RAPRA TECHNOLOGY LTD.

Screws For Polymer Processing II

A One-Day Seminar

organised by
Rapra Technology Limited

held at

Rapra Technology Limited
Shawbury, Shrewsbury, Shropshire SY4 4NR

14th May 1998

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Contents

Scale up of Extruders, Practicalities and Pitfalls

J.A. Colbert

Betol Machinery Ltd., UK

ABSTRACT

Most material and product development is carried out on a small scale, initially in Company Laboratories, Research Centres, or Universities. The successful developments are then put into large scale commercial production. However, some of these are doomed to failure at worst, or costly modifications, because insufficient attention has been paid to the task of scaling up.

This paper [8] stresses the importance of understanding scale up especially the energy balance involved, not just at the time of designing a production plant but also when carrying out the initial research. Two different approaches are used, one for single screw and one for twin screw extrusion.

SINGLE SCREW SCALE UP

This is not a new subject by any means, but despite the fact that some of the references used, and the basis for the approach advocated here, were published over 20 years ago, it is still poorly understood. Far too often processes are developed at a laboratory scale without due consideration being given to the longer term large scale production.

The extrusion process involves the transfer of energy from electrical power via viscous shear to an increase in the enthalpy of the polymer. This is then followed by a process whereby the enthalpy is reduced by transferring energy to a cooling media such as water or air. Hence it can be seen that, to fully understand the process and then achieve successful scale up, we must understand the heat transfer for processes taking place and the overall energy balance.

Energy Balance

An energy balance for a single screw extruder is illustrated in **Figures 1** and **2**. Essentially there are four main components:

Mechanical Work (E). The main drive motor, less any transmission losses, transfers its energy via viscous shear to the polymer.

Barrel Heater/Cooling (H). Heat transfer to and from the polymer is limited by the internal surface area of the barrel which is an important factor when considering scale up.

Enthalpy Change of Polymer (I). If the temperature of the feedstock and the extrudate can be measured then, by reference to enthalpy vs temperature data, available for most polymers, the energy input to the polymer can be found.

Heat Losses (S). These are difficult to calculate and are often considered as the balancing term in the equation. However, from work carried out on a range of extruders it has been possible to compile a table showing estimates for heat loss on a range of sizes and barrel temperatures. (**Figure 3**). These should only be used as a guide as obviously different guarding or insulation will make large differences to the heat loss.

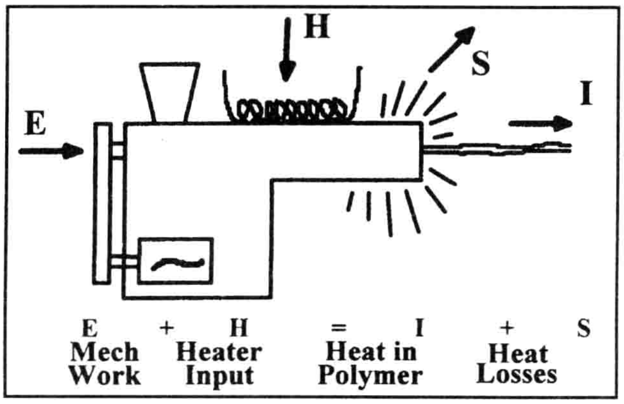

Figure 1

Extruder Energy Balance

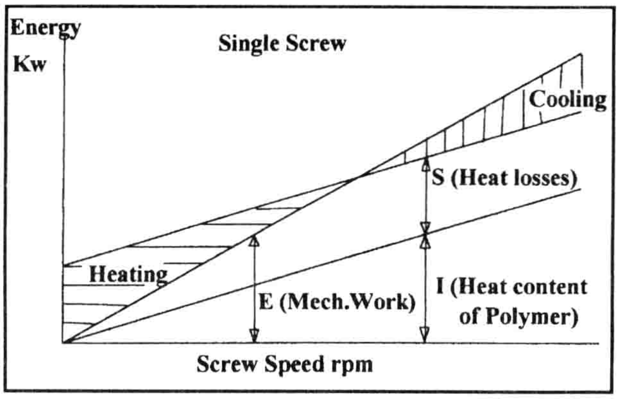

Figure 2

Energy Balance

These four components can be simply illustrated on a plot of Energy vs Screw Speed as shown in **Figure 2**. As output is broadly proportional to screw speed then, for similar temperatures, both the Mechanical Work input to the polymer, and the increase in enthalpy of the polymer will also be proportional to screw speed. With the barrel temperature set to a constant figure then the heat losses should be constant regardless of screw speed and at zero speed heat losses will be balanced by the barrel heaters. It can then be seen that, as screw speed is increased, the amount of barrel heating

Barrel Temp Deg.C	Barrel Dia. inches	Barrel Dia. mm	Radiative Heat loss W/in2	Convective Heat loss W/in2		Heat loss factor		Heat loss KW	
				Single	Twin	Single	Twin	30D single	35D twin
150	2.0	50.8	1.20	1.12	0.73	0.029	0.037	**3.48**	**5.13**
250	2.0	50.8	3.16	2.28	1.48	0.068	0.088	**8.16**	**12.35**
350	2.0	50.8	6.51	3.58	2.33	0.126	0.168	**15.14**	**23.52**
150	4.5	114.3	1.20	0.73	0.70	0.024	0.036	**14.64**	**25.59**
250	4.5	114.3	3.16	1.48	1.43	0.058	0.087	**35.27**	**61.78**
350	4.5	114.3	6.51	2.33	2.24	0.111	0.166	**67.13**	**117.85**
150	6.0	152.4	1.20	0.70	0.70	0.024	0.036	**25.65**	**45.49**
250	6.0	152.4	3.16	1.43	1.43	0.057	0.087	**61.93**	**109.82**
350	6.0	152.4	6.51	2.24	2.24	0.109	0.166	**118.14**	**209.50**

Figure 3

Heat Loss for Single and Twin Screw Extruders

required to balance the equation decreases until an adiabatic condition is reached. Above this speed barrel cooling is then utilised more and more, increasing with speed. Obviously this is a simplification utilising a global view of the process, and really each barrel zone could be viewed separately. However, there is a range of screw speed over which the process will be running close to the adiabatic condition.

At this condition the process is almost independent of heat transfer from the inner surface of the barrel to the polymer or vice versa. This is an important consideration from the point of view of scaling the process. If the process is reliant upon heat transfer then the limiting factor becomes the surface area of contact between barrel and polymer. This will only scale up by the ratio of extruder diameters to the power of 2. The volume within the extruder, on the other hand, will increase by the ratio of extruder diameters to the power of 3. Hence, by operating at or near to adiabatic conditions we are giving ourselves the potential to scale up by a larger factor.

Thermal Scale Up

The thermal considerations of scale up were recognised by Schenkel [2] as early as 1963 where he utilised a "thermodynamic parameter" in his suggested rules for scale up. This parameter is defined as the ratio of, the increase in temperature due to barrel heating at the start of the process, over the total increase in temperature due to both barrel heating and viscous shear effects once melting has commenced.

$$\text{Thermodynamic parameter } \psi = \frac{T_1 - T_0}{T_2 - T_0}$$

Although 'T_0' the feedstock temperature, and 'T_2' the final melt temperature were easy to measure, the real problem lay in establishing the value of 'T_1'. However, as the value of ψ must lay between O and 1 and it was possible to set rough limits on T_1, it was found that the thermodynamic parameter ψ normally lay between 0.3 and 0.7, although at the true adiabatic condition it would be equal to zero.

This approach worked fairly well for large, slow running polyethylene densification extruders where barrel temperature had a significant influence but tended to become inaccurate with faster running extruders.

DIMENSIONLESS NUMBER SCALE UP

Dimensionless numbers are the basis for scale up in a wide variety of applications utilising as they do critical dimensionless factors relevant to the particular process. One that most people will be aware of is

$$\text{Mach No} = \frac{\text{Velocity of Body}}{\text{Velocity of Sound}}$$

where Mach 1 is the speed of sound, or the point at which a body travelling at that speed is said to "break the sound barrier".

Another that is used extensively in the analysis of fluid flow is

$$\text{Reynolds No.}(RE) = \frac{\text{Inertia Forces}}{\text{Viscous Forces}} = \frac{\rho\sigma d}{\mu}$$

This is said to get critical at Re = 2×10^3. Above this value the flow is turbulent, and below this flow is laminar. Reynolds No. for flow in an extruder channel is around 10^{-1} to 10^{+1} which is why we treat polymer melts as laminar flow.

To utilise dimensionless numbers however we need to define which numbers are relevant to our particular process. Work was carried out on this in the early 1970's and various papers on the subject were presented in particular some by Prof. Pearson [6]. These identified Graetz and Griffiths numbers, both dealing with thermal considerations, and Q/Wbh, a volumetric efficiency, as being the three numbers most relevant to the process.

Graetz No. (Gz)

$$Gz = \frac{\text{Convection of heat from end to end of channel}}{\text{Conduction of heat to / from walls}} = \frac{H^2 V_z}{\alpha L}$$

Where V_z = Downstream velocity of Polymer $= \frac{Q}{bh}$

h = Channel depth

L = Length of section

α = Coefficient of Thermal Diffusivity $= \frac{K}{\rho Cp}$

As with Mach and Reynolds No there are significant levels of Graetz No.

Gz<1 — Temperature of polymer controlled by wall temperature. The flow will be fully developed.

1 <Gz <10 — The local wall temperature has less effect than does the previous history of the material.

10 <Gz — Wall temperature has no effect. All temperature changes will be caused by viscous heat generation.

Griffiths No. Gf

$$Gf = \frac{\text{Temp. diff. across channel generated viscous polymer flow}}{\text{Temp. sensitivity of viscosity}} = \frac{bomV2}{K}$$

Where V = Peripheral velocity of screw π ND

K = Thermal conductivity

b_0 = Temperature increase needed to decrease viscosity by $\frac{1}{e}$

μ = Viscosity

The significant levels for Griffiths Number are

Gf <1 — Insignificant generation of cross channel temperature gradients. Isothermal analysis will be valid.

1 <Gf <10 — Large cross channel temperature differences are generated. There will be errors in isothermal analysis.

10 <Gf — Isothermal analysis no longer relevant. Flow pattern dominated by generated temperature effects.

Volumetric Efficiency (Q/Wbh)

This number is obtained by taking the output equation for the metering section of an extruder, as derived from the Navier stokes equation, and making it dimensionless.

Throughput (Q)

$$Q = \text{Drag Flow}\left(\frac{Wbh}{2}\right) - \text{Pressure Flow}\left(\frac{h^3 b}{12\mu}\frac{dp}{dz}\right)$$

$$\text{Hence } \frac{Q}{Wbh} = 1/2 - \frac{h^2}{12\mu W}\frac{dp}{dz}$$

Where Q = Volumetric Output

W = Peripheral Speed of Screw

b = Channel Width

h = Channel Depth

Significant values are:

Q/Wbh = 0.33 Maximum Pressure Generation

= 0.5 Constant Pressure

> 0.5 Pressure Drop Towards Die

By keeping these three numbers constant, and utilising the power low equation for viscosity, a set of scale factors can be established that are just dependent upon the power law index 'n' in the following equation;

Viscosity = Constant * (Shear Rate) $^{(n-1)}$

The scale factors, where D = ratio of screw diameter, are as follows;

	Newtonian	**Non-Newtonian**
Channel Dept	$D^{1/2}$	$D^{\left(\frac{n+1}{3n+1}\right)}$
Screw Length	D	D
Channel Width	D	D
Screw Speed	D^{-1}	$D^{\left(\frac{-2n-2}{3n+1}\right)}$
Throughput	$D^{3/2}$	$D^{\left(\frac{5n+1}{3n+1}\right)}$
Power	$D^{3/2}$	$D^{\left(\frac{5n+1}{3n+1}\right)}$
Specific Output Q/N	$D^{5/2}$	$D^{\left(\frac{7n+3}{3n+1}\right)}$

SCALE UP IN PRACTICE

There are a variety of pitfalls one can find when scaling the single screw extruder, some of which are listed below;

a) *Heat transfer* - As mentioned earlier if the original process is dependent upon heat transfer to or from the extruder barrel then scale up is limited by the surface area. Unfortunately this is often the case with laboratory machines as the ratio of surface area to process volume is very high, and this ratio decreases significantly with production machines. It is therefore imperative that laboratory development should pay attention to the energy balance, and how close the operation is to the adiabatic condition.

b) *Basic Design Factors* - There are other dimensionless numbers related to the design of an extruder that also need to be kept constant. There are really simple ratios relating certain measurements to the diameter, such as

Length is expressed in diameters	L/D
Flight clearance is normally	D/1000
Flight width is normally	D/10

c) *Screw Speed* - This must be scaled according to the factors already set out. Too often laboratory extruders are run very fast and it is then not possible to achieve the scaled speed on the production machine.

ORIGINAL SCREW Drg No.									SCALED UP DESIGN Drg No.						
Barrel Dia. (ins)	Channel depth Feed (ins)	Inter. (ins)	Meter (ins)	Output Kg/hr	Screw Speed rpm	Specific Output Kg/hr/rpm	Scale up factor "n"	Polymer	Barrel Dia. (ins)	Channel depth Feed (ins)	Inter. (ins)	Meter (ins)	Output Kg/hr	Screw Speed rpm	Specific Output Kg/hr/rpm
4.5	0.600	0.410	0.205	400	90	4.44	0.90	Nylon	6.0	0.696	0.475	0.238	613	67	9.16
4.5	0.600	0.410	0.205	400	90	4.44	0.70	Nylon	6.0	0.703	0.480	0.240	607	66	9.25
4.5	0.600	0.410	0.205	400	135	2.96	0.50	PP	6.0	0.713	0.487	0.244	598	96	6.26
4.5	0.600	0.410	0.205	400	135	2.96	0.40	PP	6.0	0.721	0.492	0.246	592	94	6.33
						Compromise Design			6.0	0.715	0.488	0.245			

Figure 4

Scale Up Using Dimensionless Number Approach

d) *Un-Scaled Items* - The feedstock is obviously unchanged so feed throat design is more critical with smaller machines. This can mean that a laboratory machine may be feed limited whereas the larger machine is not. This can alter the nature of the process so attention must be paid to the possibility of a feed limitation.

e) *Pressure Similarity* - If an extruded product is being produced, such as a tube of certain diameter, then for a larger extruder, with a higher output, the pressure will also increase. This will again change the process considerably. For true scale up the pressure at the end of the screw must be close to the original pressure.

With compounding this is easily achieved by increasing the number of strands and keeping the output per die hole constant. With sheet a wider die can be used, or blown film a larger diameter die.

With tube however, where a fixed diameter is required, the pressure will change and hence the melt temperature of the product. This will then cause problems with sizing the product and downstream cooling.

f) *Universal Screw* - Different polymers do not scale in the same proportions hence a screw that works well on two polymers at one size of operation may scale up to two different designs at a larger diameter. This is illustrated in **Figure 4** where a 4.5 inch screw running polypropylene and nylon was scaled up to a 6 inch design. Even though this is only a small increase it can be seen that two different designs are arrived at. Eventually only one 6 inch screw was manufactured but the final design was a matter of using experience to produce a compromise design.

SCALE UP OF TWIN SCREW EXTRUDERS

This was the subject of a previous paper [7] and although dimensionless numbers or design ratios were used, the approach was more empirical. The critical design ratios to keep constant were:

Centre Line Ratio - The ratio of the distance between the centres of the shafts divided by the radius of the barrel.

Barrel Length L/D - Length of the barrel expressed in terms of the number of diameters.

Screw Clearance - This is expressed as a fraction of the diameter and is normally of the order D/100

As co-rotating twin screw extruders are segmental in design there is a lot of flexibility in screw design. Hence, although screw design should be similar, it is relatively easy to change the design slightly to modify the process. Also with the process being starve fed there is an extra variable that can be changed, feed rate, again to modify performance. This means that accurate scale up is not so critical as with single screw, but it still needs to be understood, and the pitfalls can be just as painful.

Essentially the output potential and power required will scale up under adiabatic conditions by the diameter ratio to the power of 3. If however, there is a need for some heat transfer the scale up factor will reduce to around 2.5. In practice the majority of co-rotating twin screw extruders scale up by a factor between 2.65 and 3.0.

CONCLUSION

Scale up should not be left until the production machine is required but should be included in the experimental planning to ensure that full consideration of scale up limitations is included in the development plan.

An understanding of the energy balance is critical to future scale up as thermal considerations form the basis of the key dimensionless numbers in the scale up technique. A knowledge of the viscosity/shear rate characteristics of the polymer, and hence the power law factor 'n' is required for scale up of the single screw extruder.

Co-rotating twin screw extruders can also be scaled up but with their increased design flexibility a more empirical approach can be adopted.

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BIOGRAPHICAL NOTE

John Colbert

John Colbert graduated with an honours degree in Mechanical Engineering fron City University, London, in 1970. He joined ICI Plastics Division on extrusion research, then Technical Service. Before leaving ICI in 1987 he was ICI's adviser on all forms of polymer extrusion technology. In 1987 he joined Baker Perkins, (now APV), as Process Engineering Manager, then Technical Director, and finally as General Manager, He is experienced with all types of extrusion, single screw, twin screw and continuous mixers, as well as with polymer processing in general. He is currently Technical Director of Betol UK, who manufacture complete lines for the production of fine tolerance tubing, multi-layer film and sheet, as well as twin screws for compounding.

Notes

Present Problems with Screw Extrusion

F. Fassihi, P. Prentice, J. B. Barry and F. Gao
The Nottingham Trent University, UK

INTRODUCTION

The closer collaborations between different countries and the campaign of the single currency in Europe tend to promote the competition for improvement of processing efficiency in industries between EU countries. A marginal difference in processing efficiency in the future could mean the difference between failure and success of business and loss and creation of jobs. Productivity is a measure of processing efficiency and is defined as the value added per employee. **Table 1** shows the data of productivity of major European countries in 1982 and 1992. It is important to note that Britain comes last in the order of productivity. The difference column shows the same order. If this situation continues, many industries in Britain will soon lose their competition to their rivals in Europe so that improvement of processing efficiency would play more and more important role in protection of future British business.

Table 1. Productivity of Major European Countries in 1981 and 1992

Country	1982	1992	Difference
Italy	20,420	45,615	25,195
France	21,543	39,752	18,295
Germany	21,457	38,871	17,414
UK	19,717	34,197	14,480

In view of these facts, a research programme has been lunched by the Polymer Engineering Centre (PEC) at The Nottingham Trent University aimed at identifying the problems associated with low productivity and seeking solutions to improve processing efficiency in screw extrusion industry. This paper will give a brief review of the results obtained from survey of small and medium enterprises (SMEs).

SURVEY PROCEDURES

Improvement of machinery performance is considered to be an effective way to increase productivity. The machinery performance can be improved by three ways: investment in new machinery, upgrading the existing machinery and optimising the operations. The first option is the credited cause for the massive increased productivity rate amongst the leading nations. If this approach was to be adopted the productivity increase would be higher but investment cost would be also higher. Furthermore, the extent of improvement was not quantifiable and therefore relative gains could not be measured in real terms. This option might not have been appropriate to the SMEs in particular by creating within them an awareness of their manufacturing processes.

The second and third options can be realised by identifying the extent of spare capacity within the plastic processing industry and improving work practices which demand more training as well as evaluation of the processing strategy. This promoted the PEC to conduct the "polymer processing efficiency audit" with the objective of utilisation of the expertise in the centre to support the industry in general and SMEs in particular by creating within them an awareness of their manufacturing process.

To evaluate the processing industry it was necessary to look at a range of processes. However, since the expertise of the centre was in the single screw extrusion, which is the basis of a majority of converting processes including injection moulding, it was concluded that a survey of the process using single screw extrusion was likely to reflect the overall situation in this industry. The survey was based on a random selection of over 1200 companies in a mail shot. A total of 184 returns were received. Of these 43 companies had gone away, 7 has gone bankrupt, 5 refused or were unable to give information or other. This left a total of 129 who had agree to take part in this exercise. Of this number a total of 90 respondents were from companies with screw extrusion as their main process and 39 using other processes.

Of this a random sample of 50 companies from the extruder sector were selected and a plan was drawn up to visit each of the companies in turn with their agreement. In the end, only 43 companies were able to supply the requested information. 85% of them had employees less than 250. Geographic distribution of these companies is summarised in **Table 2**. Majority of the companies were located in England.

Table 2 Geographic Distribution of the Companies Surveyed

Location	No. of companies	%
Scotland	1	2.1
Northern Ireland	2	4.3
Wales	4	8.5
England (North)	13	27.7
England (Midlands)	15	31.9
England (South)	12	25.5

Two set of data, i.e. business information and technical data were collected from these companies. The business information would help to identify the possible shortcomings while the technical data were used for finding the extent of spare capacity in the processing processes.

BUSINESS INFORMATION COLLECTED

The following information was gathered during visits to the participating companies:

1) 29% of the companies involved had more than one product.
2) 100% of the companies manufactured to order.
3) 40% of the companies designed to order. Most companies stated that the design work was on conjunction with their capabilities along with customer requirements.
4) 55% of the companies maintained stock.
5) 68% of the companies adhered to ISO 9000 Part 2 (production quality). One company was aiming for ISO 9002 Part 1 (design quality).
6) 55% of the companies had someone with a HNC/HND/Degree. This could be misleading as there was only one person with a qualification. The qualification was not usually in plastics or production engineering.
7) 93% of employees had been at the company for more than 5 years.
8) 38% of employees had been at the company for more than 10 years.
9) 76% of the companies claimed that staff had attended courses, seminars or exhibitions. However in virtually all cases it was senior staff alone that attended these activities.
10) 46% of the companies claimed it was easy to replace workers.
11) 100% of the companies had a desktop computer of some description.
12) 23% of the companies used Electronic Data Interchange
13) 83% of the companies said that their business was driven by high demand.
14) 64% of the companies said that their business was driven by effective marketing.
15) 76% of companies said that they had overseas business - this was mainly in Europe and tended to be a small part of their business.
16) 85% of the companies said that the industry is price sensitive.
17) 95% of companies said that production had increased over the last 5 years. This could be more to do with business growth rather than technological advances although it is felt that that too could have played a part.

From examination of these results, it can be drawn that education, control of product quality and information exchange are the major areas which could be further improved. The skill of a operator plays an important role in control of processing in the optimum condition and reducing waste. Lack of any formal qualification was the fact in 45% of the companies visited. For those 55% of companies having someone with a qualification, the person in charge of processing is usually a chemist who had little knowledge in polymer rheology and polymer engineering. This could be due to the lack of polymer engineering content in the majority of engineering degree courses which makes chemistry and polymer science graduates more suited for the job. In addition to these, discussion with industrialists also revealed that they do not consider the training programmes on offer as benefiting their companies or representing good value for money.

Quality of products is directly relevant to processing technology and conditions. 68% of the companies visited were aspiring to ISO 9000. This means that a good proportion are seeking quality. However it is significant that more than 30% of the companies were not concerned with quality measures.

Information exchange is an effective way to seek technical support and market information. Computer equipment in those companies visited was insufficient. Most of them had not connected to the internet system. However this survey was conducted before 1996. Situation might have changed in the past two years since the rapid development of network technology.

TECHNICAL PROBLEMS AND SOLUTIONS

The technical data collected covered material characteristics, screw geometry and operating conditions. This data was then analysed via computer simulation of screw extrusion process. The computer simulation programme was originally developed by Imperial College and was successively upgraded by the Polymer Engineering Centre in the past two decades.

The simulation involves repeated running of the programme with different screw geometry until a satisfactory combination of output rate and melt quality is achieved. The computer predictions have been extensively tested and compared with experimental data obtained using a fully instrumented screw extruder. Such a solution was also tested through a case study based on an extrusion process of manufacturing plastic bags in B&H Plastic Ltd. in Nottingham. The processing equipment comprised a 55.5 mm diameter single screw extruder, 24D in length. The film bubble was inflated from a tube produced from an annular die 200 mm in diameter with a gap of 1.4 mm. The material was a blend of LD, HD and LLDPE. The screw design was optimised via the computer simulation. The computer designed screw was then manufactured by Stanley Vickers Ltd. The data on output rate at various motor speeds for both old and redesigned screws are plotted in **Figure 1**. It can be seen that a significant improvement of output rate has been obtained with the new designed screw. The company also claimed that the machine was much quieter with less vibration at higher screw revs using the new screw.

A similar exercise was carried out with other companies. The results of the audit is summarised in **Table 3**. The information in the table is based on calculating the percentage increase in calculated output using a redesigned system compared with the existing output rate. The average value is calculated by discarding the best and the worst value and averaging the remaining data.

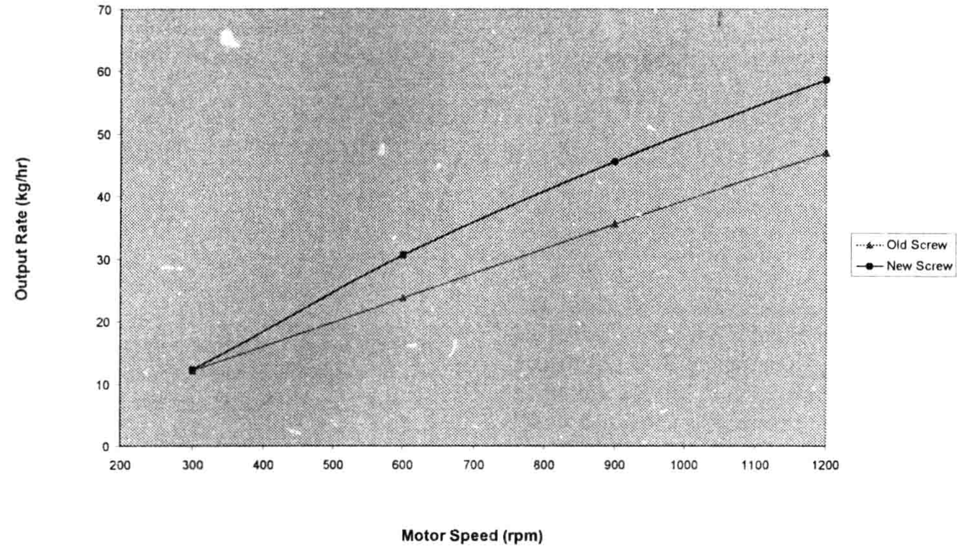


Figure 1

Output Rate at Various Motor Speeds for Old and Redesigned Screws

Table 3. Efficiency Improvement After Optimisation Via Computer Simulation				
No. of Companies	**Material**	**High (%)**	**Low (%)**	**Average (%)**
25	PE	94	0	30.7
8	PVC	485	220	343
8	PS	213	0	37
2	PP	3	0	1.5
1	PET	14	14	14
1	ABS	9	9	9

This data indicates that most of the companies did not run their processes at optimum capacity. The output increment via computer optimisation is significant. Increase of productivity can be achieved by minor technical upgrading.

REMARKS

The study was based on the evaluation of screw extrusion in small and medium enterprises. The data collected suggest that there is a spare capacity in UK plastic processing industry. The current problems include poor awareness of manufacturing processes, insufficiently qualified employees, insufficient awareness of product quality and significant spare capacity of processing. Computer simulation of optimisation of processing has proved to be an effective method to explore the spare capacity in present production processes. Further education of employees and promotion of interactions between industries and universities would be helpful to overcome other problems.

ACKNOWLEDGEMENTS

We would like to thank DTI and ERDF for financial support for this work and participating SMEs for their collaboration in this programme.

BIOGRAPHICAL NOTES

Fengge Gao

Dr. Fengge Gao is currently a Senior Lecturer and the Manager of the Polymer Engineering Centre at the Nottingham Trent University. He obtained his BSc and MSc in Polymer Science and Engineering in China and PhD in Composite Materials from Loughborough University. Since his graduation in 1982, he has worked as an Engineer with Xinjing Iron and Steel, a Lecturer at Shengyang University of Chemical Technology, a Research Officer with Loughborough University and a Research Fellow at both the University of Surrey and University of Salford. He has more than 16 years of research experience in a wide area of polymer science and engineering. His major current interests include polymer processing, flame retardancy of polymers, polymer recycling and polymer degradation.

Notes

What Screw?

Karol Braun

ER-WE-PA Davis-Standard, UK

INTRODUCTION

The forerunner of the single-screw extruder, as we know it today was first applied to plastics processing between the wars; initially for PVC and later for Polyethylene. By late 1940's the basic layout of today's extruder had already been adopted. However, the concept of the single-screw extruder as a pump goes back several thousand years. Today extrusion is without question the most important process in the manufacture and processing of plastics. In addition to the better-known finishing applications, such as film, sheet or pipe, extruders are used as melt pumps, reactors, compounders and reclaimers, as well as being an essential part of the injection moulding and blow moulding processes.

This paper is mainly concerned with extruder screws for finished-product applications, although many of the considerations may well also apply to other processes. The distinctions between different processes are often blurred, e.g. end-product extrusion may be combined with reclaim or with compounding.

The principal end-product processes are:-

- Film, Blown and Cast
- Sheet
- Pipe and Profile
- Coating and Laminating
- Wire and Cable Covering
- Filaments and Strapping

These processes have some common and some differing requirements. Some of these are examined in the context of screw design and operating conditions.

THE BASIC SCREW

The basic single-stage screw is illustrated in **Figure 1**. The majority of single-screw extruders through the 50's and 80's used this type of screw and the only variables were the dimensions.

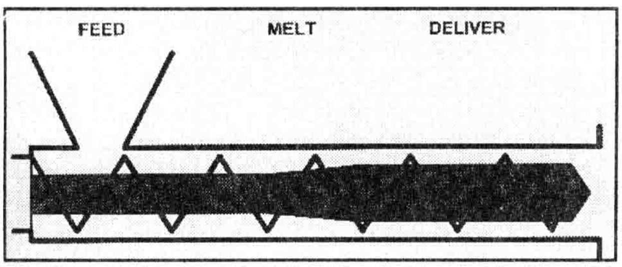


Figure 1
Single Stage Screw

The fundamental aim of any extrusion process, **Figure 2**, is to deliver to the die a homogeneous melt at a steady rate, acceptable melt temperature and sufficient throughput for economically viable operation. The simple screw, as illustrated is very limited in its ability to achieve all these requirements, Output is normally determined by the dimensions of the metering section which will also provide a small measure of distributive mixing, but little or no dispersive mixing. If the metering section is shallow output may be limited by polymer overheating as screw speed increases and if it is deep the pressure generating capability is reduced, the screw may become prone to surge and melt uniformity is likely to be poor.

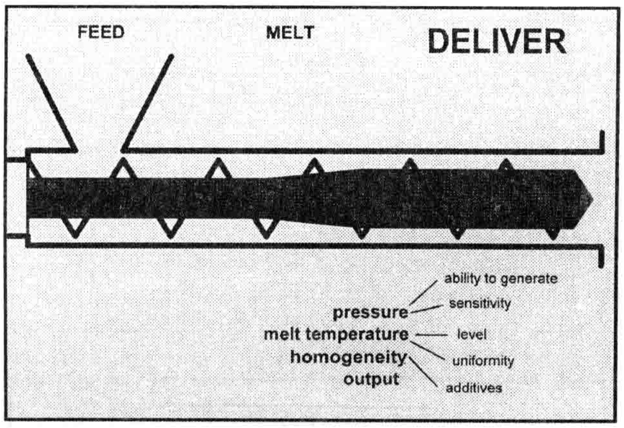


Figure 2
The Extrusion Process

MIXING DEVICES

It became clear early on that the simple screw configuration was incapable of satisfying the requirements of many applications and a number of mixing devices began to appear. These broadly fell into two categories of **Dispersive** and **Distributive** mixing. Distributive mixing is required to ensure adequate uniformity of the melt and dispersive mixing is needed to break down agglomerates of additives or unmolten polymer.

The distributive mixing devices are generally positioned at the end of the screw as indicated in **Figure 3** and take the form of slotted disc or pineapple pattern, their purpose being to achieve random spatial orientation within the polymer melt (i.e. homogeneity). Typical examples of this type of mixer are shown in **Figures 5** and **6**. There are several variants on this theme. Another type of distributive mixer shown in **Figure 4** consists of one or more circumferencial rows of pins positioned in the channel of the metering section of the screw.

With all these mixers the aim is to achieve maximum uniformity with the least possible input of energy and at the

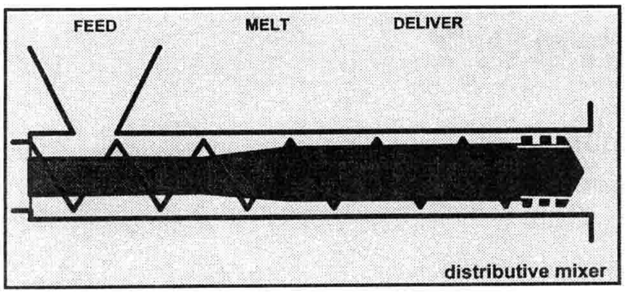


Figure 3
Distributive Mixing Device

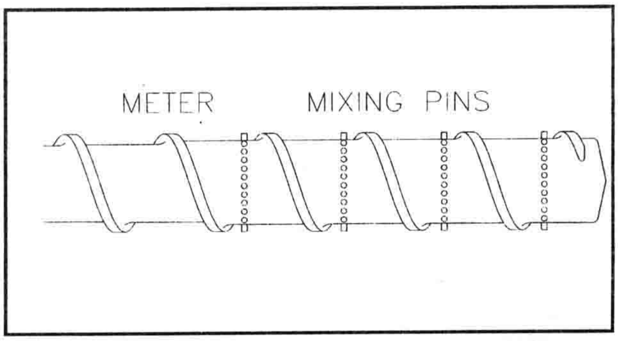


Figure 4
Pin Mixing Section

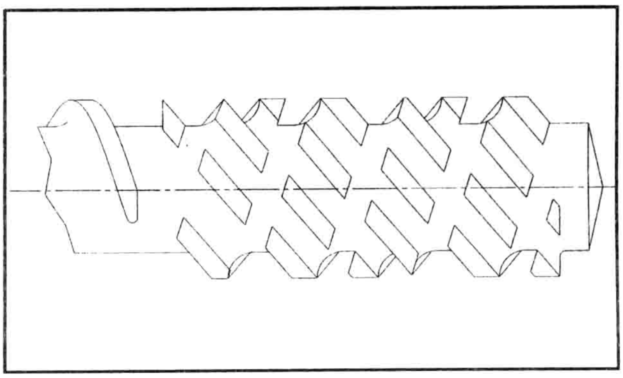


Figure 5
Saxon Mixing Section

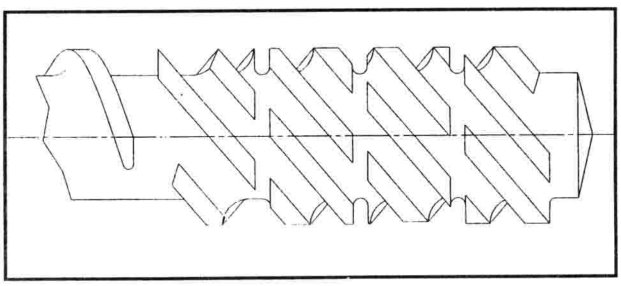


Figure 6
Dulmage Mixing Section

same time to minimise any hang-up of material that might lead to polymer degradation, gel formation and poor purge performance. Perhaps surprisingly, the round pin mixer is not particularly good in this respect.

The dispersive mixers shown diagrammatically in **Figure 7** are normally placed near to but not necessarily at the end of

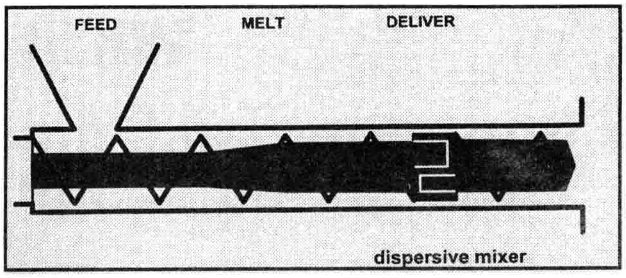
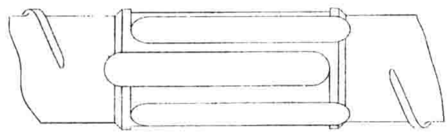


Figure 7
Dispersive Mixer (diagrammatically)

the screw and generally take the form of a barrier which subjects all the melt to intensive shear by forcing it through a small gap. The critical dimension is the clearance over the barrier. The further back along the screw the mixing element is positioned the greater the mixing effect but at the expense of lower output and higher melt temperature. Examples of this type of mixer appear in **Figures 8, 9,** and **10**.

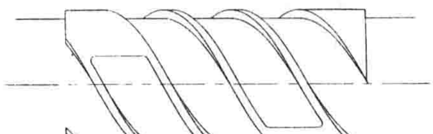


Screw Diameter	# Of Channels	Screw Diameter	# Of Channels
Less Than 2½"	3	To 130mm	6
To 90mm	4	To 165mm	7
To 120mm	5	To 8.0"	8

Figure 8
Union Carbide Mixing Section (UCC)

The original UCC or Maddock mixer, **Figure 8**, has alternate inlet and outlet channels separated on one side by a barrier with clearance designed to produce the required amount of shear and on the other side minimal clearance equal to the screw outside diameter. This results in material deposited on the inside wall of the extruder barrel by the barrier being wiped oft by the small-clearance land. So there are two distinct mixing actions.

The Egan mixer, **Figure 9**, has all flights with the same barrier clearance but the helical arrangement results in lower pressure drop because of the pumping action of the barrier flights.



Screw Diameter	# Of Channels
All Diameters To 8.0"	2

Figure 9
Egan Mixing Section

The helical UCC (Mattock) mixer, **Figure 10**, combines the virtues of the original UCC mixer and the Egan mixer although its mixing is a little less intensive than the original design. **Figure 11** shows a single-stage screw with a helical UCC mixer.

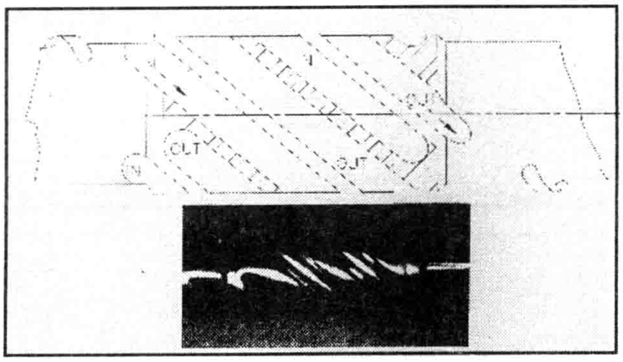


Figure 10

Helical Union Carbide Mixing Section (UCC-T)

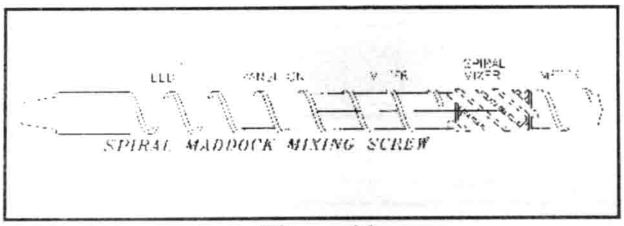


Figure 11

Single Stage Maddock Mixing (SSMM-T)

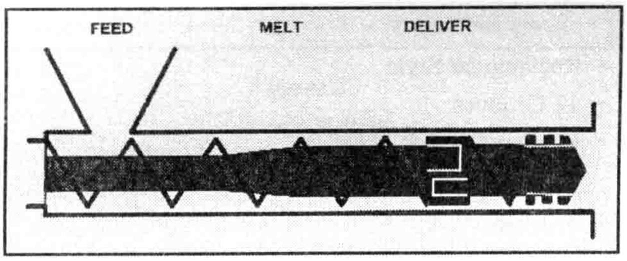


Figure 12

Dispersive and Distributive Mixing

In some applications, for example extrusion coating where high temperatures are not a problem, two mixing elements may be used to enhance the melt homogeneity.

A different type of mixer, which is both distributive and dispersive in its action is the RAPRA Cavity Transfer Mixer (CTM), depicted diagrammatically in **Figure 13**. It consists of a screw (**Figure 14**) and barrel extension, each containing hemispherical cavities. Mixing performance is good, specific energy consumption low and it purges well. The main drawback is cost.

BARRIER SCREW

The concept of the barrier screw, **Figure 15**, is not new. It was first applied to the extrusion of plastics by Mailiefer around 1980 but it took many years for this design to become widely accepted, partly through better analysis and

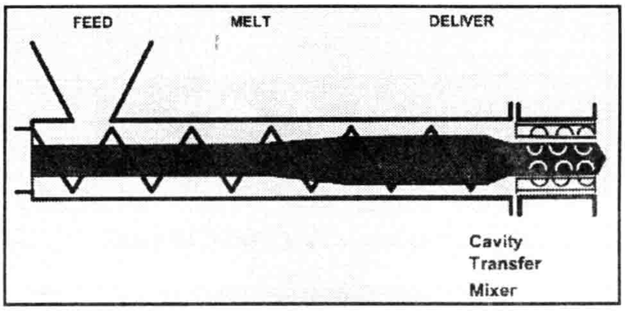


Figure 13

Cavity Transfer Mixer (Diagrammatically)

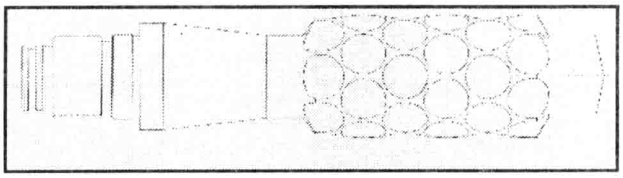


Figure 14

Cavity Transfer Mixer (CTM)

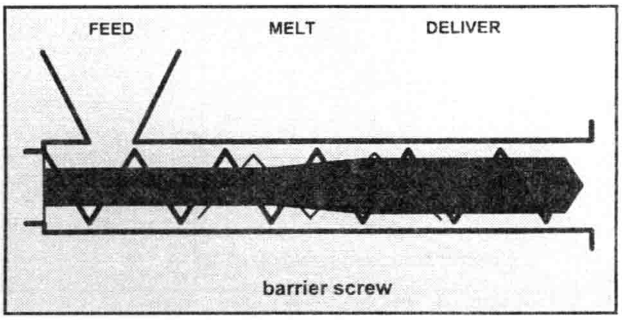


Figure 15

Barrier Screw

understanding of the process and partly through more versatile machine tools more recently available for the manufacture of the screws.

The purpose of the barrier screw is to create more efficient conditions for melting of the polymer by preventing early break-up of the solids bed, by trapping the solids between a barrier flight and the trailing edge of the main screw flight, at the same time allowing the melt to spill over the solids barrier or dam (**Figure 16**).

Clearly, the way the dimensions of the two channels in the barrier design vary and the clearance over the barrier are vital parameters in the design of barrier screws.

Questions such as whether it is better to vary the melt channel width or depth are often asked. The answer is always.. it depends on the polymer, it depends on the end product.

Theorists will argue that a barrier screw can only operate efficiently on one material at one screw speed and one barrel temperature. In practice this is not the case. Whilst some polymers require the screw to be quite closely matched to their melting behaviour, most polyolefines, for example, are much more accommodating. Properly designed barrier screw has higher out-put, better stability and for the most part is very flexible.

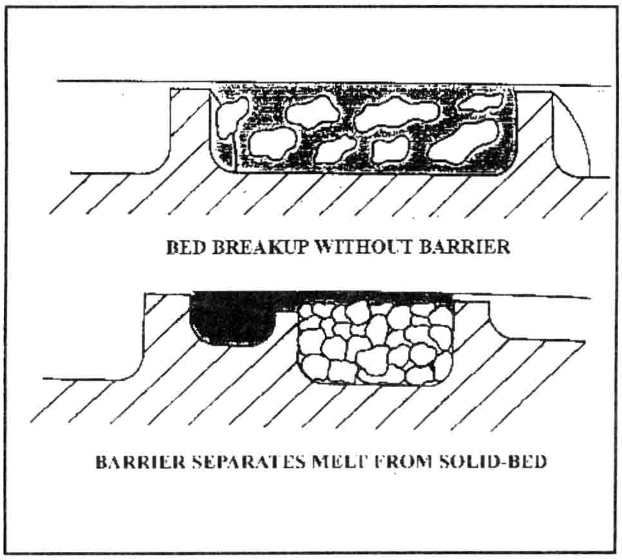


Figure 16
The Purpose of the Barrier Screw

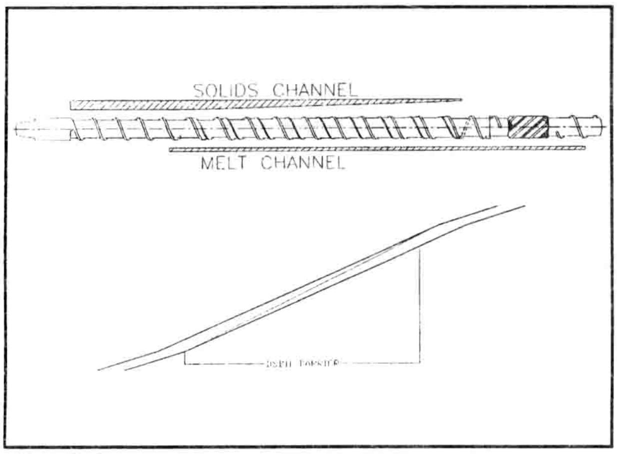


Figure 17
DSB II - Barrier Section

Figures 17 and **18** show two examples of different, single-stage barrier screws.

GROOVED FEED SYSTEM

Grooved feed extruders have been around in the plastics industry for almost 30 years although the basic idea is much older. The humble meat mincer is a testimony to that.

The extruder utilises an intensively cooled feed section with axial grooves over a length of 3-4 diameters of the screw as shown diagrammatically in **Figure 19**. The grooves are typically rectangular in section, full depth at the rear end tapering to nothing at the discharge end. **Figure 20** shows a typical arrangement for a 90 mm extruder.

On hard-pellet feed stock grooved system has the highest specific output with the lowest specific energy consumption. On materials which melt easily i.e. soft pellets or fluff, which may melt in the grooves, grooved-feed systems do not perform well.

High pressure in the feed zone of the extruder means that the screw and barrel in that area require special treatment to achieve acceptable life. Also grooved feed systems are less

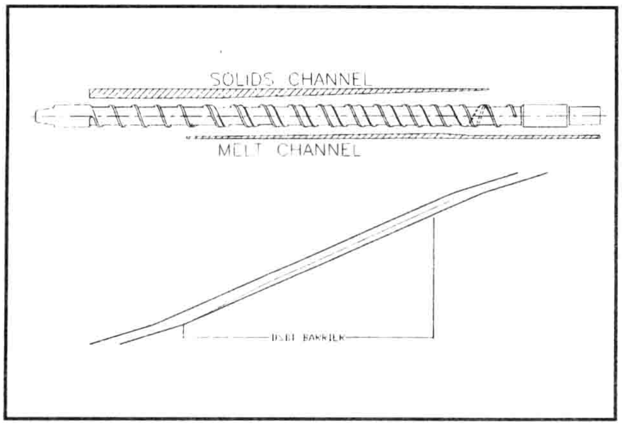


Figure 18
DSBI - Barrier Section

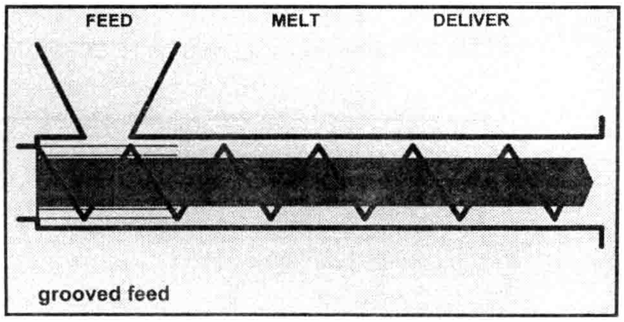


Figure 19
Groove Feed Systems

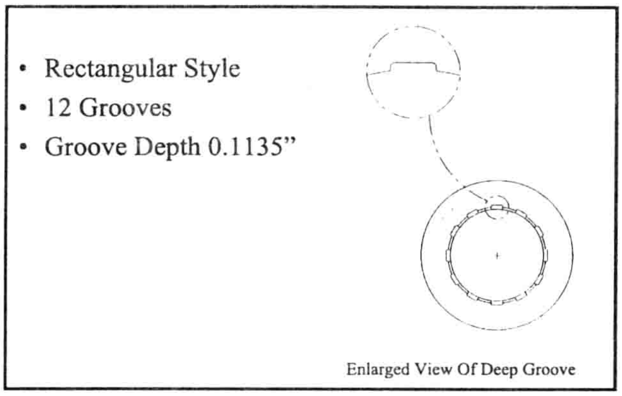


Figure 20
Design of Deep Grooves

suitable for the injection of liquid additives, such as PIB, for example, which, because of the high pressures at the fed end, have to be injected near the discharge end, thereby losing the benefit of mixing in the extruder.

On the other hand it would be un-thinkable to use anything other than a grooved-feed extruder for say HM-HDPE film.

Figure 21 shows a mini-groove design, which is an attempt to achieve a halfway house system between smooth and grooved system.