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**Heat recovery in tubular
heat exchangers for
medium viscosity food
products**

2001



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Research Association Group



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Heat recovery in tubular heat exchangers for medium viscosity food products

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SUMMARY

Recovery of heat in heat exchangers is presently only used in the food industry with low viscosity food products such as milk, fruit juices and beverages, where it is referred to as regeneration. In a tubular heat exchanger set up to recover heat, the hot 'processed' product is cooled down in the tubes by the cold 'unprocessed' product, which in turn is heated in the shell. This ensures that sterilised product flows through the tubes where the flow paths are well defined and cleanability issues are minimal.

In order for food processing companies to adopt this energy saving technology for foods of increased viscosity, this project aimed to demonstrate that heat recovery was economically viable. This was achieved using starch solutions of varying viscosity to represent the flow behaviour of food types that could be processed in tubular heat exchangers. Determination of the limitations on maximum viscosity were critical in that (a) below this viscosity heat recovery can be realised with existing commercial exchangers, and (b) above this viscosity re-design of the shell-side is required to prevent stagnation and poor distribution of flows. This limit was found to be at viscosities of approximately 20 cP (at 10 s shear rate), although starch solutions are highly shear thinning and the shear rate is critical for viscosity calculation.

The energy calculations showed that potential savings of between 50 and 70% could be achieved by utilising heat recovery over a wider range of foods than is currently commercially acceptable. The general rule was that higher savings were achieved with lower viscosity foods because of improved flow conditions and heat transfer. For example, for a food with a viscosity equivalent to a 2 wt% starch solution (e.g. a thin soup), the savings were of the order of 58 %. These equate to annual energy cost savings for a typical single factory of £6,707 on conversion of existing tubular heat exchangers to heat recovery and £16,288 on conversion of existing batch systems to tubular heat exchangers with heat recovery. Minimal capital costs are required for this conversion. Additional annual savings were also achieved in terms of costs for cooling water and effluent treatment of £8,043 and £5,712 respectively for a single factory.

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1. INTRODUCTION

The overall project objective was to demonstrate the application of heat recovery methods to medium viscosity foods such as fruit products, tomato products, cook-in-sauces, dressings and baby foods. This was achieved using various starch solutions of equivalent viscosities to the foods suitable for processing with tubular heat exchangers. By utilising recovered heat from the analogue food products, the project estimated improvements in the energy efficiency of tubular heat exchangers.

The disadvantages of pumping food products through the exchanger shell was that the flow paths were designed for fast flowing water and not for relatively slow flowing foods. This has implications for poor heat transfer and shear damage to the foods. Hence, a secondary objective was to assess the additional shear effects on product texture and consistency that may occur when foods are pumped through the shell-side. In addition, there were likely to be places in the exchanger shell where the flow could become too slow for effective heat transfer or could stagnate and be difficult to clean.

There is a need for tubular heat exchangers to be 'cleaned-in-place' between product changeovers, which was addressed by developing computational fluid dynamics (CFD) models to illustrate flows around tube supports and product entry/exit ports. This highlighted areas where the current shell design required modification before heat recovery could be adopted for a wide range of foods.

The project aimed to deliver information in a number of technical areas:

- Provision of rheological data taken at high temperatures that described the flow behaviour of the materials studied.
- Comparison of the effectiveness of heat recovery systems for medium and low viscosity foods, in order to determine the viscosity at which the rheology (flow behaviour) limits the recovered heat.

- Provision of data on the beneficial effects of food particulates on heat transfer efficiency.
- Derivation of heat transfer correlations of the form $Nu = b.Re^c.Pr^d$, which could be used for heat exchanger design.
- Information on the adverse processing effects (e.g. temperature and shear) on food product structure, when highly structured foods were passed through the exchanger shell.
- Quantification of the reduced energy requirements for heating media, and the savings in cooling water use in terms of water costs and effluent treatment costs.
- Data on the additional pumping requirements for the food products, in order to determine when the pressure drop limits the heat recovery process.
- CFD predictions of regions in the shell where flow can stagnate or become blocked.

The project was divided into phase 1 of 12 months duration and phase 2 of 6 months duration. The objectives of phase 1 were to prove the principle of heat recovery and to determine the operational limits for food viscosity where heat recovery was uneconomic. Phase 2 quantified the energy savings and reduced requirements for cooling water, in addition to providing the technical information that would allow heat recovery systems to be designed.

If the project was successful, the proposed total potential for energy savings to the year 2010 was estimated to be of the order of 680 TJ. This included two main changes to the current methods used by food companies:

- Existing users of tubular heat exchangers converted their systems to use heat recovery, which could save 10 TJ (£k 23),

- Companies that currently process in steam-jacketed vessels replaced these with energy efficient tubular heat exchangers that have heat recovery modules, which could save 670 TJ (£k 1,523).

1.1 Glossary of Terms

Apparent viscosity: The viscosity at a quoted shear rate. Foods with a high content of thickening agents will display a shear dependent viscosity. Units: Pa.s or cP.

Heat Transfer Coefficient (h): Heat transfer rate per unit area and per unit temperature difference between a solid surface and a fluid in contact with it. Units: $\text{W.m}^{-2}.\text{K}^{-1}$.

Hydraulic diameter: Four times the cross-sectional flow area divided by the wetted perimeter. Units: m.

Laminar Flow: The flow where fluid particles follow smooth streamlines and there is no obvious mixing between adjacent layers of fluid. It is also known as streamline flow.

Nusselt number (Nu): The ratio of convective heat transfer to heat transfer by conduction in the fluid. Higher Nusselt numbers represent effective heat transfer. Dimensionless.

$$\text{Nu} = \frac{h \cdot d}{k}$$

Where, h = heat transfer coefficient, $\text{W.m}^{-2}.\text{K}^{-1}$

d = hydraulic diameter, m

k = thermal conductivity, $\text{W.m}^{-1}.\text{K}^{-1}$

Prandtl number (Pr): The ratio of momentum diffusivity to the thermal diffusivity, and represents the ratio of heat transfer to fluid motion. Dimensionless.

$$\text{Pr} = \frac{c_p \cdot \eta_a}{k}$$

Where, c_p = specific heat of liquid, $\text{J.kg}^{-1}.\text{K}^{-1}$

η_a = apparent viscosity, Pa.s

Reynolds number (Re): The ratio between the fluid's dynamic forces and viscous drag forces. Values of the Reynolds number indicate the flow regime, whether the flow can be described as laminar, transitional, or turbulent. Re below 2,100 can be taken as laminar. Dimensionless.

$$\text{Re} = \frac{\rho.d.v}{\eta_a}$$

Where, v = velocity, m.s^{-1}

ρ = density, kg.m^{-3}

Rheology: The science of deformation and flow of matter.

Turbulent Flow: An irregular unsteady well mixed flow that leads to high rates of heat transfer.

2. BACKGROUND

2.1 Food Processing using Tubular Heat Exchangers

Tubular heat exchangers are widely used in the food industry for thermal treatment of foods as diverse as milk to tomato paste¹. The flowing viscosity is the key factor in the selection of the appropriate tube designs in order to maximise heat transfer and minimise pressure drops. Tetra Pak were the equipment manufacturer involved in this project, hence the heat exchangers used were from the Tetra SpiraFlo range. These used hot water as the heating medium and cold water as the cooling medium. Water was preferred to steam because it offered a less severe means of transferring heat from the medium to the product, and minimised undesired reactions such as fouling and thermal damage to the flowing food. The hot water was generated with a plate heat exchanger and was pumped through the exchanger shell in a counter current direction to the food flowing in the tubes.

Figure 1 illustrates the commercial SpiraFlo tube configurations in cross-section for the concentric, mono- and multi-tube exchangers. The tubes were inserted into a 70 mm diameter shell that allowed the heating or cooling media to flow around the tubes (see figure 2). The velocity of the medium was designed to be turbulent in order to maximise heat transfer. All of the SpiraFlo exchanger materials were 316 stainless steel.

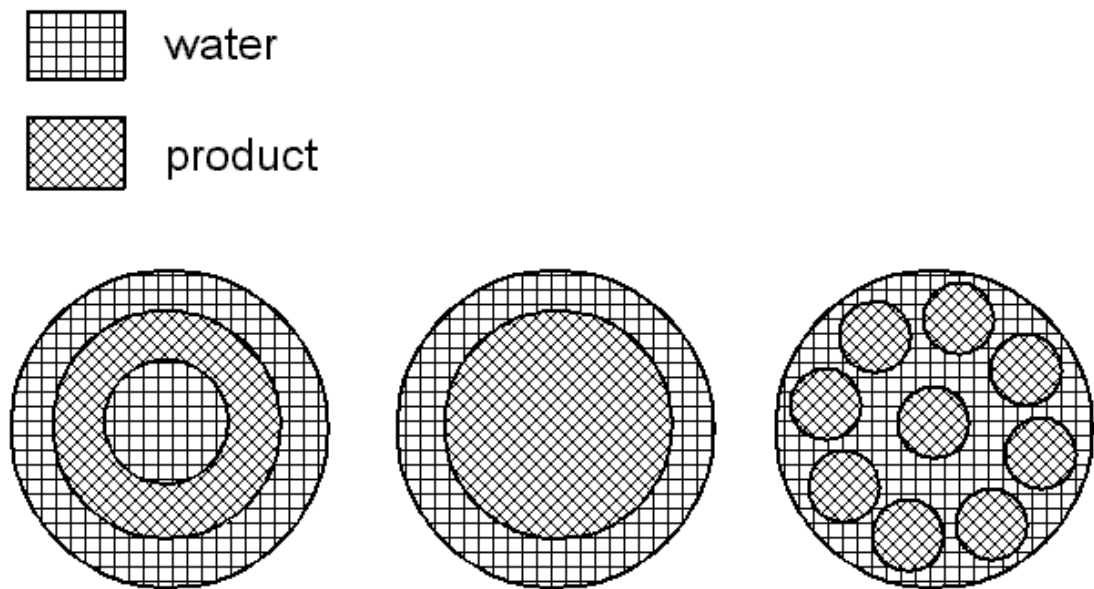


Figure 1: Illustration of the tube configurations in cross section for concentric, mono- and multi-tube exchangers. (Diagram courtesy of Tetra Pak)

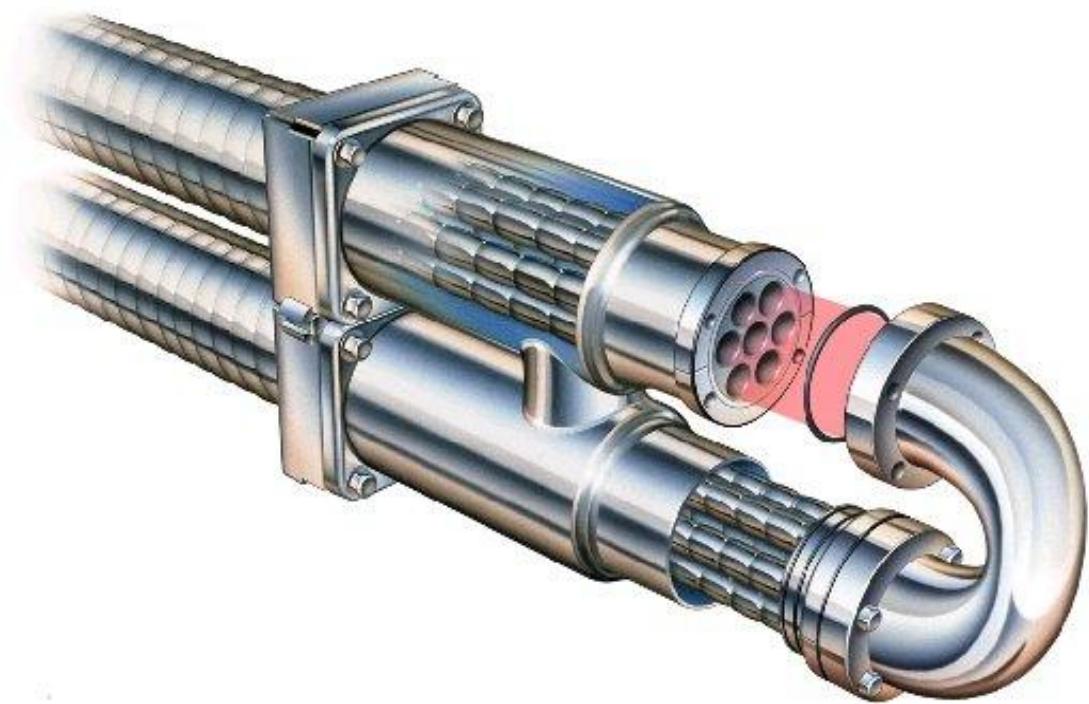


Figure 2: Drawing of the multi-tubes inside the shell, illustrating the spiral surface patterns that promote turbulence in the flow. (Diagram courtesy of Tetra Pak)

2.2 Heat Recovery Principles

Heat recovery, or regeneration, is used in the food industry for heating and cooling of low viscosity products such as milk, fruit juices and beverages. In a tubular heat exchange system set up to recover heat, the hot 'processed' product is cooled down in the tube inserts by the cold 'unprocessed' product, which in turn is heated in the surrounding shell (see figure 3). The product passes counter current through the tube inserts to the product passing through the surrounding shell.

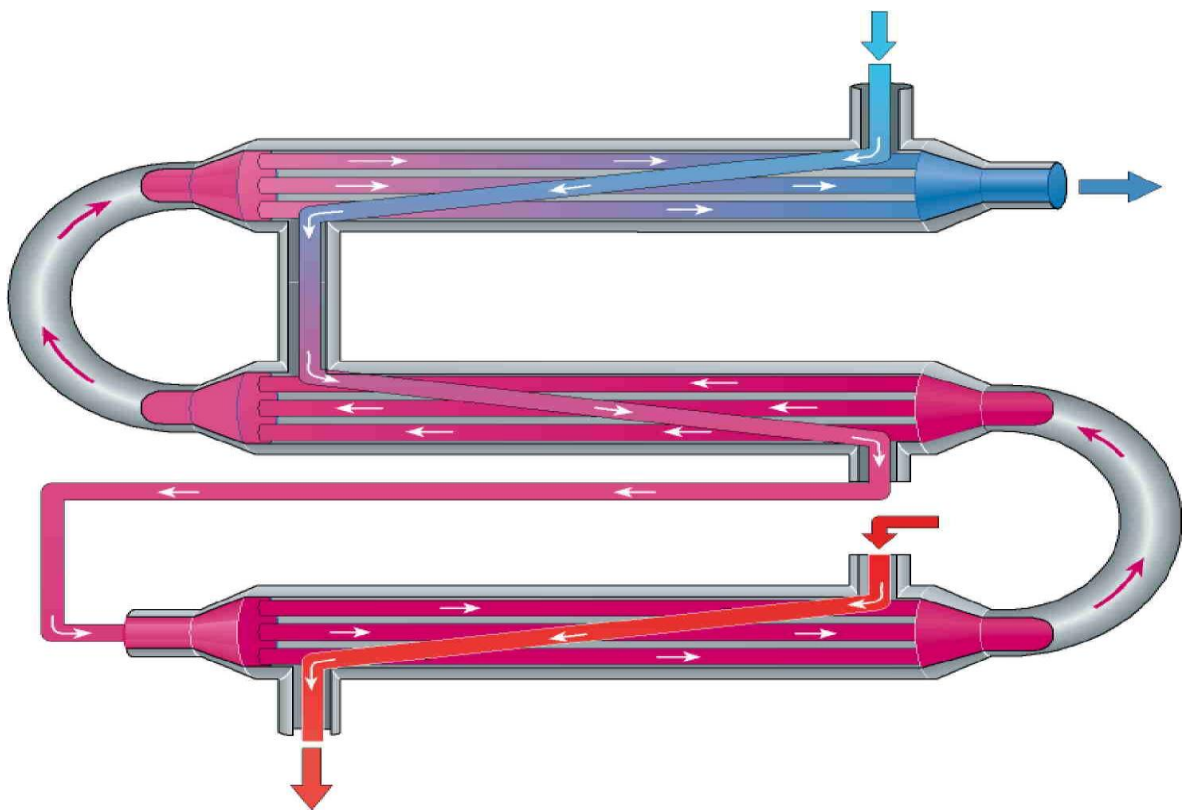


Figure 3: Conventional tube configuration for heat recovery with liquids of low viscosity. The shading represents the temperature of the various product and media flows. (Diagram courtesy of Tetra Pak)

Apart from the heat recovery systems used for milk, fruit juices and beverages, little work has been published on foods of higher viscosities. There is one report on ghee (clarified butter) production² that used heat recovery, but at the high temperatures associated with this pasteurisation operation (85-100°C) the ghee viscosity was reduced and flow conditions were close to the transitional region or even turbulent. Ghee has a simple molecular structure that does not require such a careful treatment as the foods studied within this project. The technical issues addressed included the need to avoid damage to the delicate macromolecular structures that gives rise to the 'consistency' of finished food products.

By removing heat from a hot product via exchange with the cold feed, the number of cooling and/or heating tubes can be reduced. Moreover, the heating, as well as the cooling energy consumption, will be significantly reduced, making the exchanger more compact and energy efficient. The hurdle to overcome in the project was that laminar flow of hot product was dominant in the shell flows ($10 < Re < 500$) compared with turbulent flow of the heating or cooling media. This created adverse conditions; for example it:

- limited the heat transfer from the medium (hot food product) to the food (cold food product),
- raised the pressure drop through the exchanger,
- created the potential for low velocity regions in the shell, leading to flow stagnation,
- may have incurred adverse shear damage to products with delicate structures.

3. EXPERIMENTAL METHODS

Colflo 67 starch (National Starch) was used as the test solution in the Spiraflo experiments because it is one of the most commonly used thickeners in heat processed foods. Starch concentrations from 0 to 5 wt% provided the data to evaluate heat recovery over a wide range of solution viscosities, representing food viscosities from fruit juice to custard.

Two systems were set up to compare energy and cooling water savings both with and without heat recovery. Eight Spiraflo tube passes were available for the experiments. Only 4 and 5 wt% starches were evaluated using the 34 mm mono-tubes because these tubes were designed for higher viscosity foods, whereas the range from 0 to 5 wt% was used for the 7 x 16 mm multi-tubes and concentric tubes. For each wt%, two flowrates and two temperatures (70 and 90 °C) were evaluated in order to generate data that gave four 'sets' of points for plotting heat transfer correlation data.

System 1 without heat recovery used four water-heated and four water-cooled tubes (see figure 4), and system 2 with heat recovery used four tubes where the heat transfer was product-product, two water-heated and two water-cooled tubes (see figure 5). With the concentric tubes, all four tubes were used in product-product heat transfer, and the product was heated to the final target temperature of 70 or 90 °C using multi-tubes. The section of interest to the project comprised the heat recovery tubes.

Various improvements were made to the exchanger and its instrumentation during phase 1 of the project to improve the accuracy of the measurements. This included the use of a large insulating cover to minimise heat loss to the atmosphere.

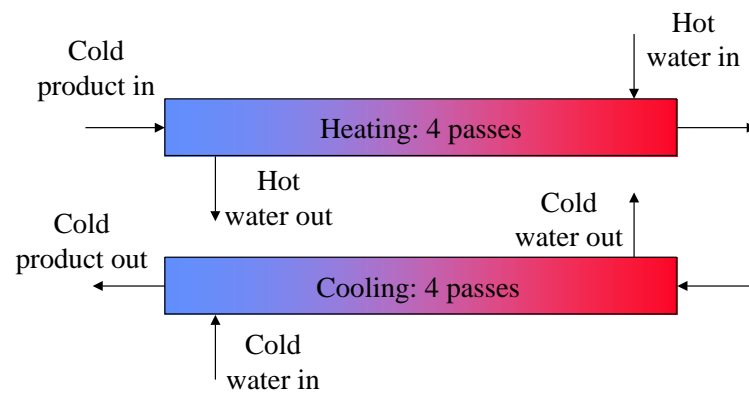


Figure 4: Existing exchanger set-up with 4 x heating and 4 x cooling tube passes

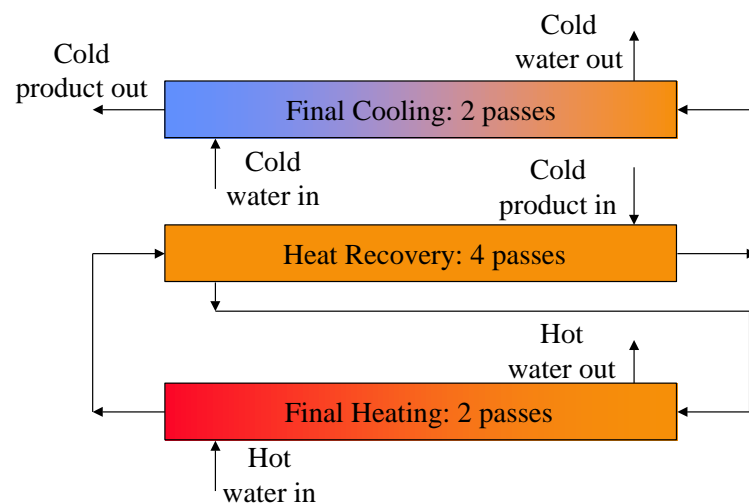


Figure 5: Heat recovery set-up with 4 x heat recovery, 2 x final heating and 2 x cooling tube passes

4. RESULTS AND DISCUSSIONS OF PHASE 1 TESTS

4.1 Heat Transfer Efficiency and Equilibrium Time

The time for the exchanger to reach thermal equilibrium was one of the measures used to evaluate the thermal efficiency with product-product heat transfer. At thermal equilibrium, these temperature differences (ΔT) will be the same. It would be expected that the equilibrium time would be lower for the 'thin' liquids because heat transfer is more effective in turbulent flow compared with laminar flow³. Figures 6 and 7 illustrate this for 1 and 5 wt% starch respectively, using the temperature difference between the inlet and outlet of both tubes and shells as the criteria. The flow conditions of 1 wt % starch were turbulent and those of 5 wt % starch were laminar. These experiments were conducted in the multi-tubes.

The experiments with 4 and 5 wt% starches in the 34 mm mono-tubes, and to a lesser extent in the concentric tubes, required unacceptable times (> 15 minutes) to reach thermal equilibrium. However, the percentage of energy transferred within the heat recovery section was still close to 40 % of the total transferred, once thermal equilibrium had been established. It was concluded that while heat recovery with 34 mm mono-tubes could provide energy advantages, the restriction of excessive times to reach thermal equilibrium was such a disadvantage that this could not be applied in industry.

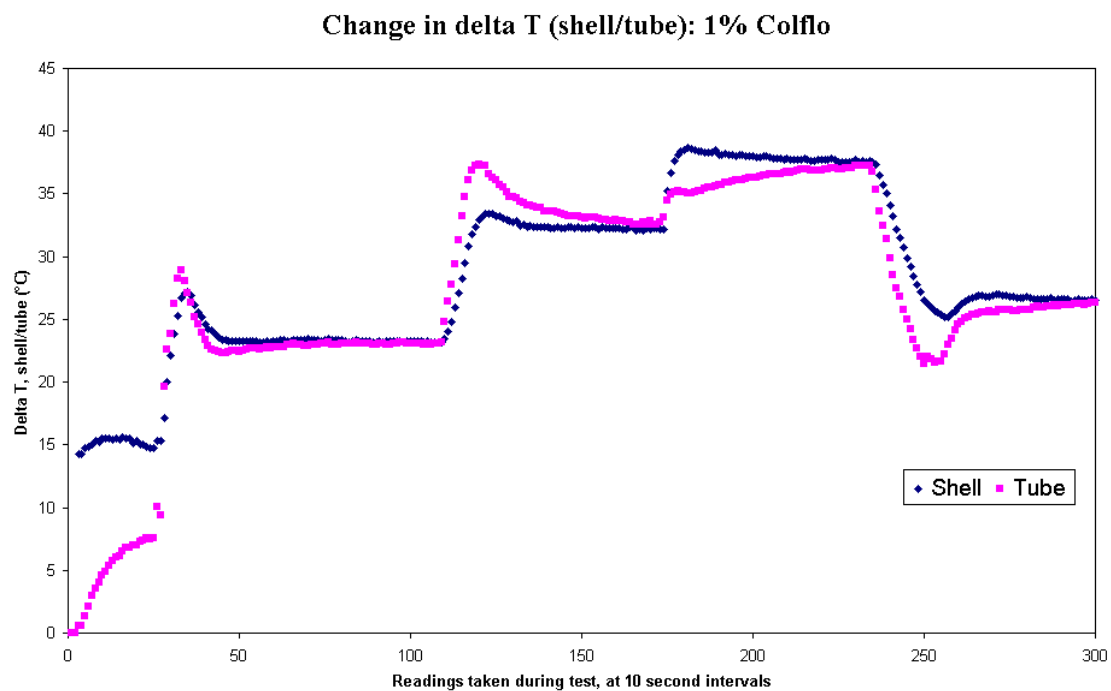


Figure 6: Time to reach thermal equilibrium for 1 wt % Colflo 67 in the multi-tubes

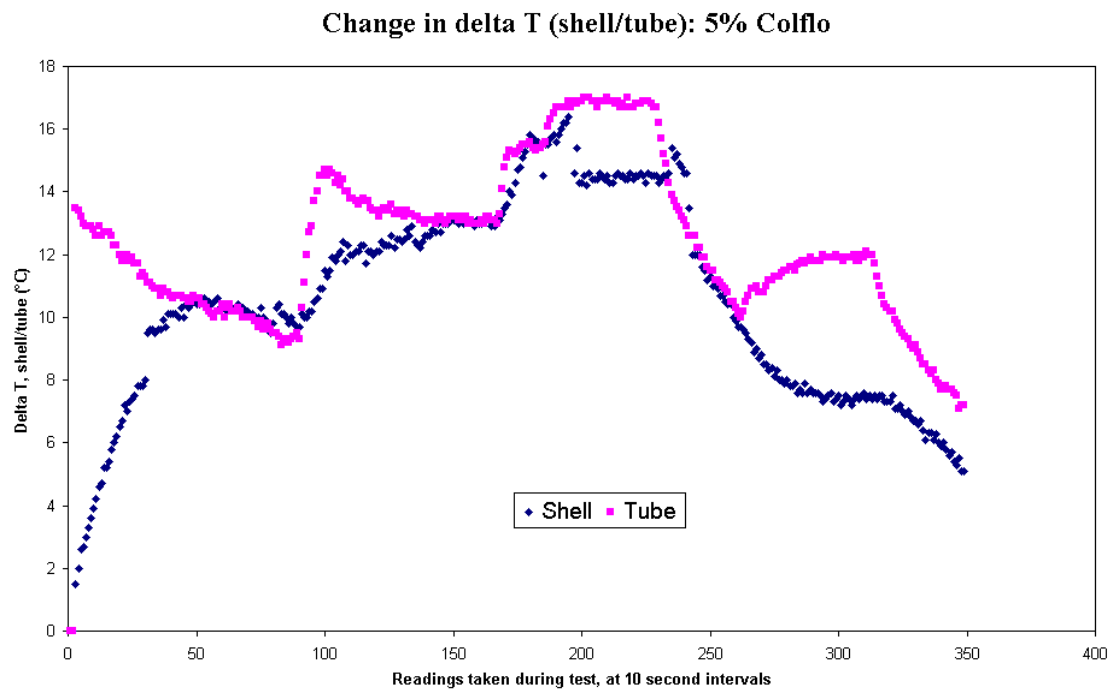


Figure 7: Time to reach thermal equilibrium for 5 wt % Colflo 67 in the multi-tubes

4.2 Effects on Heat Transfer of Small Particulates

Heat recovery with 34 mm mono-tubes showed that insufficient heat transfer surface existed to make this viable, therefore particle experiments were not relevant. The more energy efficient 7 x 16 mm multi-tubes did not allow particles > 2 mm to flow around the tight-packed tube bundle on the shell-side, and this prevented experiments from being conducted. However, reports from literature showed enhancement of heat transfer of the order of 20% due to the presence of particulates in tube-side flow and it would be reasonable to expect similar benefits in shell-side flow³.

4.3 Effects of Heat Recovery on Food Product Structure

Tomato paste, white sauce and apple sauce were provided by Heinz and Hazlewood, and these products were evaluated with the multi-tube exchanger set up for product-medium and for product-product heat transfer (see figures 4 and 5). The viscosities of tomato paste and apple sauce were found to be too high to achieve adequate heat transfer, and for tomato paste the pressure drop was excessive. Economic heat was recovered for a white sauce and tomato ketchup made from paste.

Some degradation in the product structure for the white sauce was measured at 20°C and the power law coefficients are given in Table 1 to illustrate the viscosity reduction. At least 4 hours was allowed for the structure to re-develop before measuring viscosities. For example, viscosities at a shear rate of 10 s^{-1} were 3,161 cP before processing compared with 2,298 cP after processing. Therefore, some permanent structure degradation had occurred. To achieve thermal equilibrium required re-circulation of sauce, and therefore there was a greater chance of structure breakdown caused by shear in the shell-side.

Table 1: Power law parameters for a white sauce used for heat recovery experiments in multi-tubes. Pre- and post-process refer to the heat recovery experiments with four product-product tube passes.

White sauce description	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
Pre-process at 21 °C	16.4	0.29	3,161
Post-process at 20 °C	12.0	0.28	2,298

The tomato ketchup experiments showed interesting results in heat recovery, with an initial viscosity increase after 15 minutes as the starch gelled in the exchanger. The formulation included starch addition as a powder. However, on re-circulation for 30 and 45 minutes the viscosity gradually reduced, until after 60 minutes an increase was measured. All the viscosity tests were done on samples retained until the following day in order to allow temporary structures to re-build. Table 2 shows the viscosity data. It was suspected that the increased shearing after 60 minutes re-circulation had caused the tomato cells to partially break in a similar way to the effect of an homogeniser, resulting in increased viscosity. Thus, this effect could be beneficial.

Table 2: Power law parameters for tomato ketchup used for heat recovery experiments in multi-tubes.

Time of re-circulation (minutes)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
0	1.25	0.44	345
15	3.49	0.46	1,017
30	2.71	0.45	762
45	2.13	0.46	613
60	2.84	0.46	812

5. RESULTS AND DISCUSSIONS OF PHASE 2 TESTS

Phase 1 had proven that there was economic potential for using heat recovery with the multi-tubes and concentric tubes, but the 34 mm mono-tubes did not have sufficient heat transfer surface. This latter finding excluded the experiments with particulates.

5.1 Energy Savings

Energy calculations were done for all products processed in the multi- and concentric tubes. For example, for 4 wt% starch flowing at 27 l.min^{-1} in the multi-tubes, the energy transferred in the product-product (heat recovery) section was 27 kJ.s^{-1} or 45 % of the total of 60 kJ.s^{-1} . For 5 wt% starch in the multi-tubes, the energy transferred in the product-product sections was just under 40 %. Figures 8 and 9 present the percent energy transferred for the multi-tube experiments, as a function of wt% Colflo (figure 8) and of log Colflo viscosity (figure 9). Figures 10 and 11 present the percent energy transferred for the concentric tube experiments, as a function of wt% Colflo (figure 10) and of log Colflo viscosity (figure 11).

The conclusions on energy transfer in product-product heating were that significant quantities of energy could be saved with multi-tubes and concentric tubes. The limitation on heat transfer with the starch solutions of increasing viscosity were in the time required to reach thermal equilibrium, observed markedly with 4 and 5 wt% solutions. These would require re-circulation of product for 15-30 minutes before thermal equilibrium was reached, which may have implications for structural damage in formulated products. Realistically, Colflo solutions of 3.0 wt% and above (equivalent viscosity of 200 cP at a shear rate of 10 s^{-1}) were not suited to heat recovery without re-design of the shell-side flows.

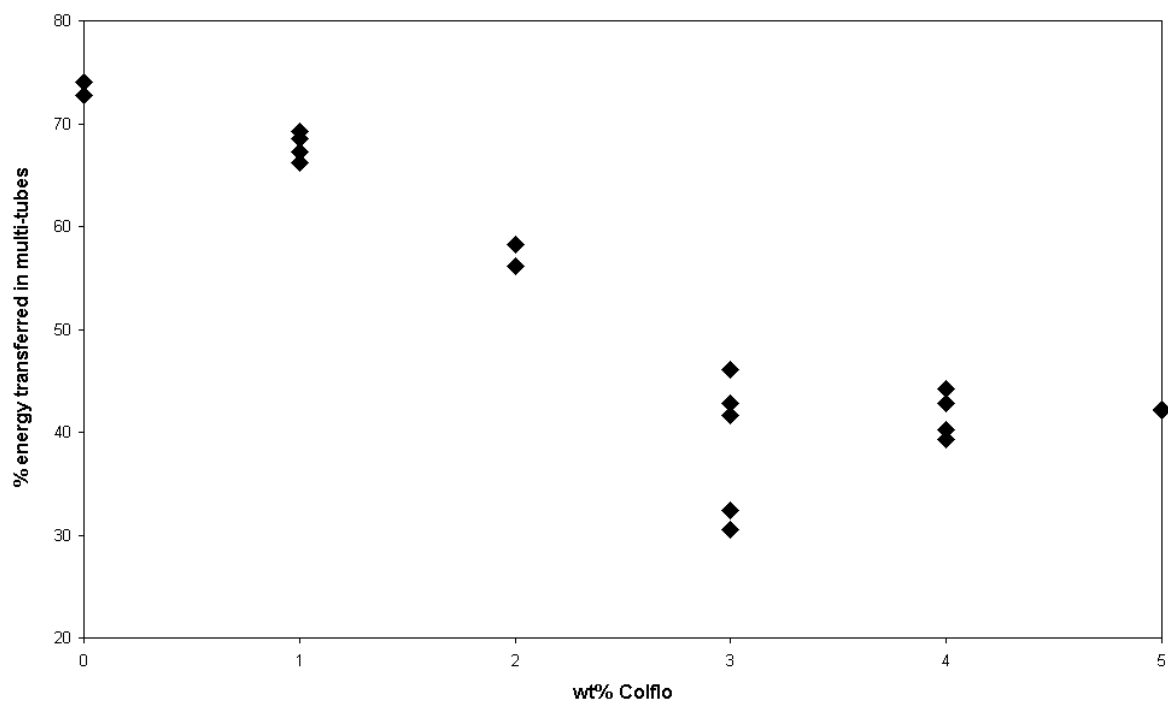


Figure 8: Percentage energy transferred in the four product-product multi-tubes as a percentage of the total transferred, plotted against wt%.

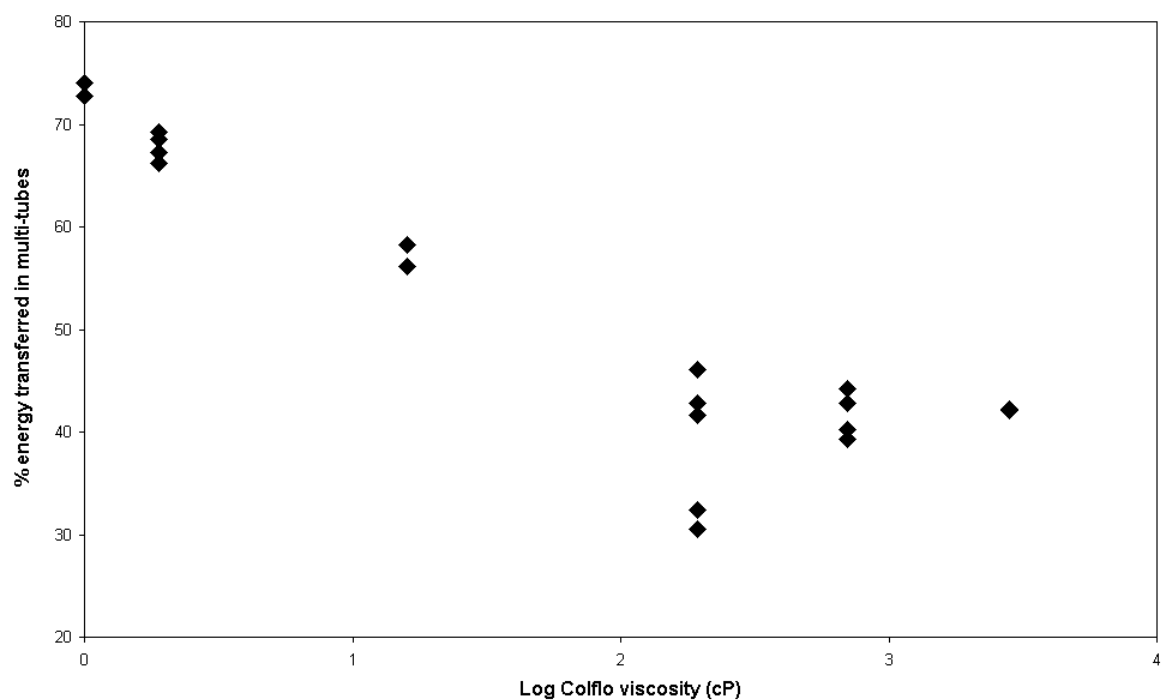


Figure 9: Percentage energy transferred in the four product-product multi-tubes as a percentage of the total transferred, plotted against log viscosity.

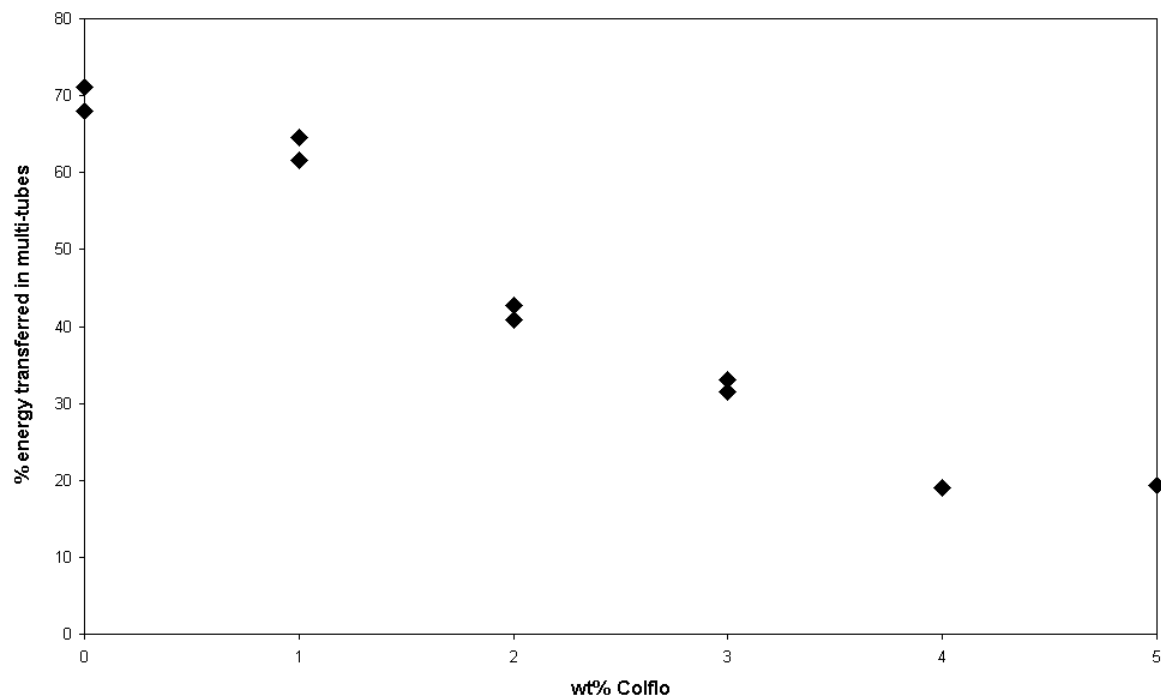


Figure 10: Percentage energy transferred in the four product-product concentric tubes as a percentage of the total transferred, plotted against wt%.

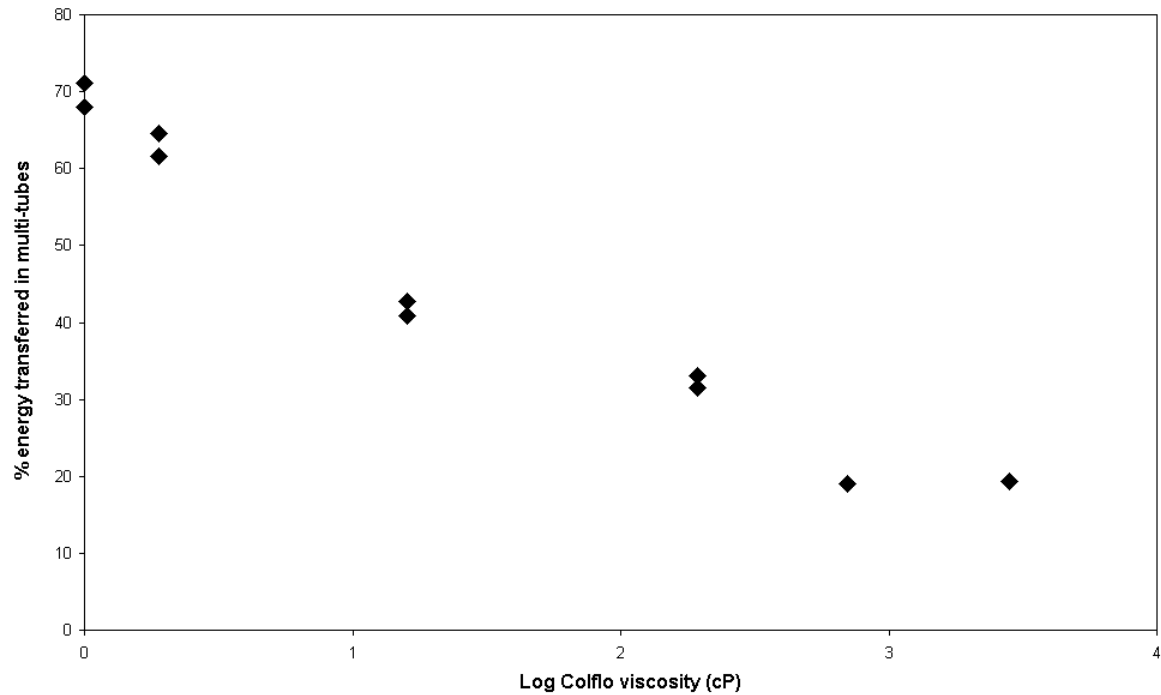


Figure 11: Percentage energy transferred in the four product-product concentric tubes as a percentage of the total transferred, plotted against log viscosity.

5.2 Cooling Water Savings

Taking the same example of 4 wt% starch flowing at 27 l.min^{-1} in the multi-tubes, to cool the product from 67 to 20 °C without heat recovery required a cooling water flowrate of $6,110 \text{ kg.h}^{-1}$ compared with $3,000 \text{ kg.h}^{-1}$ with heat recovery. The calculations assumed that the cooling water entered at 8 °C and exited at 20.4 °C. Thus, a reduced cooling water flowrate was required when extra cooling from the product-product heat transfer was taken into account.

Table 3 gives the measured cooling water flowrates required for 1 to 5 wt% Colflo solutions. The data presented without heat recovery was calculated from the Quantum program (Tetra Pak) in order that the cooling water exit temperatures were the same as those used in the heat recovery experiments.

Table 3: Cooling water (CW) flowrates required to achieve equivalent starch cooling duty for 1 to 5 wt% Colflo solutions in the 7x16 multi-tubes. Data with heat recovery (HR) was measured experimentally, and data without heat recovery was calculated from the Quantum program.

Wt% Colflo	Product flowrate (l/min)	CW temperature rise (°C)	Product temperature fall (°C)	Measured CW flowrate, with HR (kg/h)	Calculated CW flowrate, without HR (kg/h)
1	126	8.8 to 37.5	80.6 to 34.3	2,268	11,834
2	98	8.8 to 36.3	61.6 to 37.0	2,268	5,205
3	107	8.8 to 26.5	50.2 to 42.3	2,268	2,852
4	27	8.0 to 20.4	67.0 to 20.2	3,000	6,110
5	32	9.1 to 18.4	58.2 to 30.9	3,156	5,581

5.3 Heat Transfer Correlations

There was a marked change in the heat recovery efficiency as the flow regime in the exchanger shell changed from turbulent (0–2 wt% starch) to laminar (3–5 wt% starch). This was illustrated by the values for average heat transfer coefficients in Table 4. These coefficients were used as relative numbers to compare the thermal efficiencies.

The information required to calculate the number of tube passes for an industrial installation can be obtained from the plots of these coefficients as correlations of dimensionless numbers. The correlation plots (see figures 12 and 13) also highlight the two distinct regions of laminar and turbulent flow, with Tables 5 and 6 presenting the numerical values of the shell-side correlations for the multi- and concentric tubes respectively.

Table 4: Heat transfer coefficients calculated for flow of the starch solutions in the multi-tube shell.

Wt% Colflo 67	Heat transfer coefficient, $\text{W.m}^{-2}.\text{K}^{-1}$	Flow Regime
0	6,000	Turbulent
1	6,000	Turbulent
2	620	Laminar
3	610	Laminar
4	580	Laminar
5	410	Laminar

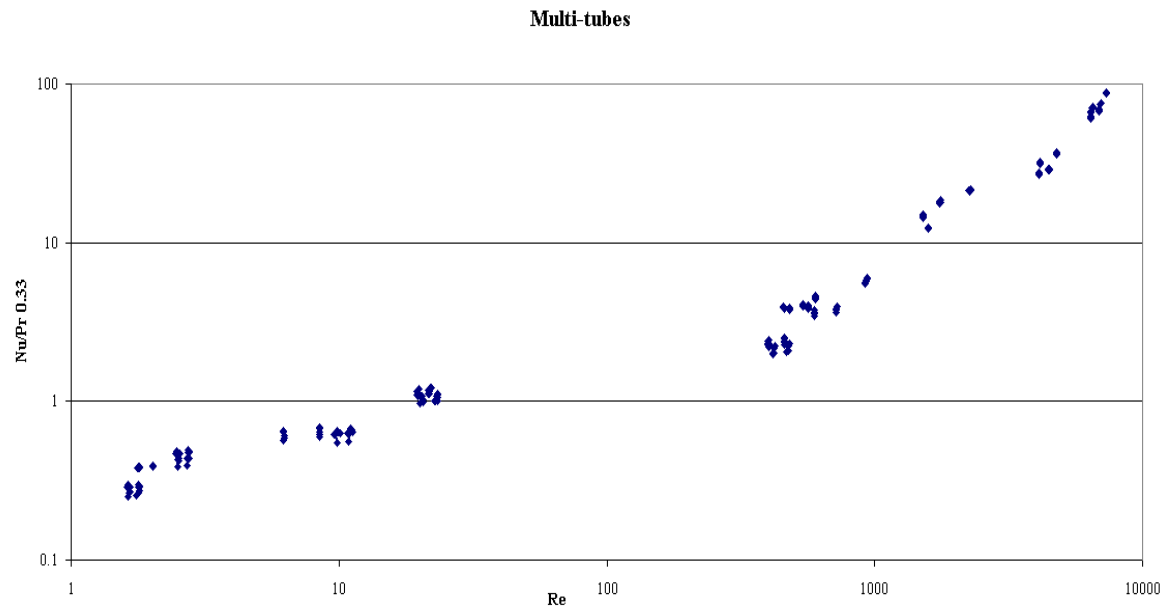


Figure 12: Heat transfer correlations for the starch solutions and foods flowing in the shell of a 7x16 multi-tube exchanger

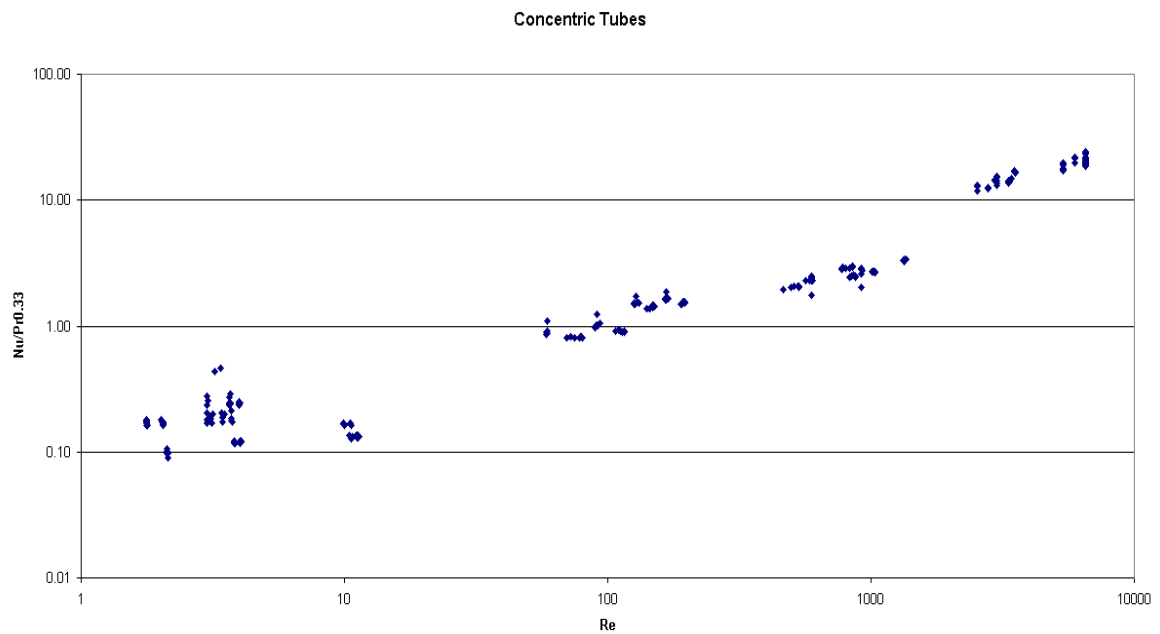


Figure 13: Heat transfer correlations for the starch solutions and foods flowing in the shell of a 7x16 concentric tube exchanger

Table 5: Shell-side correlation data for the 7x16 multi-tubes, as a function of the Reynolds (Re), Nusselt (Nu) and Prandtl (Pr) numbers.

Reynolds number range (shell)	Shell-side correlation
$1 < Re < 575$	$Nu/Pr^{0.33} = 0.288 Re^{0.384}$
$575 < Re < 7,500$	$Nu/Pr^{0.33} = 0.0987 Re^{0.688}$

Table 6: Shell-side correlation data for the concentric tubes, as a function of the Reynolds (Re), Nusselt (Nu) and Prandtl (Pr) numbers.

Reynolds number range (shell)	Shell-side correlation
$1 < Re < 450$	$Nu/Pr^{0.33} = 0.081 Re^{0.542}$
$450 < Re < 7,000$	$Nu/Pr^{0.33} = 0.0031 Re^{1.025}$

5.4 Additional Pumping Requirements

In a heat recovery system, the flow path for the food will be increased in length with the likely result that the pressure drop will also increase. Precise comparisons of pressure drop were difficult to test experimentally because of the number of variables that needed to be controlled. Therefore, the Quantum program was used to estimate pressure drops where experimental data could not be provided. Quantum had proven accuracy for pressure drop prediction in tube flows but did not contain data for flows in the shells. Hence, the approach taken was to use pressure drop data from the heat recovery experiments and to predict the corresponding pressure drops for the closest equivalent systems without heat recovery.

Table 7 presents the pressure drop across the complete heat exchanger systems, comparing the heat recovery set-up (as described previously) with that required if no heat recovery was used. Exact temperature and flowrate conditions could not be matched for the heating and cooling water, and as a result this table should be used as an approximate guide. The

conclusions were that there was an increase in pressure drop as a result of the food flowing through an extended and more complex pathway, but that this was not excessive.

Table 7: Pressure drop across a multi-tube heat exchanger, operating with and without heat recovery.

Wt% Colflo	Pressure drop with heat recovery (kPa)	Pressure drop without heat recovery (kPa)
1	188	105
2	216	159
3	1,908	1,852
4	467	332
5	1,290	1,070

5.5 Computational Fluid Dynamics (CFD)

The 7x16 multi-tube exchanger was used for CFD simulations, since this proved to be the most advantageous in terms of the energy transferred by product-product heat transfer. Regions of slow flow between the seven tubes were identified with 3 wt% Colflo starch (see Figure 14) as the solution entered the shell. A coarse grid was used to speed up the run time of the computations but the disadvantage of this was in the less than ideal definition for flow within the tube bundle. The distribution of flow in the shell around the seven parallel tubes was satisfactory for 0, 1 and 2 wt% starches but it was possible to identify regions of poor distribution and reduced velocity as the solution viscosity increased from 3 wt% to 5 wt%. To overcome poor flow distribution within the tube bundle the tube spacing should be increased or the number of tubes reduced from 7 to 5.

Figure 15 highlights the situation for 3 wt% starch by simulating axial velocities in a series of cross-sectional contour maps. The first picture was upstream of the inlet starch and showed zero velocity at this end of the shell. The second picture was also upstream but

closer to the inlet, and showed higher velocities towards the top of the shell. The inlet pipe was shown in the third picture and it required two further cross-sectional pictures before the flow stabilised. The zone around the central tube never flowed at a velocity adequate to induce good heat transfer, therefore the hot product flowing inside this tube would not give up much of its heat. The outside tube surfaces would remain hot and provide the conditions for surface fouling to occur.

Additional areas for concern were the entry manifolds for the product as it flowed into the shell. The present single entry port gave rise to non-uniformity in velocity from top to bottom of the shell. This could be resolved by increasing the number of entry ports. The implications for a food with shear dependent viscosity arise in regions where the velocity is reduced and as a result the food viscosity increases. This increases the drag forces and acts to further reduce the velocity, with the ultimate effect of producing stagnation zones. In the context of a food operation, such zones must be avoided because they will be difficult to clean and promote bacterial growth.

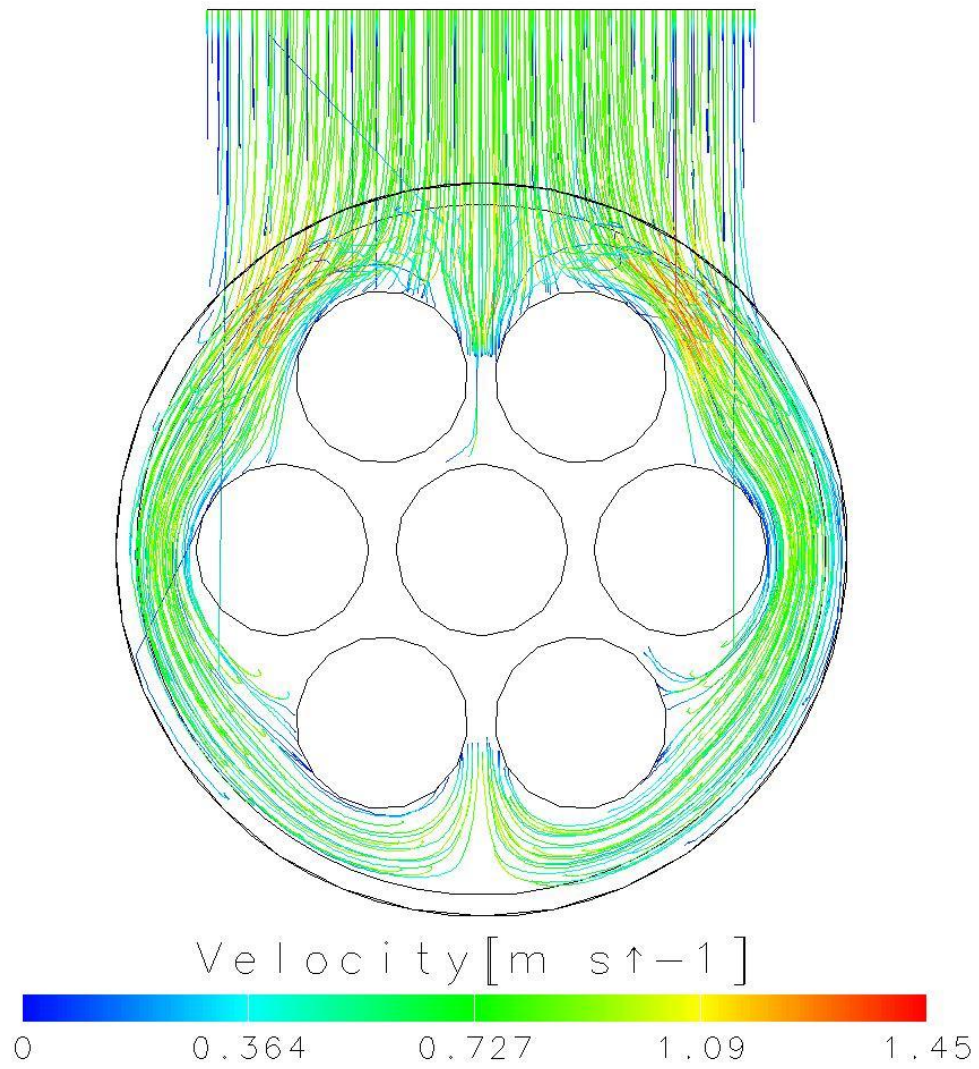


Figure 14: Streamlines for 3 wt% Colflo entering the shell of a 7x16 multi-tube exchanger.

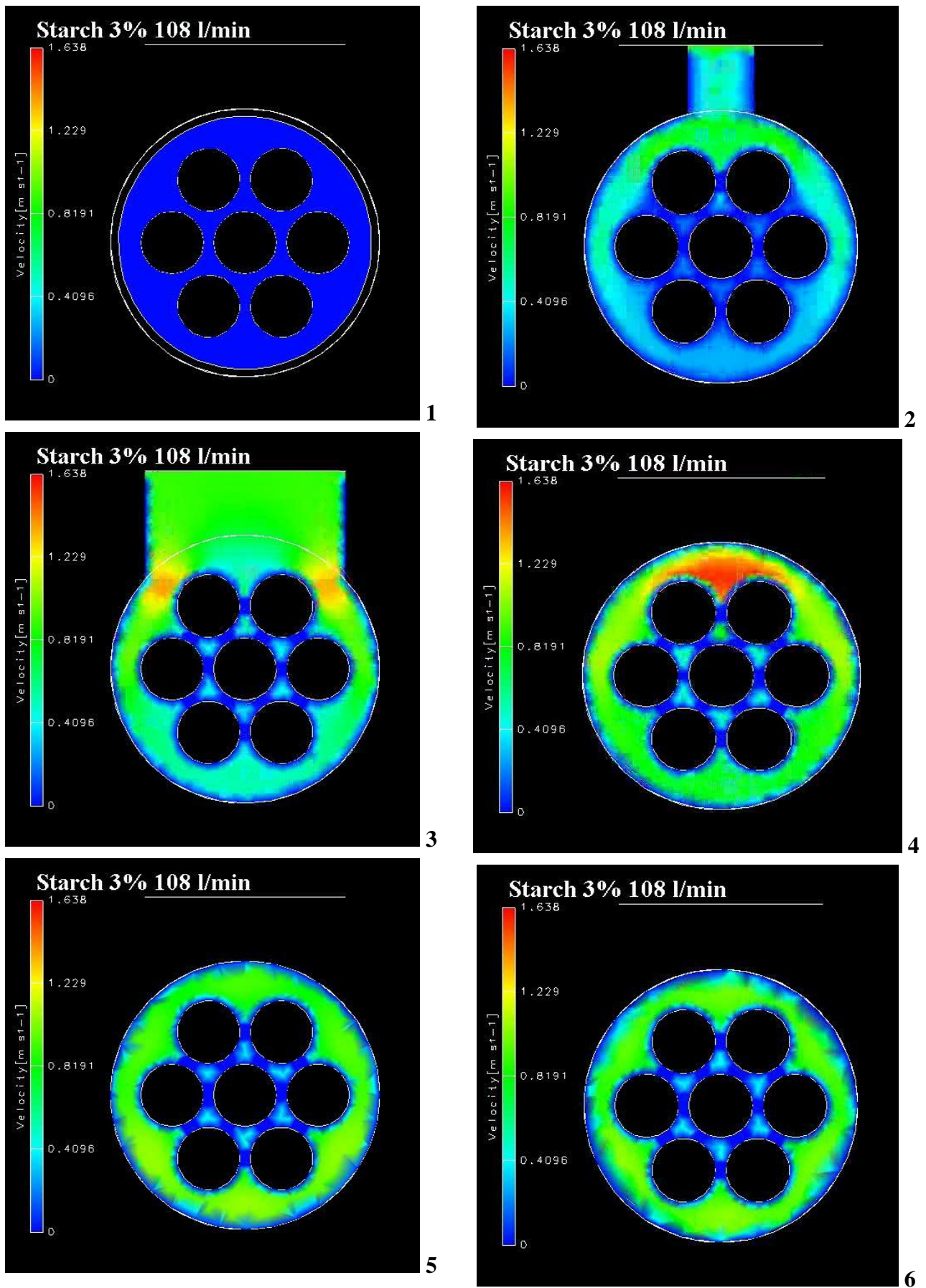


Figure 15: Axial velocities for 3 wt% starch at 108 l/min, displayed as cross-sections along a multi-tube exchanger from prior to the inlet (1) until the flow stabilises (6).

6. COST BENEFIT ANALYSIS

6.1 Energy Savings

The calculations of energy and cost savings were made for a factory that processed 40,000kg of an equivalent food product (to 2 wt% Colflo starch) per day and operated for 320 days per year. Tables 8 and 9 give the potential annual savings for the single factory and also scaled to 2010 if four factories converted each year. Table 8 is for the conversion of existing tubular heat exchangers to heat recovery, and Table 9 for conversion of existing batch process vessels to tubular heat exchangers with heat recovery. Batch vessels are the more common style of process operation used in the food industry and represent the most significant savings because they are less energy efficient than heat exchangers.

Table 8: Potential annual savings in energy by converting existing tubular heat exchangers to heat recovery (no additional capital costs are required).

Converting existing tubular exchangers to heat recovery	2001, single factory	2010, four factories p.a.
Energy use (TJ)	4.79	172.47
Energy use with heat recovery (TJ)	2.02	72.62
Energy savings (TJ)	2.77	99.85
Energy savings (£) *	6,707	241,439

* Calculated from 4th quarter 1999 figures for an average gas price of 0.531p/kWh (DUKES, DTI web-site) and converted to steam bill figures by assuming 61 % boiler efficiency (ECG 67).

Table 9: Potential annual savings in heating energy by converting existing batch vessels to energy efficient tubular heat exchangers with heat recovery. Additional capital costs are required.

Converting existing batch systems to tubular heat exchangers with heat recovery	2001, single factory	2010, four factories p.a.
Energy use (TJ)	11.63	418.85
Energy use with heat recovery (TJ)	4.90	176.36
Energy savings (TJ)	6.74	242.49
Energy savings (£) *	16,288	586,351

* Calculated from 4th quarter 1999 figures for an average gas price of 0.531p/kWh (DUKES, DTI web-site) and converted to steam bill figures by assuming 61 % boiler efficiency (ECG 67).

The total potential energy saving to 2010 was estimated to be 342 TJ or £ 826k for the example chosen. These figures compared with 680 TJ and £ 1,546k quoted as the target in the project proposal. It should be stated that there are two less years to 2010 since the proposal figures were estimated.

6.2 Cooling Water Savings

By removing heat from the hot product in the heat recovery section, the quantity of water required to cool a batch of food product was substantially reduced. This has the potential to reduce costs for factory cooling water and for effluent treatment. Table 10 gives the cost data based on the same assumptions on factory throughput as in Tables 8 and 9. The total cooling water savings to 2010 were estimated to be £ 495k for the example chosen.

Table 10: Potential annual costs and savings of cooling water by converting existing tubular heat exchangers to use heat recovery (no additional capital costs are required).

Converting existing tubular exchangers to heat recovery	2001, single factory	2010, four factories p.a.
Mains water costs (£)	14,254	513,160
Effluent treatment costs (£)	10,123	364,418
Mains water costs with heat recovery (£)	6,211	223,602
Effluent treatment costs with heat recovery (£)	4,411	158,790
Mains water cost savings (£) *	8,043	289,558
Effluent treatment cost savings (£) *	5,712	205,628

* Mains water cost 0.69 £/m³, effluent cost 0.49 £/m³

7. CONCLUSIONS AND FURTHER WORK

The data generated within this project on energy and cooling water savings highlighted the considerable potential for increasing operational efficiency by adopting heat recovery. Minimal changes to the Spiraflo exchangers are required if the viscosity of the food is below the equivalent to a 2 wt% starch solution (< 20 cP at shear rate 10 s^{-1}). This is the critical viscosity above which it will be necessary to improve the flow distribution in the shell. Changes such as increasing the tube spacing and streamlining the tube supports can be made without recourse to complex computational design techniques. These will allow heat recovery to be utilised for foods in the viscosity range between 2 and 3 wt% starch (20 to 200 cP at shear rate 10 s^{-1}). However, to extend the use of heat recovery to foods equivalent to 5 wt % starch (3,000 cP at shear rate 10 s^{-1}) a more radical design approach will be required.

To address the issue of foods with viscosities above 200 cP, a MAFF LINK project has been initiated in which CFD techniques will be used to design a novel tubular heat exchanger. The objective will be to arrive at a design that will recover heat from variable viscosity foods and provide thermal efficiencies similar to those for thin liquids. The types of tubular exchanger appropriate to this project will include 7 x 16 mm multi-tubes and 4 x 16 mm 'fibre' tubes. The food products will include gelatinised starches for generic work and foods manufactured by the industrial partners, e.g. tomato ketchup and salsas, fruit purees, baby foods, cook-in-sauces, soups and dressings. Of specific interest to the development of new food products are those foods that contain particulates in the size range 2 to 5 mm.

8. REFERENCES

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APPENDIX 1

Rheological characterisation of food products

Nomenclature used in appendix 1

σ	= shear stress	Pa
σ_0	= yield stress	Pa
K	= consistency coefficient	Pa.s ⁿ
$\dot{\gamma}$	= shear rate	s ⁻¹
n	= flow behaviour index	

Rheology is defined as the science of deformation and flow of matter. The term itself originates from the Greek ‘rheos’ meaning to flow. Rheology is applicable to all types of materials, from gases to solids.

Rheology is used in food science to define the consistency of different products.

Rheologically the consistency is described by two components, the viscosity (“thickness”, lack of slipperiness) and the elasticity (“stickiness”, structure). In practice, therefore, rheology stands for viscosity measurements, characterisation of flow behaviour and determination of material structure. Basic knowledge of these subjects is essential in process design and product quality evaluation.

Characterisation of materials

One of the main issues of rheology is the definition and classification of materials. Normal glass, for instance, is usually defined as a solid material, but if the thickness of an old church window is measured from top to bottom a difference will be noted. Glass does in fact flow like a liquid, albeit very slowly. Also, an air bubble captured in a glass object will rise, safely but slowly, due to the difference in density between air and glass. The process may, however, take centuries to observe and is of course not evident to human beings – the time of observation is too short.

Another way of defining materials rheologically is by the terms viscous, elastic or viscoelastic. Gases and liquids are normally described as viscous fluids. By definition an ideal viscous fluid is unable to store any deformation energy. Hence it is irreversibly deformed when subjected to stress; it flows and the deformation energy is dissipated as heat, resulting in a rise in temperature.

Solids, on the other hand, are normally described as elastic materials. An ideal elastic material stores all imposed deformation energy and will consequently recover totally upon release of stress. A viscous fluid can therefore be described as a fluid, which resists the act of deformation rather than the state of deformation, while an elastic material resists the act as well as the state of deformation.

A number of materials show viscous as well as elastic properties, i.e. they store some of the deformation energy in their structure while some is lost by flow. These materials are called viscoelastic and there are many examples among foodstuffs, such as starch-based puddings, mayonnaise, and tomato purées.

Shearing

In rheology, shearing of a substance is the key to knowledge of flow behaviour and structure. A sheared flow is achieved through flow between parallel plates, rotational flow between coaxial cylinders where one cylinder is stationary and the other one is rotating, telescopic flow through capillaries and pipes, and torsional flow between parallel plates.

To enable study of the viscosity of a material, the shearing must induce stationary flow of the material. The flow occurs through rearrangement and deformation of particles and through breaking of bonds in the structure of the material.

Shear stress (σ) is defined as:

$$\sigma = \frac{F}{A}$$

F = force, N

A = area, m²

shear rate as:

$$\dot{\gamma} = \frac{d\gamma}{dt} = \frac{dv}{dy}$$

and apparent viscosity of a fluid as:

$$\eta_a = \frac{\sigma}{\dot{\gamma}}$$

If we want to study the elasticity (structure) of a material, the shearing must be very gentle so as not to destroy the structure. One way to achieve this is to apply an oscillating shear to the material with amplitude low enough to allow an unbroken structure to be studied.

Shearing between parallel plates is normally used for the basic definition of shear stress and shear rate, corresponding to how much deformation is applied to the material and how fast.

Newtonian fluids

Newtonian fluids are those having a constant viscosity dependent on temperature but independent of the applied shear rate. One can also say that Newtonian fluids have direct proportionality between shear stress and shear rate in laminar flow:

$$\sigma = \eta \cdot \frac{dv}{dy} = \eta \cdot \dot{\gamma}$$

The proportionality constant is thus equal to the viscosity of the material. The flow curve, which is a plot of shear stress versus shear rate, will therefore be a straight line with slope η for a Newtonian fluid. The viscosity curve, which is a plot of viscosity versus shear rate, will show a straight line at a constant value equal to η .

A Newtonian fluid can therefore be defined by a single viscosity value at a specified temperature. Water, mineral and vegetable oils and pure sucrose solutions are examples of Newtonian fluids. Low-concentration liquids in general, such as whole milk, skim milk and most single strength fruit juices, may for practical purposes be characterised as Newtonian fluids.

Non-Newtonian fluids

Materials that cannot be defined by a single viscosity value at a specified temperature are called non-Newtonian. The viscosity of these materials must always be stated together with a corresponding temperature and shear rate. If the shear rate is changed, the viscosity will also change. Generally speaking, high concentration and low temperature induce or increase non-Newtonian behaviour.

Apart from being shear rate dependent, the viscosity of non-Newtonian fluids may also be time dependent, in which case the viscosity is a function not only of the magnitude of the shear rate but also of the duration and, in most cases, of the frequency of successive applications of shear. Non-Newtonian materials that are time independent are defined as shear thinning, shear thickening or plastic. Non-Newtonian materials that are time dependent are defined as thixotropic, rheopectic or anti-thixotropic.

Shear thinning flow behaviour

The viscosity of a shear thinning fluid (sometimes also denoted as a pseudoplastic fluid) decreases with increasing shear rate. Most liquid food systems belong to this category of fluids, e.g. yoghurt, fruit juice concentrates, salad dressings and tomato ketchup. The shear rate dependency of the viscosity can differ substantially between different products, and also for a given liquid, depending on temperature and concentration. The reason for shear thinning flow behaviour is that an increased shear rate deforms and/or rearranges particles, resulting in lower flow resistance and consequently lower viscosity. It should be noted that although sucrose solutions show Newtonian behaviour independent of concentration, fruit juice concentrates are always significantly non-Newtonian.

Hence a non-Newtonian fluid like yoghurt or fruit juice concentrate being pumped in a pipe shows decreased apparent viscosity if flow rate is increased. This means in practice that the pressure drop of a non-Newtonian fluid in laminar flow is not directly proportional to the flow rate as for Newtonian fluids in laminar flow.

Shear thickening flow behaviour

The viscosity of a shear thickening fluid increases with increasing shear rate. This type of flow behaviour is rare but can be found among suspensions of very high concentration. A shear thickening fluid exhibits dilatant flow behaviour, i.e. the solvent acts as a lubricant between suspended particles at low shear rates but is squeezed out at higher shear rates, resulting in denser packing of the particles. Typical examples of shear thickening systems are wet sand and concentrated starch suspensions.

Plastic flow behaviour

A fluid that exhibits a yield stress, is called a plastic fluid. The practical result of this type of flow behaviour is that a significant force must be applied before the material starts to flow like a liquid (often referred to as the ketchup effect). If the force applied is smaller than the force corresponding to the yield stress, the material stores the deformation energy, i.e. shows elastic properties, and hence behaves as a solid. Once the yield stress is exceeded, the liquid can flow like a Newtonian liquid and be described as a Bingham plastic liquid, or it can flow like a shear thinning liquid and be described as a viscoplastic liquid.

Typical plastic fluids are quarg, high pectin pineapple juice concentrate, tomato paste and certain ketchups. Outside the liquid food world, toothpaste, hand cream and greases are typical examples of plastic fluids.

A simple but still very effective way of checking a fluid's possible plastic properties is to just turn the jar upside down. If the fluid will not flow by itself, it probably has a significant yield value. If it flows by itself, but very slowly, it probably has no yield value

but a high viscosity. Information of this kind is of vital importance to process plant design regarding the dimensions and layout of storage and process tank outlets and pump connections.

Time-dependent flow behaviour

A thixotropic fluid can be described as a shear thinning system where the viscosity decreases not only with increasing shear rate but also with time at a constant shear rate. Thixotropic flow behaviour is normally studied in a loop test. In this test the material is subjected to increasing shear rates followed by the same shear rates in decreasing order. The time-dependent thixotropic flow behaviour is seen from the difference between the ascending and descending viscosity and shear stress curves. To recover its structure, the material must rest for a certain period of time which is characteristic for the specific material. This type of flow behaviour is shown by all gel-forming systems. Typical examples of thixotropic fluids are yoghurt, mayonnaise, margarine and ice cream. Outside the liquid food world perhaps brush paint is the most well-known product showing thixotropic behaviour.

A rheopectic fluid can be described as a thixotropic fluid but with the important difference that the structure of the fluid will only recover completely if subjected to a small shear rate. This means that a rheopectic fluid will not rebuild its structure at rest.

An anti-thixotropic fluid can be described as a shear thickening system, i.e. one where the viscosity increases with increasing shear rate, but also with time at a constant shear rate. This type of flow behaviour is very uncommon among foodstuffs. As with thixotropic fluids, the flow behaviour is illustrated by a loop test where the shear rate is ramped up and down in a fixed time.

Flow behaviour models

For the adaptation of viscosity measurement data to process design calculations some kind of mathematical description of the flow behaviour is required. For that purpose several models are available, e.g. Ostwald, Herschel-Bulkley, Steiger-Ory, Bingham, Ellis and Eyring. These models relate the shear stress of a fluid to the shear rate, thus enabling the apparent viscosity to be calculated, as always, as the ratio between shear stress and shear rate.

By far the most general model is the Herschel-Bulkley model, also called the generalised power law equation, which in principle is an extended Ostwald model. The main benefit of the generalised power law equation is its applicability to a great number of non-Newtonian fluids over a wide range of shear rates. Furthermore, the power law equation lends itself readily to mathematical treatment, for instance in pressure drop and heat transfer calculations.

The generalised power law equation is applicable to plastic as well as shear thinning and shear thickening fluids according to the following equation:

$$\tau = \sigma_0 + K \cdot \dot{\gamma}^n$$

Suitable modification of the generalised power law equation makes it possible to rewrite it to express each type of flow behaviour.

For Newtonian fluids the power law equation has $K = \eta$ and $n = 1$:

$$\sigma = K \cdot \dot{\gamma}^n = \eta \cdot \dot{\gamma}$$

For a plastic fluid the power law equation is used in the fully generalised form, with $n < 1$ for viscoplastic behaviour and $n = 1$ for Bingham plastic behaviour.

For a shear thinning or shear thickening fluid the power law equation has $n < 1$ and $n > 1$ respectively, and becomes:

$$\sigma = K \cdot \dot{\gamma}^n$$

The power law equation was used for this project where the foods and starch solutions displayed shear thinning behaviour.

APPENDIX 2

Rheological data on starches

Rheological data was taken for the starches and products as the project progressed through phases 1 and 2. In all cases, the power law model was used to relate shear rate to shear stress data. The following 5 tables give the data for the Colflo 67 solutions used throughout the project, with Figure 1 displaying the viscosities at each wt%. Data for the food products were taken as required, using the same format, and are presented in the main text.

Table 1: Power Law Parameters for 1 wt% Colflo 67			
Temperature (°C)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
22	0.0018	1.03	1.9
40	0.0012	1.04	1.3
60	0.0014	1.00	1.4
80	0.0008	1.07	0.9

Table 2: Power Law Parameters for 2 wt% Colflo 67			
Temperature (°C)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
19	0.024	0.82	16
41	0.013	0.86	10
59	0.007	0.91	6
76	0.005	0.93	4

Table 3: Power Law Parameters for 3 wt% Colflo 67			
Temperature (°C)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
20	0.47	0.62	193
34	0.26	0.66	119
40	0.20	0.69	99
57	0.10	0.73	54

Table 4: Power Law Parameters for 4 wt% Colflo 67			
Temperature (°C)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
23	2.54	0.44	699
39	1.43	0.50	454
60	1.46	0.48	444
67	1.25	0.50	396

Table 5: Power Law Parameters for 5 wt% Colflo 67			
Temperature (°C)	Consistency Coefficient, k (Pa.sⁿ)	Flow Behaviour Index, n	Viscosity at Shear rate 10 s⁻¹ (cP)
29	12.19	0.36	2,814
42	11.74	0.33	2,495
58	9.48	0.32	1,986
70	8.71	0.29	1,715