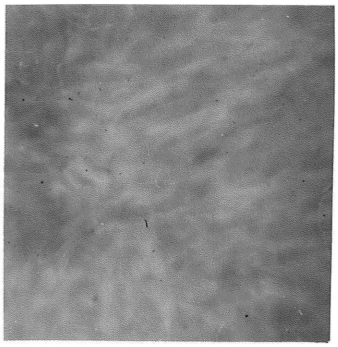


415 lb/hr.



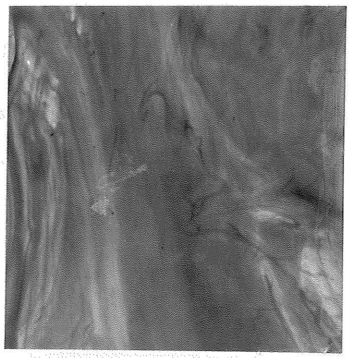
420 lb/hr.

Extruded At A Screw Speed Of 30 rev/min. Using Breaker Plate Assembly Type E. (200 mesh)



150 lb/hr.

200 lb/hr.



210 lb/hr.

APPENDIX 6

Comparison Of The Quality Of Mixing Achieved With Screws X. and Y. At Various Screw Speeds Using Breaker Assembly Type D. (120 mesh)

Screw X. (Dow Torpedo)

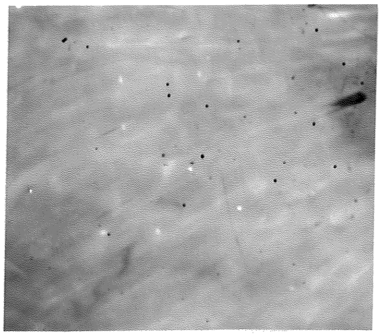
30 rev/min.

160 lb/hr.

60 rev/min.

255 lb/hr.

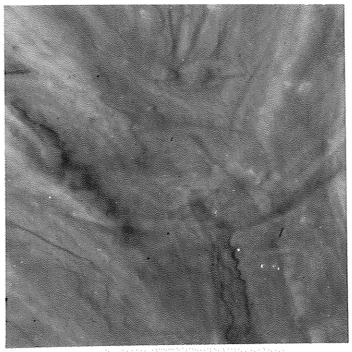
90 rev/min.



335 lb/hr.

Screw Y. (Pineapple Torpedo)

30 rev/min.

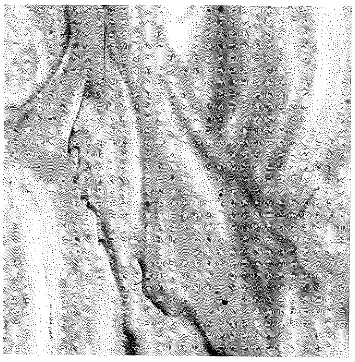


205 lb/hr.

60 rev/min.

315 lb/hr.

90 rev/min.



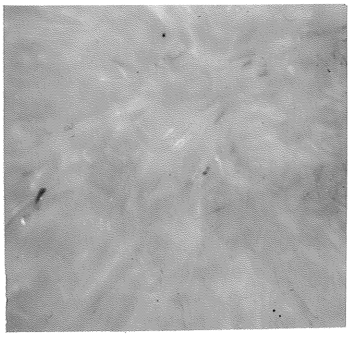
400 lb/hr.

APPENDIX 7

Comparison Of Quality Of Mixing Achieved With Screws X. and Z. At 30 rev/min. Using Various Screen Packs

Screw X. (Dow Torpedo)

Breaker Plate Assembly
Type C. (60 mesh)

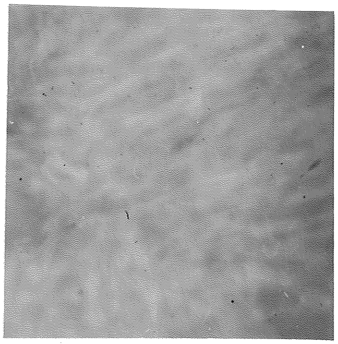


170 lb/hr.

Breaker Plate Assembly
Type D. (120 mesh)

160 lb/hr.

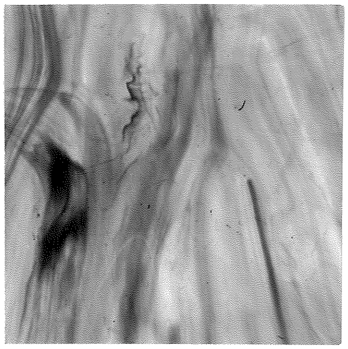
Breaker Plate Assembly
Type E. (200 mesh)



150 lb/hr.

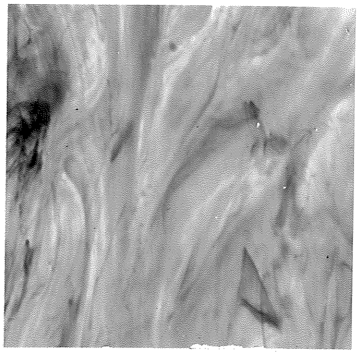
Screw Z. (No Torpedo)

Breaker Plate Assembly
Type C. (60 mesh)



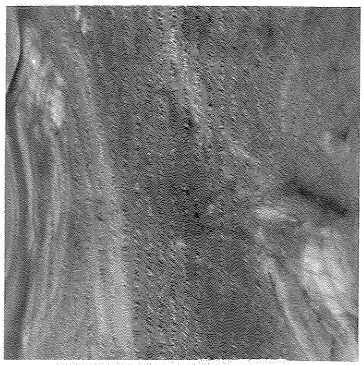
240 lb/hr.

Breaker Plate Assembly
Type D. (120 mesh)



225 lb/hr.

Breaker Plate Assembly
Type E. (200 mesh)



210 lb/hr.