THE SEPARATION OF CARBOHYDRATES BY CONTINUOUS PRODUCTION SCALE CHROMATOGRAPHY

A thesis submitted

by

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Summary

A review of general chromatographic theory has been made together with a survey of the factors which may affect the performance of larger scale chromatographic columns. The mechanism of the chemiadsorption chromatography employed in this research has been described in detail. Various industrial chromatographic processes used for the refining of sugars have been outlined.

The design and construction of a liquid-solid semi-continuous chromatographic refiner, (SCCR) has been reported. The equipment comprised ten 108 mm internal diameter stainless steel columns each containing a packed bed approximately 650 mm in length. Counter-current operation of the mobile and stationary phases was achieved by a programmed sequencing of inlet and outlet valves associated with each column.

Commissioning and operation of the SCCR has been reported. An ion exchange resin, Zerolit 225, which was charged in the calcium form, was used as the stationary phase. Deionised water was the mobile phase. For what is believed to be the first time, a complete separation of a fructose-glucose mixture into two 99.9% pure products was achieved continuously in one pass through the equipment. The effects of feed concentration, feed flow rate and change of switch period were investigated.

Refining of a three component system: fructose, glucose and dextran, a polyglucose, was also undertaken. 99.9% pure fructose was obtained with a solids concentration of greater than 10% w/v.

Significant increases in product concentrations have been achieved by collection of selected fractions of each product stream.

An equilibrium plate model has been used in an attempt to simulate the performance of the semi-continuous chromatograph. Infinite dilution equilibrium data for the model were obtained from batch experimental work. Various improvements have been made in calculating the average on-column concentrations of the sugars. The model was also adapted to handle the three component system. Results achieved show partial agreement with experimental findings.

A thesis submitted by John Charles Gould, B.Sc. to the Faculty of Engineering at the University of Aston in Birmingham, for the Degree of Doctor of Philosophy.

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Key Words: Production-scale, chromatography, fructose, glucose, dextran.

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CHAPTER 1

INTRODUCTION

1 Introduction

Chromatography is well known as an analytical tool. It offers a fast and efficient method of analysis for both gaseous and liquid samples. The various mechanisms of chromatographic separation include physical or chemical adsorption, exclusion, ion exchange and partition. A typical analytical chromatography column may be 0.25 m long and 6 mm in diameter.

The chromatographic process is essentially batch with a small sample of feed material being eluted through a column using a suitable mobile phase. Recently much interest has been centred on chromatography as a commercial chemical engineering unit operation. Well known examples are found in the sugar industry, many since the disclosure by the Boehringer Mannheim Company in a patent (1) of a chromatographic process for the separation of sugars. This includes the separation of the isomers fructose and glucose.

Examples of these applications are: the Finnsugar (2) molasses desugarisation plant which uses columns in excess of 1 m diameter. The Sudzuker (3) process for the same duty uses columns of a similar size. These production scale units operate in the batch mode.

If high purity products and reasonable throughputs are to be achieved using repetitive batch techniques, then either total resolution must occur in a short time or a considerable quantity of contaminated products recycled. If a continuous countercurrent flow scheme is used where the mobile phase and the stationary phase move in opposite directions to each other, only partial resolution is necessary to collect pure products. The whole column length may be used to achieve this partial resolution thus allowing severe overloading by co-current standards and consequently greater throughput with high purity.

In attempts to achieve a satisfactory continuous counter-current flow scheme three basic stages of development have occurred - moving bed, moving column and simulated moving bed. The most successful method to date has been the simulated moving bed mainly because the need for solids handling equipment or mechanical seals has been eliminated.

A successful version of what is known as a semi-continuous chromatographic refiner (SCCR) was constructed by Barker and Deeble (4) in 1974 at the University of Aston in Birmingham. This was a gas chromatograph with twelve interlinked 76 mm diameter columns. Halocarbon mixtures were successfully separated with purities of 99.9% by partition chromatography using this refiner.

A liquid-solid chromatograph was constructed by Barker, Ellison and Hatt (5) in which ten 50 mm diameter columns were packed with a porous silica. This SCCR utilised exclusion chromatography to fractionate the polymer dextran.

A ten column liquid solid chromatograph was constructed by Barker and Ching (6) to investigate the separation of sugars by chemisorption chromatography. The column diameter used was 25.4 mm. Total separation was not achieved. This was thought to be due to the relatively large liquid hold-up volume in the valves and pipe lines associated with each column.

In the current research programme a larger 108 mm diameter SCCR has been designed and constructed. This was a production scale unit which made the hold-up volume insignificant compared with the total column volume. This unit had the added advantage of allowing commercial assessments to be made using an industrial by-product as a feedstock for production scale chromatography.

During experimental work a study was to be made of the rig's ability to refine or separate sugars under various operating conditions. Throughput, product purities and product concentrations were also to be investigated. Data would be collected to help develop and improve an existing mathematical model for predicting SCCR performance. A comparison of data from the 25.4 mm diameter SCCR would allow scale-up factors to be assessed in a qualitative and quantitative manner.

Most of the experimental work was to be carried out on the binary fructose-glucose system. Not only would this allow comparison with previous data but may also have industrial application. This is because a growing number of countries are manufacturing sugar syrups from starch sources using enzyme action. The raw material for such a process may be corn, maize or rice. If the fructose content could be enhanced in this syrup, a natural sweetener with a lower calorie intake is produced which is of use to the food industries. This is possible because fructose is about 1.5 times as sweet (7) as an equivalent amount of sucrose in cold solution.

Until quite recently when large scale batch chromatographs were developed, separation of fructose and glucose was difficult and expensive to achieve. A continuous process for the separation of the two isomers therefore merits investigation.

CHAPTER 2

LITERATURE SURVEY

2 Scope

It is convenient to divide the review into three parts. In the first part terminology, basic concepts and theory are reported. In part two the literature describing the separation of sugars and production of high purity fructose syrups using chromatographic processes is reviewed. The final section deals with the scale up of the system from an analytical tool to production scale equipment.

2.1 Introduction to Chromatography

Chromatography may be described as a separation or refining process. Resolution of two or more components will occur if they exhibit different equilibrium distributions between two immiscible phases. This means that they have different distribution coefficients K_d where

$$K_d = \frac{\text{concentration of a component in the stationary phase}}{\text{concentration of a component in the mobile phase}}$$

Most of the chromatography practised today takes place in a packed column through which an eluting fluid is allowed to flow. This is generally referred to as the mobile phase whilst the packing is the stationary phase. The stationary phase is selected so that each component has a different affinity for this phase and thus a different K_d. The components therefore migrate through the column at varying rates. Migration only occurs when components are in the mobile phase so those components with a distribution favouring this phase will migrate faster than those with a distribution favouring the stationary phase. Thus, depending on the length of the packed bed partial or total separation will take place.

Chromatographic techniques may be classified according to the type of mobile phase and stationary phase selected. Gas chromatography, GC, includes all those methods in which the mobile phase is a gas. Liquid chromatography, LC, describes all those methods in which the mobile phase is a liquid, whilst in GC and LC different stationary phases give rise to the names gas-liquid chromatography GLC, gas-solid chromatography GSC, liquid-liquid chromatography LLC and liquid-solid chromatography LSC. A further classification of liquid chromatography arises from the different retention mechanisms in operation - adsorption, exclusion, ion exchange and partition chromatography.

Adsorption chromatography involves the association of solute molecules and the active sites within the stationary phase. Such associations may be physical or chemical but do not involve the exchange of ions. Suitable ion exchange resins however are often used as the stationary phase. Such separations are often known as ligand exchange chromatography.

Exclusion chromatography uses a porous solid such as silica for the stationary phase. Separation occurs because of differences in the sizes of the various solute molecules. Those small enough to penetrate the porous matrix are retained for a longer period than the larger molecules which elute first. The method is widely used for the fractionation of proteins and polymers.

Ion exchange chromatography involves the continuous reversible exchange of ions between electrolytes and the ion exchange resin. Separation occurs as a result of different affinities of different solute molecules for the resin.

Partition chromatography relies on the absorption of solutes by an inert solid support coated with a liquid stationary phase. There are three modes of chromatographic operation: elution, frontal and displacement. In liquid chromatography the elution technique is the only one which allows quantitative separation and consequently most theoretical work has been performed in relation to this type. In this method, the sample components are injected at the beginning of the chromatographic column and development occurs as the eluting mobile phase flows through the bed.

2.1.1 Definitions and Terminology

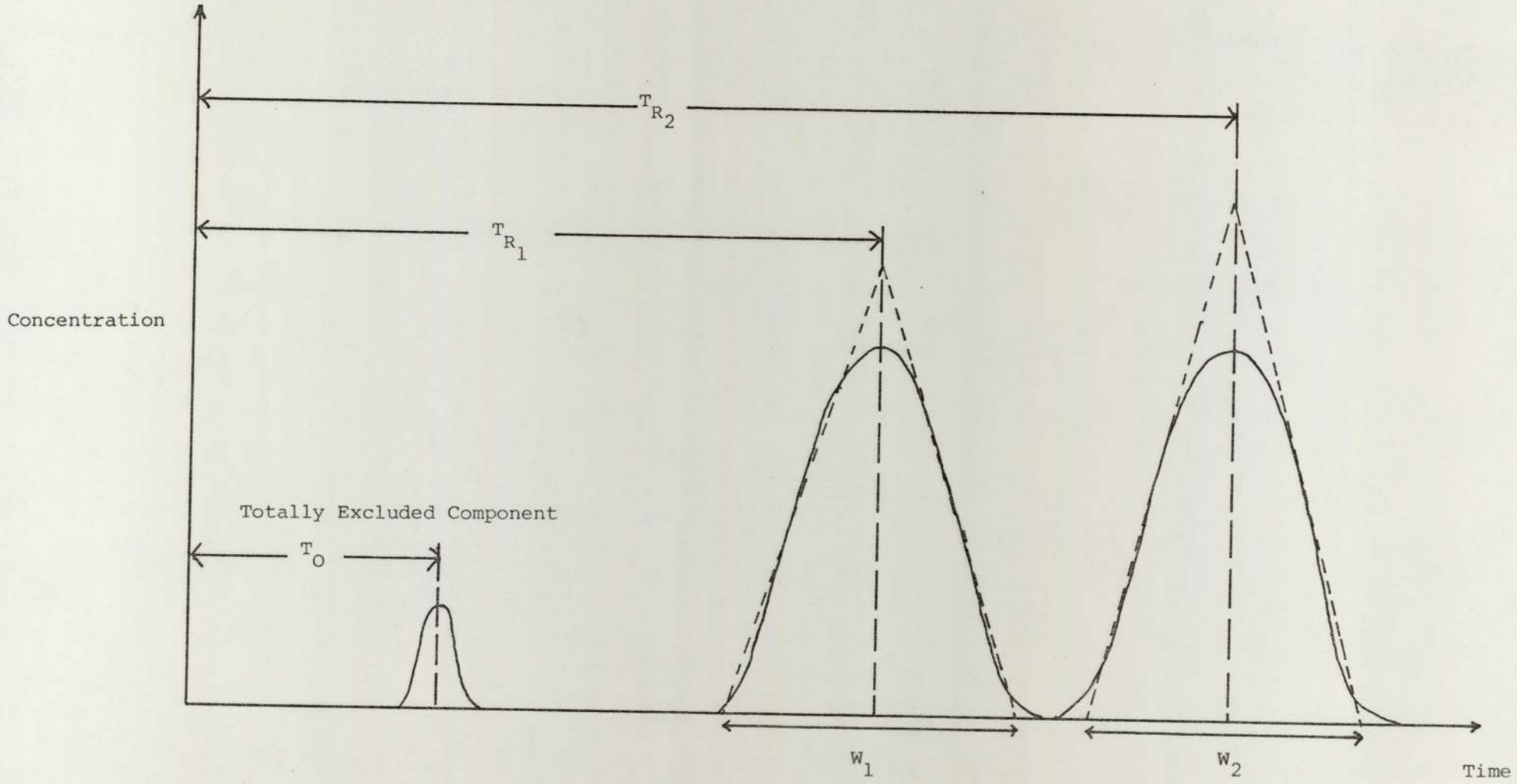
If, after a small sample injection, the eluent flowing from a chromatographic column is monitored, an approximately Gaussian concentration curve of that component will result. Under such conditions it is possible to relate the time taken for elution of the peak to the equilibrium distribution coefficient K_d (8). If this time period measured at the peak centre, known as the retention time T_R, is multiplied by the eluent flow rate the retention or elution volume V_R is calculated. When a symmetrical peak is obtained, the amount of feed sample has no effect on the retention time. This is because a linear distribution isotherm exists between the concentrations of the components in each of the phases.

The curves which result after an elution are known as chromatograms. A typical example is shown in Fig. 2.1.

The fundamental retention equation (9) for any chromatographic process is given as equation 2.1

$$V_R = V_M + K_d V_S \qquad 2.1$$

FIG. 2.1 A TYPICAL CHROMATOGRAM



Where V_R = component retention volume

V_M = volume of mobile phase in column

K_d = equilibrium distribution coefficient

V_S = volume of stationary phase in column

An important parameter in chromatography is the capacity factor k'.

$$k' = K_d \frac{V_S}{V_M} \qquad 2.2$$

Combining equations 2.1 and 2.2 gives

$$V_R = V_M(1 + k') \qquad 2.3$$

Rearranging equation 2.3 gives k' in terms of retention volumes.

$$k' = \frac{V_R - V_M}{V_M} \qquad 2.4$$

The retention volume may be obtained directly from the chromatogram and the mobile phase volume will be a known property of the column. Also from the chromatogram, a measure of the degree of separation or resolution may be obtained.

$$R_s = 2\left(\frac{T_{R_2} - T_{R_1}}{W_1 + W_2}\right) \qquad 2.5$$

Where R_s = resolution

T_R = retention time of component

W = peak width at base line.

The larger the value of R_s the better the separation.

The two characteristics which determine the peaks' "overlap" i.e. incomplete separation, are (1) the distance between the peak centres and (2) the width of the peaks. This may be illustrated by studying Fig. 2.2.

In Fig. 2.2a the peaks overlap which indicates incomplete separation. In Fig. 2.2b complete separation has occurred but the peaks are broad and the distance between the peak centres is large. This implies a long elution time. In Fig. 2.2c complete separation has occurred and the peaks are narrow and close. This represents complete separation in an acceptable elution time. From equation 2.5 an R_s value of 1.0 is considered satisfactory. Purnell (10) developed a relationship between resolution and the fundamental parameters:

$$R_s = \frac{1}{4}\left[\frac{\alpha-1}{\alpha}\right]\left[\frac{k'_2}{1+k'_2}\right][N]^{\frac{1}{2}} \qquad 2.6$$

Where $\alpha = \frac{K_{d_2}}{K_{d_1}}$ (Ratio of equilibrium distribution coefficients)

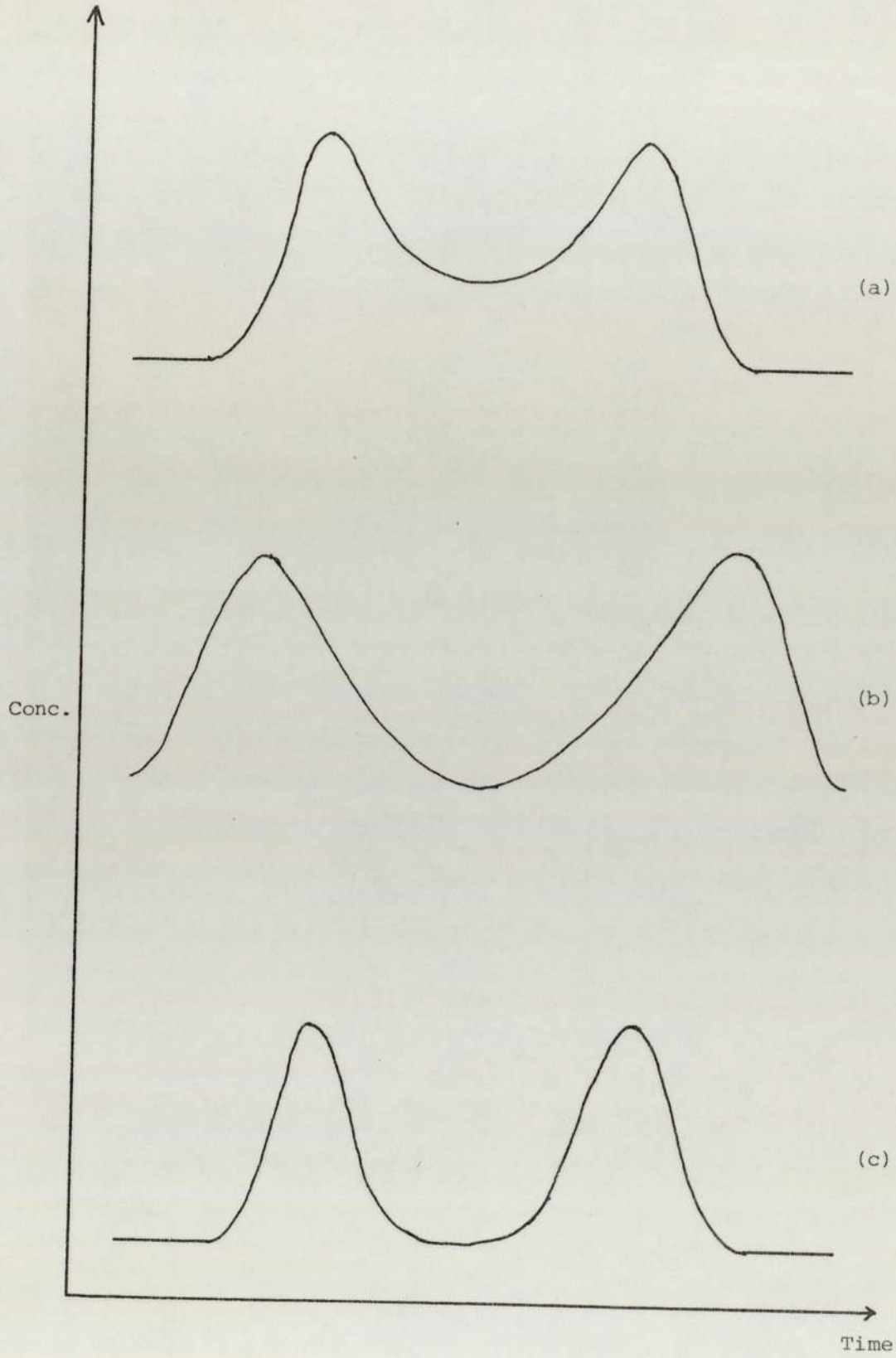
k'_2 = capacity factor of the most retarded component

N = no of theoretical plates in the column.

2.1.2 Theory of Band Broadening

Two factors determine whether a separation can be achieved in elution chromatography. These are (i) the distance between the peak centres (ii) the fact that the peak must be kept compact to avoid overlap. The control of the migration rates which affect the distance between the peak centres is related to the thermodynamic equilibrium of the process. The peak or band broadening is a result of the column dynamics and it is this which theories try to predict.

FIG. 2.2 CHROMATOGRAMS ILLUSTRATING RESOLUTION



2.1.2.1 The theoretical plate concept

This concept was introduced into chromatography by Martin and Synge (11) in 1941 because of its success in describing distillation processes. They suggested that the chromatographic column could be considered to consist of a number of layers of packing each of which was equivalent to a theoretical plate. The height of such a layer is referred to as the height equivalent to a theoretical plate (HETP). They postulated a number of assumptions about the theoretical plate. (1) The concentration of solute issuing from each plate is in equilibrium with the average concentration of the solute in the stationary phase of that plate. (2) The diffusion of solute from one plate to the next is negligible. (3) The mobile phase flow is discontinuous consisting of stepwise additions of volumes of mobile phase equal to the volume of mobile phase per plate. (4) At equilibrium the distribution ratio of one solute between the mobile and stationary phase within the plate is independent of the other solutes present.

Using these assumptions their model suggested that a single solute band would spread into a Gaussian distribution curve and the degree of spreading of this curve could be quantified by its variance. The HETP could then be defined as:-

$$H = \frac{d\sigma_z^2}{dz}$$

where σ_z^2 = variance or length based second moment

z = distance along a column of length L.

Martin and Synge also reported the dependence of HETP on the mobile phase velocity u, and particle diameter d_p. They also mentioned

that as the flow rate is reduced so the significance of the longitudinal diffusion from plate to plate increases.

2.1.2.2 The continuous model

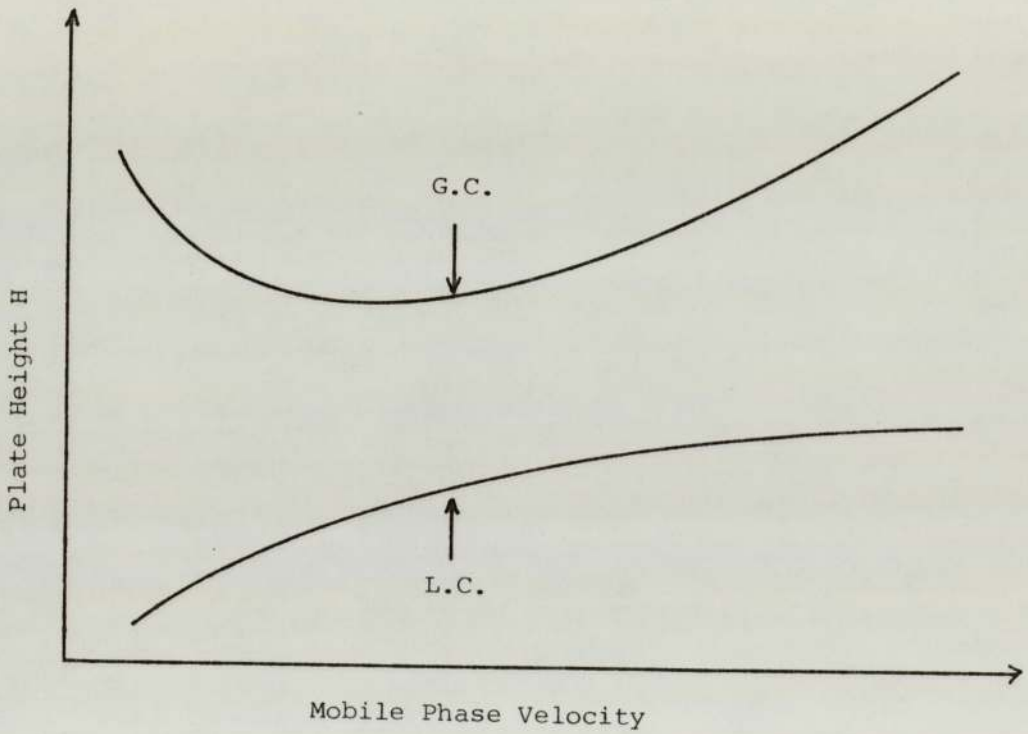
Glueckauf (12) converted this discrete plate model into a continuous one by reducing the volume of the plate to an infinitessimally small value. The predicted concentration profile exhibited a Poisson distribution which became Gaussian when N was greater than 100.

The most significant deviation of the plate models' concepts from real column processes rests on the assumption of plate wide equilibrium. In actual situations equilibrium is only reached at the peak maximum. In addition the plate models fail to account for the contributions of molecular structure, sorption phenomena, temperature, molecular distribution and flow pattern towards band spreading. However the plate height is a useful and widely used parameter for the characteristisation of band spreading and column efficiency.

2.1.3 GC and LC Theory: A comparison.

Nearly all early theoretical treatment was for gas chromatography. This was generally considered in the batch mode. The similarities between GC and LC are more prominent than the differences. Band broadening and separation occur by virtue of the same kinds of thermodynamic, flow, kinetic and diffusional processes. Thus they are subject to the same theoretical laws. The differences are in the physical properties of the mobile phase. These differences may be illustrated by comparing the variation of plate height with mobile phase velocity and the equations that describe the two curves. See Fig. 2.3.

FIG. 2.3 COMPARISON OF THE CHROMATOGRAPHIC PLATE HEIGHT CURVES FOR GAS CHROMATOGRAPHY AND LIQUID CHROMATOGRAPHY



The difference in shape occurs primarily because diffusion coefficients in liquids are 10^4 to 10^5 times smaller in liquid than those in gases. Thus they are far less significant. The GC curve may be represented by equation 2.7

$$H = A + \frac{B}{u} + C_s u \qquad 2.7$$

Where H = plate height

A = the eddy diffusion contribution

B = the axial diffusion contribution

C_s = the stationary phase mass transfer resistance term

u = the mobile phase velocity.

The LC curve may be represented by

$$H = C_s u + C_m u + \left[\frac{1}{A} + \frac{1}{C_m u} \right]^{-1} \qquad 2.8$$

Equation 2.7 was based on a model proposed by Lapidus and Ammundson (13) which was improved by Van Deemter, Klinkenberg and Zuiderweg (14). Equation 2.8 was proposed by Giddings (15) in his generalised non-equilibrium theory. Each of the terms in this equation contribute independently and additively to band spreading in LC and are discussed below.

The $C_s u$ term relates to the non equilibrium which results from the resistance to mass transfer in both the mobile and stationary phase. This may arise from slow adsorption or desorption from a surface.

The extent of broadening is a function of the rate of flow of mobile phase relative to the rate of transfer of solute between phases. Giddings suggests that true equilibrium only exists at the centre of the band - the stationary phase has a lag in its equilibrium value whilst the concentration of a solute in the mobile phase will always be ahead of its equilibrium concentration. The extent of non-equilibrium can be reduced by having a low mobile phase flowrate thus preventing rapid concentration changes. This is illustrated in Fig. 2.4.

The $C_m u$ term represents the slow rate of mass transfer of solution by diffusion in the mobile phase that is stagnant in the pores of the stationary phase. Giddings suggests it may be evaluated thus

$$C_m u = \left[\frac{1}{30} f(\Phi, k') \right] \frac{d_p^2 u}{Dm} \qquad 2.9$$

Where Φ is the fraction of mobile phase occupying the stagnant volume in the stationary phase.

k' = the capacity factor

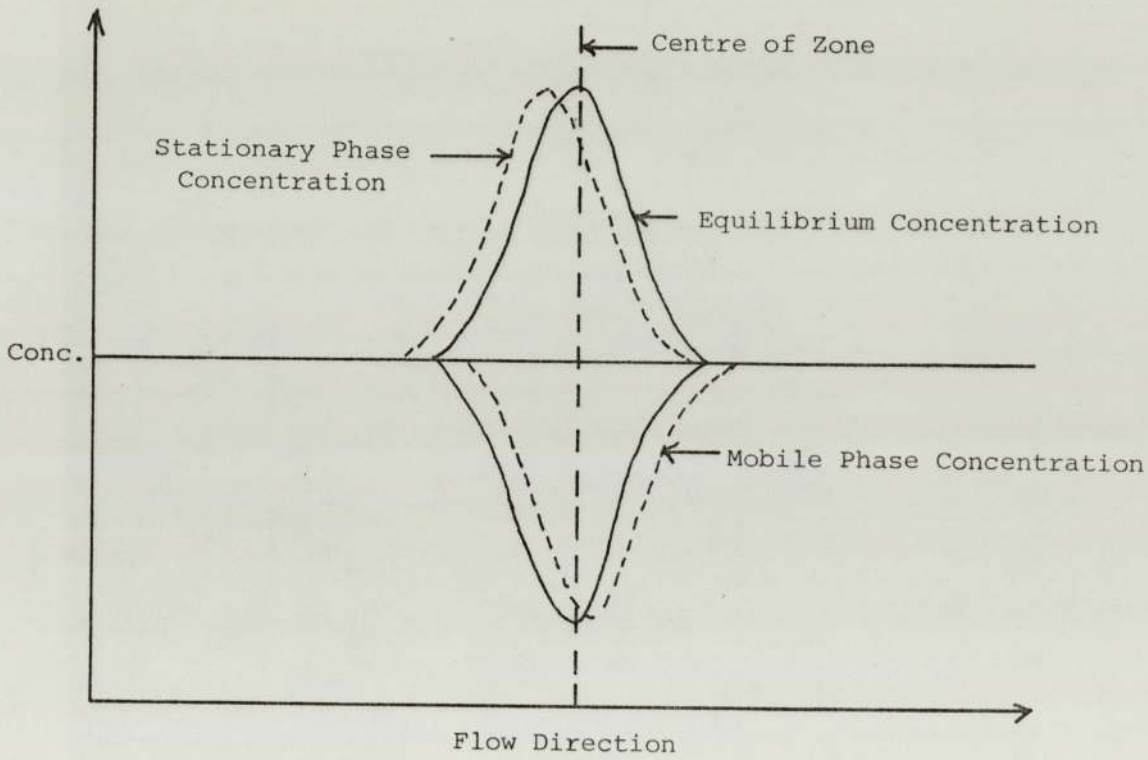
d_p = average particle diameter

u = mobile phase velocity

Dm = radial diffusion coefficient

Fig. 2.3 shows a much flatter rise of HETP with increasing u. This is a result of complex flow processes occurring in the mobile phase liquid. Since solute diffusion is far slower in liquids lateral mixing plays an important part in the shape of the profile. Giddings proposes that this results in a coupling effect of the eddy diffusion and mobile phase resistance to mass transfer terms to give the final term in equation 2.8.

FIG. 2.4 COMPARISON BETWEEN ACTUAL AND EQUILIBRIUM COMPONENT CONCENTRATION PROFILES FOR NORMAL ELUTION CHROMATOGRAPHY



$$\left[\frac{1}{A} + \frac{1}{C_m u}\right]^{-1} = \left[\frac{1}{2\lambda d_p} + \frac{Dm}{d_p^2 u}\right]^{-1} \qquad 2.10$$

2.1.4 A direct measurement of the number of theoretical plates

If it is assumed that the solute is introduced at the beginning of a column as a small narrow band, due to the band broadening phenomena already discussed an approximately Gaussian concentration profile is achieved. The width of this band is a measure of the efficiency of the chromatographic column. The broader the band - the less efficient the column. Glueckauf (12) has related the elution time necessary and the variance of the band to the number of theoretical plates in the column. He selects a reference point to measure the band width as the height at the peak maximum divided by 'e' the base of exponential logarithms.

Equation 2.11 then allows calculation of N, the number of plates. See Fig. 2.5.

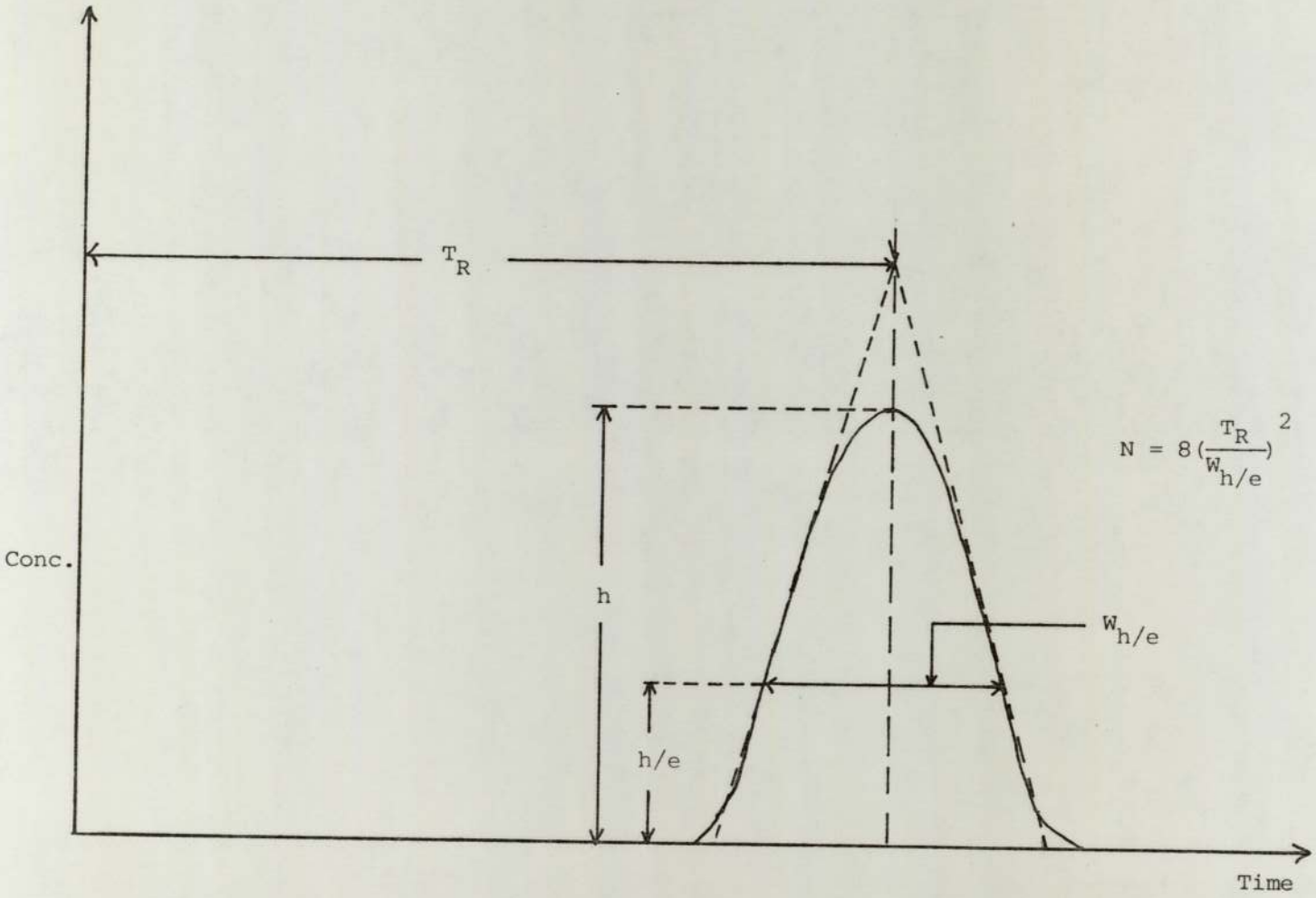
$$N = 8\left(\frac{T_{R_i}}{W_{h/e}}\right)^2 \qquad 2.11$$

T_{R_i} = retention time of component

h = height at peak maximum

$W_{h/e}$ = width of band at $\frac{h}{e}$ expressed in time units

FIG. 2.5 RELATIONSHIP BETWEEN NUMBER OF PLATES AND PEAK CHARACTERISTICS



2.2 Separation of fructose from carbohydrate mixtures

Fructose is the sweetest of all natural sugars. It offers the lowest calorific intake per unit weight for the same sweetness. It causes no hyperglycemia and no insulin release. In cold solution it is about 1.5 times as sweet as an equivalent sucrose solution (7).

A higher concentration of fructose in sweeteners would thus result in a lower calorie food with no loss in enjoyment. There are two main methods of producing high fructose syrups, (i) enzymatic conversion of starch (ii) inversion of sucrose.

Countries in various parts of the world where neither sugar cane nor sugar beet is homegrown have been investigating conversion of local starch sources to sweetening syrups. These include corn, maize and rice. Enzymes are used to produce firstly a glucose rich product from the starch source and then to produce a fructose-glucose mixture.

The major sources of sucrose are sugar cane in the tropical areas and sugar beet in the more temperate zones of the world. As sucrose molecules consist of one molecule of glucose and one molecule of fructose the product from an inversion process contains an equal amount of both sugars. A complete inversion is difficult to obtain in bulk and the product often contains some residual disaccaride. A further refining process is therefore necessary to enrich the fructose content.

During the last fifteen years large scale separations of the two sugars have been performed using chromatographic columns packed with ion exchange resins. The most notable of these processes were those of the Colonial Sugar Refining Company (16) and the Boehringer Mannheim Company (1).

2.2.1 Processes for manufacturing high purity fructose syrups

2.2.1.1 The Colonial Sugar Refining Co. Process (16)

The Company was granted a patent in 1967 for the separation of fructose and glucose. The ion exchange resin used was Dowex 50W which was a sulphonated polystyrene resin with 4% cross linking with divinyl benzene. It was charged in the calcium form. This resin was able to hydrolyse and separate invert sugar into glucose and fructose fractions. Their 1.8 m packed bed was operated in a co-current batch mode with various recycle fractions. From their published results (16) the fractions collected varied in purity and concentration. One fructose rich product had a concentration of 29% w/w solids of which 82% was fructose another was 24% w/w solids of which 95% was fructose. The process was operated at 60^{o}C which necessitated the lagging of all process lines, vessels and columns. This was to keep the viscosity of the syrup low enough to enable reasonable pumping costs to be maintained.

2.2.1.2 The Boehringer Mannheim Process (1)

The German company was granted a British patent in 1967. In it they claimed a process in which fructose and glucose products were obtained using sucrose as a feed material. They used an ion exchange resin, Dowex 50W in six glass columns in series each 15 cm in diameter. A total separating length of 9 m was used. The eluent flow rate was quoted as between 1-2 $cm^{3}min^{-1}cm^{-2}$. The resin was charged at room temperature and at particular conditions of pH with calcium chloride solution to render it in the calcium form. This method allows the resin to retain between 5 and 30% of its active sites in the hydrogen form. It is this property which allows complete hydrolysis of sucrose to take place in situ. The calcium

ions then retard the fructose by forming a complex and thus effect the separation.

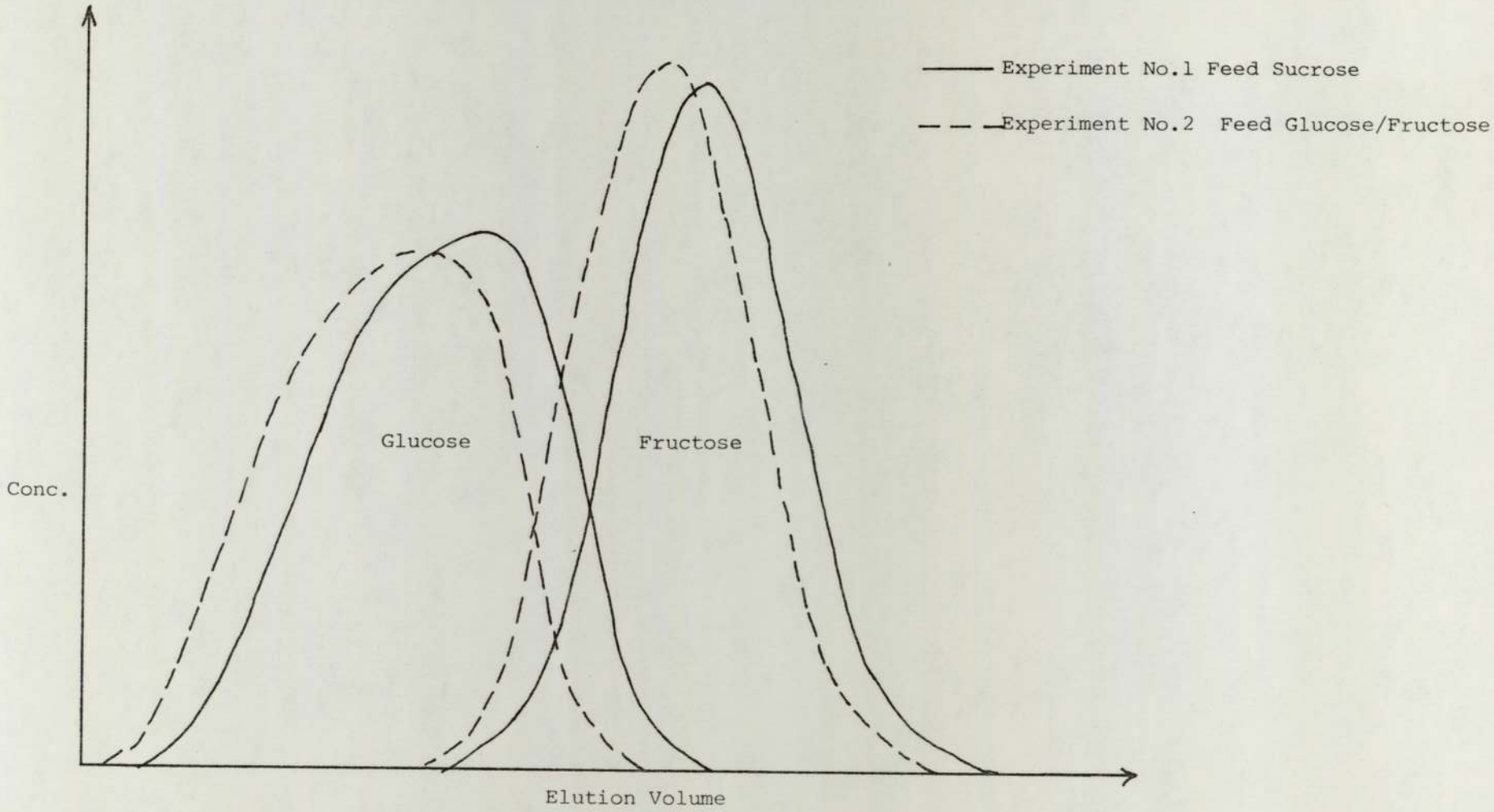
The process is operated at 60°C to keep the viscosity low and deionised water is the eluting solvent. Two sets of results have been published. The first set when only sucrose was used as the feed, the second set when a mixture of fructose, and glucose was used as the feed. The published concentration profiles are shown in Fig. 2.6 and as can be seen no sucrose emerges with either product thus confirming total hydrolysis. The company used a repeated feed injection operation but did not achieve total separation. The 'mixture' or 'overlap' fraction is recycled.

This simultaneous hydrolysis and separation process offers a more flexible alternative for the production of high purity fructose. Work has been performed by Chuah (17) using a semi-continuous chromatographic system which shows complete inversion and fructose purities of up to 95% from a pure sucrose feed material.

2.2.1.3 Other processes for fructose production

In 1895 Bruyn and Eckenstein (18) discovered that glucose could be isomerised to fructose using an alkaline catalyst at an elevated temperature. In more recent times numerous patents were granted which detailed the isomerisation of glucose to fructose using sodium hydroxide and ion exchange resins. Scallet (19-21) and Ozazaki (22) have been prominent in this field. The use of alumina with various liquid coatings has been reported by Parish (23), Tsao (24) and Barker (25) to achieve separation.

FIG. 2.6 RESULTS PUBLISHED BY THE BOEHRINGER PATENT



The advent of enzyme technology led to a higher conversion rate from glucose to fructose. Various micro-organisms were used by different research teams who have published successes: Yamanaka (26), Danno (27) and Takasaki (28).

A commercial enzyme process was reported in 1974 by Newton and Wardrip (29) of the Clinton Corn processing company for manufacturing a fructose rich corn syrup.

The enzyme glucose isomerase was used to manufacture 'sweetzyme' a particulate material containing this enzyme (30). This was then used commercially to produce high purity fructose products from maize. Large scale production plants operate in the Netherlands, Belgium and Spain. Also in 1974 the Albion Sugar Company built a plant at Tilbury-on-Thames (31) to produce a high purity fructose syrup by enzyme action and a chromatographic technique. Due to a sugar subsidy policy within the European Economic Community the plant has never been commissioned.

2.2.2 Mechanism of separation used in this Research

It is the structural differences in the two isomers which enables separation to be achieved. It has long been known that fructose forms a complex with calcium ions in aqueous solution. Reviews of the research carried out by earlier workers on the metal complexing of fructose have been published by von Lippman (32) and Vogel (33). In these works no conclusions have been presented about the mechanism of the complexing. More recently Angyal (34-37) has published a series of papers proposing a hypothesis for the mechanism. His results are based on evidence obtained using nuclear magnetic resonance or n.m.r.

2.2.3 Molecular structures of glucose and fructose

The conventional representation of glucose is a straight chain. However when in solution glucose exists as four isomeric ring structures. Two of these structures are six-membered rings and are referred to as glucopyranoses. The other two are five-membered rings called glucofuranoses. See Fig. 2.7.

An equilibrium exists between these structures and this is shown in Table 2.1. These values are for a 1 molar solution in pure water at 25°C.

In his hypothesis Angyal (34) suggested that for a successful complex formation, a sequence of eq-ax-eq-ax arrangement of oxygen atoms on adjacent carbon atoms is desirable. This is not exhibited in any of the glucose isomers. The abbreviations "eq" stands for equatorial - which means that the oxygen atoms lie in a plane horizontal or nearly horizontal to a carbon atom, whereas "ax" means axial when the oxygen atom lies in a plane vertical to the carbon atom.

In solution fructose exists as isomeric ring structures as shown in Fig. 2.8. The equilibrium proportions present of these anomers in a 1 molar solution in pure water is shown in Table 2.2.

It will be seen from Fig. 2.8 that an equilibrium exists between two forms of the β-D-fructopyranose. This is the component present in the highest concentration under the conditions described above. The equilibrium arises from the variation in the "chair" shape of the ring in relation to the equatorial or axial position of the CH_2OH group. These are referred to as the $^5C_2$ fructopyranose and the $^2C_5$ fructopyranose and describe the ring structure as viewed from a certain position. The equilibrium between these two structures lies predominantly towards the $^2C_5$ form.

FIG. 2.7 FORMS OF GLUCOSE IN SOLUTION

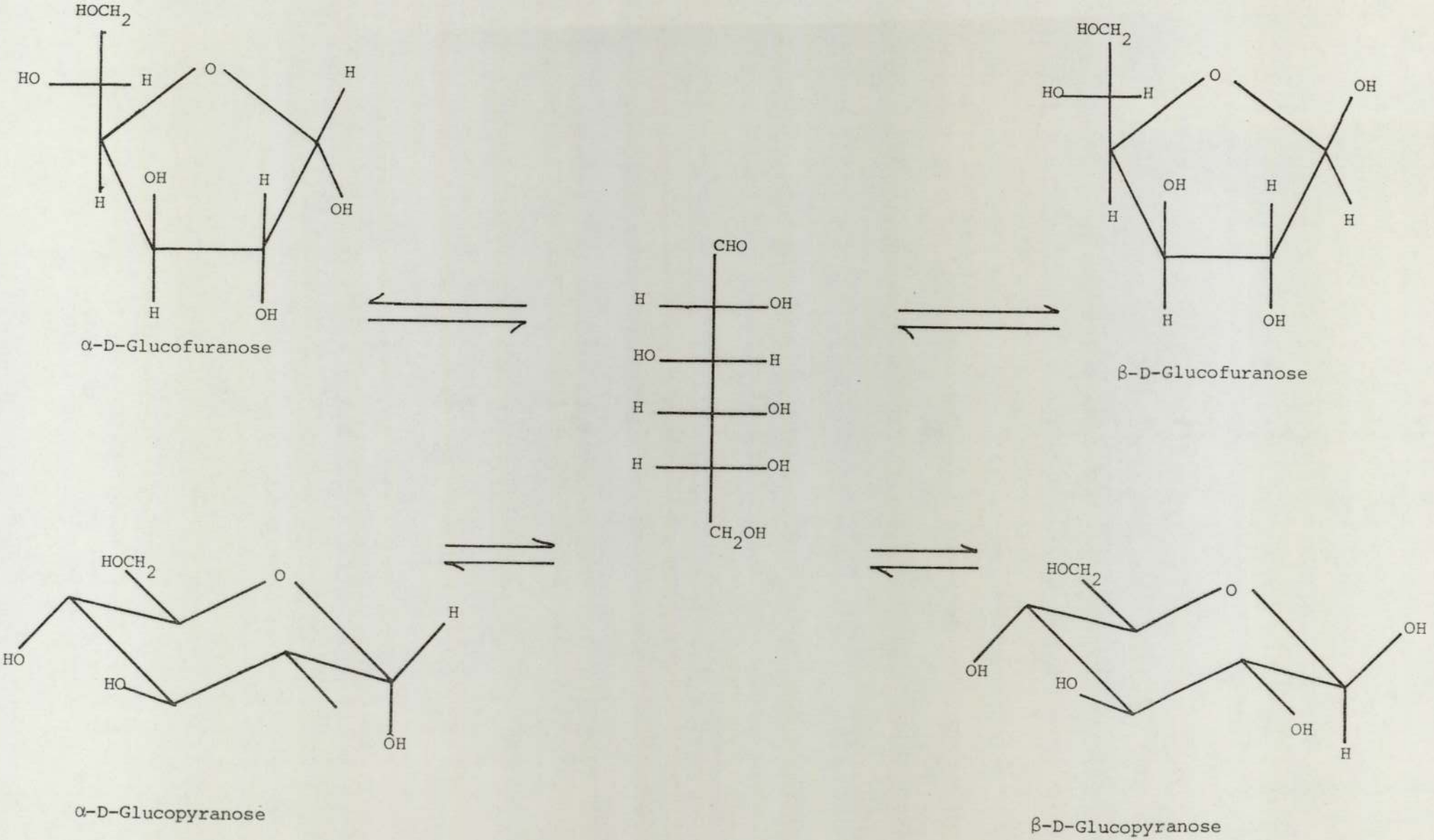


FIG. 2.8 FORMS OF FRUCTOSE IN SOLUTION

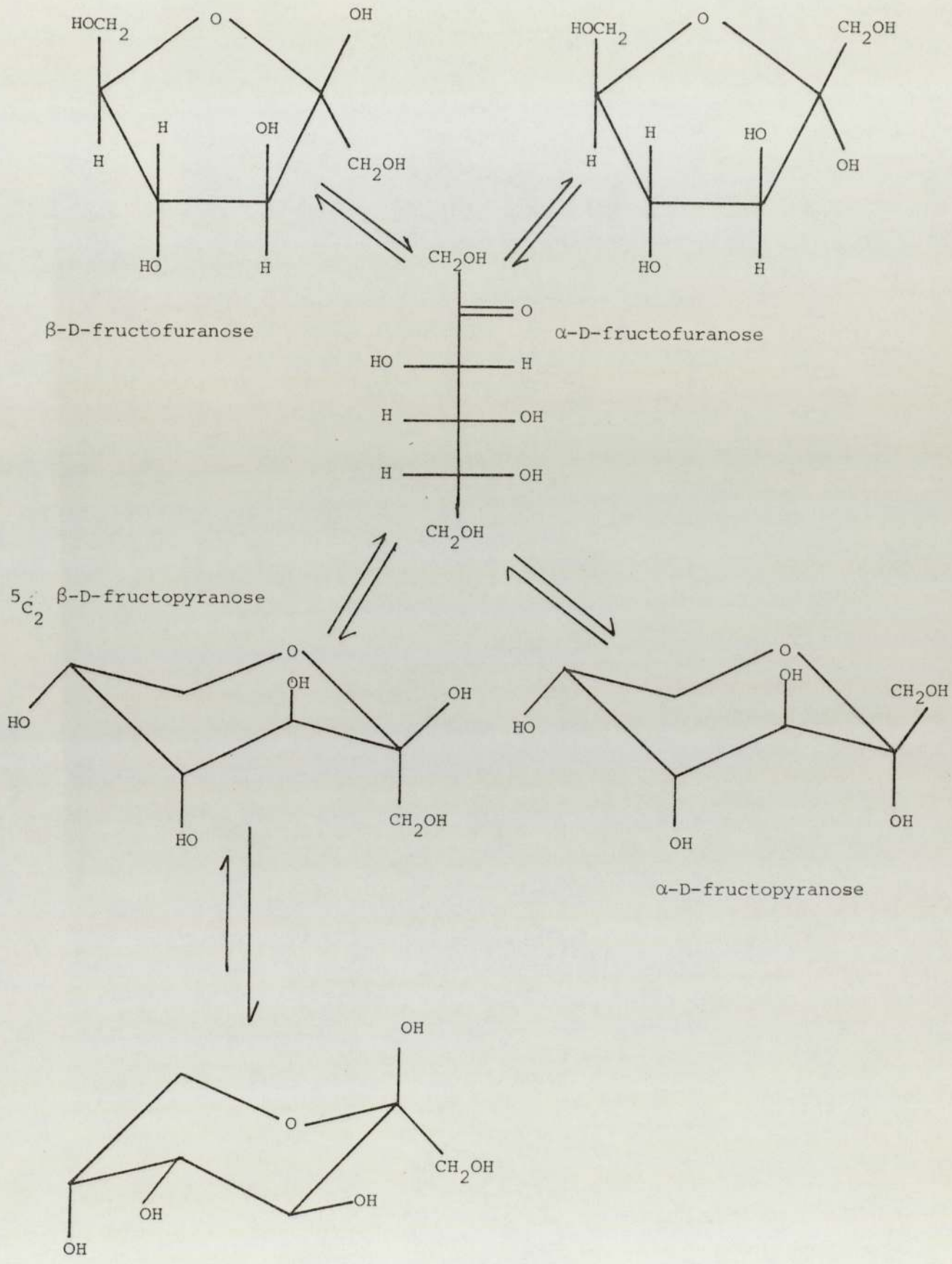


Table 2.1 Equilibrium composition of glucose forms in a molar solution at 25°C	
α-D-glucofuranose	< 1%
β-D-glucofuranose	< 1%
α-D-glucopyranose	35%
β-D-glucopyranose	64%

Table 2.2 Equilibrium composition of forms in a 1 molar solution at 25°C	
α-D-fructofuranose	5%
β-D-fructofuranose	20%
α-D-fructopyranose	0%
β-D-fructopyranose	75%

It is this structure which exhibits the sterically favourable form of the ring, exhibiting the eq-ax-eq-ax arrangement of the oxygen atoms for complexing to occur.

2.2.4 Complex formation

The complex forms when the structurally favourable fructose molecule approaches the hexahydrated calcium ion $[Ca(H_2O)_6]^{2+}$. Since a higher entropy and hence greater thermodynamic stability results if these two units react to form four units, a complex forms and three water molecules are released from around the calcium ion. The available positions are filled by bonds from the hydroxyl groups which are in the ax-eq-ax positions around the fructose ring. The group is known as a chelate complex. It can be represented as shown:

$$C_6H_{12}O_6 + [Ca(H_2O)_6]^{2+} \underset{k_2}{\overset{k_1}{\rightleftharpoons}} [Ca(C_6H_{12}O_6)(H_2O)_6]^{2+} + 3H_2O$$

The equilibrium lies to the right due to the entropy effect resulting from the chelate formation (38). From the law of mass action the equilibrium constant K for the reaction is defined in terms of the concentration of the components as:

$$K = \frac{[Ca(C_6H_{12}O_6)(H_2O)_3^{2+}][H_2O]^3}{[C_6H_{12}O_6][Ca(H_2O)_6^{2+}]} = \frac{k_2}{k_1}$$

The amount of water present must be increased to break the complex and return the fructose to solution. This occurs when a column is purged to remove fructose: a high water flow rate is used and the complex is therefore in contact with more molecules of water per unit time and the equilibrium moves to the left.

2.3 Scaling up the chromatographic process

Chromatography is a technique normally used in the laboratory as an analytical tool. It is inherently a batch process and only minute quantities of materials are generally used. A typical analytical column may be 6 mm in diameter and 250 mm in length. The resolving power exhibited by these analytical columns prompted many workers to attempt a scale up of the process to the preparative then production level. Much of the early work was done using gas chromatographs and generally using a repetitive feed batch technique. More recently liquid chromatography has been studied in greater detail. This section of the survey deals with developments in this field and in particular with continuous chromatographic refining techniques. It tries to identify the physical and operational factors which affect the scale up and reviews methods for minimising if not overcoming them. In some cases findings for batch processes are cited as a practical guideline since studies in continuous chromatography are, as yet, rather limited.

2.3.1 Factors affecting scale-up

2.3.1.1 Flow patterns in large diameter packed columns

The velocity profile of fluid flowing through small diameter - say less than 1" diameter columns may reasonably be assumed to be flat. This means that the solute band travels along the column with a flat uniform front. The concentration therefore at any point along the same radial plane will not vary. As column diameters increase, packing techniques are not good enough nor repeatable enough for a flat profile to be assumed. In his approach to band broadening Giddings (15) has cited five mechanisms by which he believes velocity inequalities may occur (see section 4.4.3).

He indicates that the most significant of these causes is the trans column effect - which is a direct result of the efficiency of the packing.

The deterioration of packing efficiency in larger diameter columns gives rise to an uneven velocity profile, generally parabolic. Various workers have tried to quantify this effect and its contribution to increased plate height. An additional term H_c has been added to the van Deemter equation (eqn. 2.7)

$$H = A + \frac{B}{u} + C_s u + H_c$$

Giddings (39) assumes a parabolic profile and expressed the contribution as

$$H_c = G \frac{r_c^2 u}{\gamma D_m} \qquad 2.12$$

Where G = constant

r_c = column radius

γ = radial labrynth factor

D_m = diffusivity of solute in mobile phase

u = mobile phase velocity.

Results found experimentally (40) using column diameters from 0.6 cm - 5.1 cm gave close agreement with his correlation.

Huyten (41) obtained a good agreement using a 7.5 cm column and similar expressions were proposed by Higgins and Smith (42) and Rijinders (43).

Bayer, Hupe and Mack (44) assumed a velocity profile concave to the direction of flow, caused they suggested by the higher packed density in the central area of the bed. They developed an expression empirically which related the contribution to plate height H_c to the column radius and the mobile phase velocity

$$H_c = 2.83 \frac{r_c^{0.58}}{u^{1.86}} \qquad 2.13$$

Where r_c = column radius

u = mobile phase velocity

This gave very good experimental agreement for columns from 1.3 cm to 10.2 cm diameter.

The preceding expressions predict a decrease in column efficiency as column diameter increases. Pretorious and de Clerk (45) proposed that the plate height was a function of the particle diameter and the column diameter. The resultant profile they developed was a 'W' shape with the maximum velocity several particle diameters into the bed. The plate height contribution H_c was:

$$H_c = \frac{M' \, d_c^2 \, u}{2 \, D_r \, d_p}$$

Where $M' = \frac{1}{100} \, EXP \left(- \frac{d_c}{10 \, d_p}\right)$

d_c = column diameter

u = mobile phase velocity

D_r = radial diffusivity of solute

d_p = average particle diameter

This concept was supported by Spencer (46) and Knox (47). Giddings (48) suggests that if radial equilibrium is not reached, then plate height becomes independent of diameter but the resolution decreases.

2.3.1.2 Concentration Effects

Within an analytical chromatographic column the feed sample is so small ~20 µl compared to the column volume ~10 cm^3 that the elution can be considered to take place at infinite dilution. In preparative chromatography the feed volume and concentration will be far more significant. In the models proposed by van Deemter (14) and Giddings (15) a linear adsorption isotherm is assumed, i.e. the distribution of solute between the two phases is independent of solute concentration. In reality as concentration increases a deviation from ideallity occurs. Helfferich (49) redefined the retention volume equation to account for this - cf eqn. 2.1.

$$V_R = V_M + V_S \frac{\partial q}{\partial c} \qquad 2.14$$

where q = concentration of solute in stationary phase

c = concentration of solute in mobile phase.

The curve representing this function which applies at constant temperature is the adsorption isotherm. Three main types of isotherms may be classified by their different effects on column performance. Vermulen (50) describes favourable and unfavourable isotherms for resolution. The slope of a favourable isotherm is a decreasing

function of c (see Fig. 2.9a) whereas an unfavourable isotherm is an increasing function of c, Fig. 2.9b. The intermediate or linear isotherm is represented by a dashed line.

When a 'favourable' isotherm is exhibited both the retention volume and the K_d increase with an increase in feed concentration, with an unfavourable isotherm both parameters decrease. Operation in the non linear region requires extra column length to compensate for the decrease in resolution. In production scale operation a degree of contamination of products is tolerated for the increase in throughput gained by higher feed concentration. This will however be accompanied by an increase in viscosity and thus higher pressure drops. A compromise is often reached after an economic assessment.

2.3.1.3 Increase in mobile phase velocity and column length

When designing production scale equipment throughput plays an important rôle. To increase the throughput in a chromatograph the mobile phase velocity is increased. This causes a shorter residence time and an increase in plate height as the columns' ability to resolve two or more components decreases. To maintain the same degree of resolution for a higher throughput a longer column is necessary. This in turn results in a longer elution time. Generally a compromise is reached between throughput and product purity in conjunction with pressure drop and pumping energy.

2.4 Practical Solutions to Production Scale Chromatography

2.4.1 Column Packing

As mentioned in section 2.3.1.1 poor packing techniques

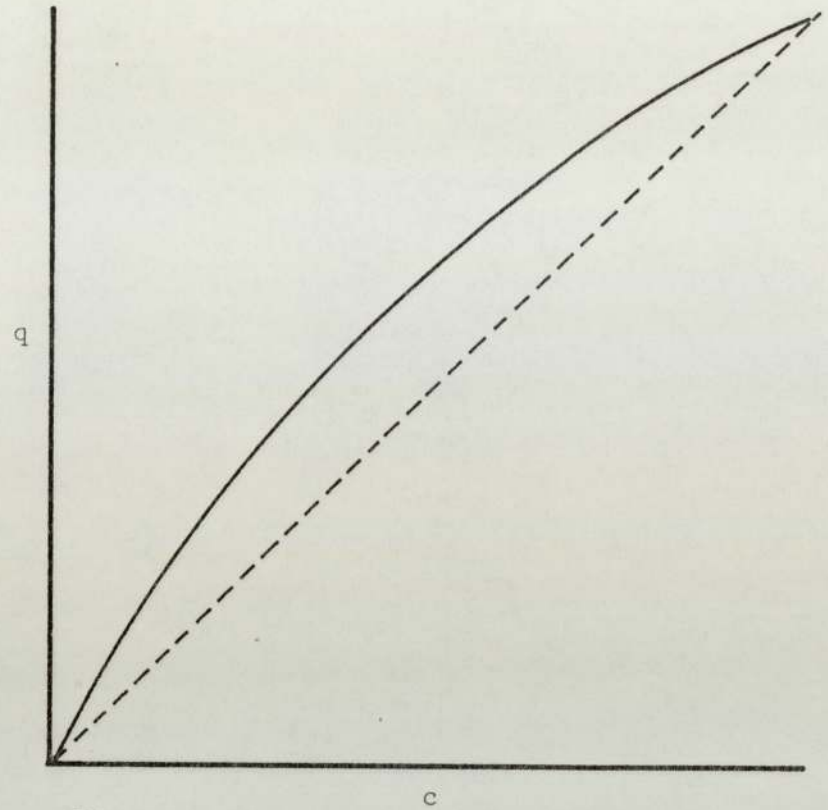


FIG. 2.9a FAVOURABLE ADSORPTION ISOTHERM (LANGMUIR)

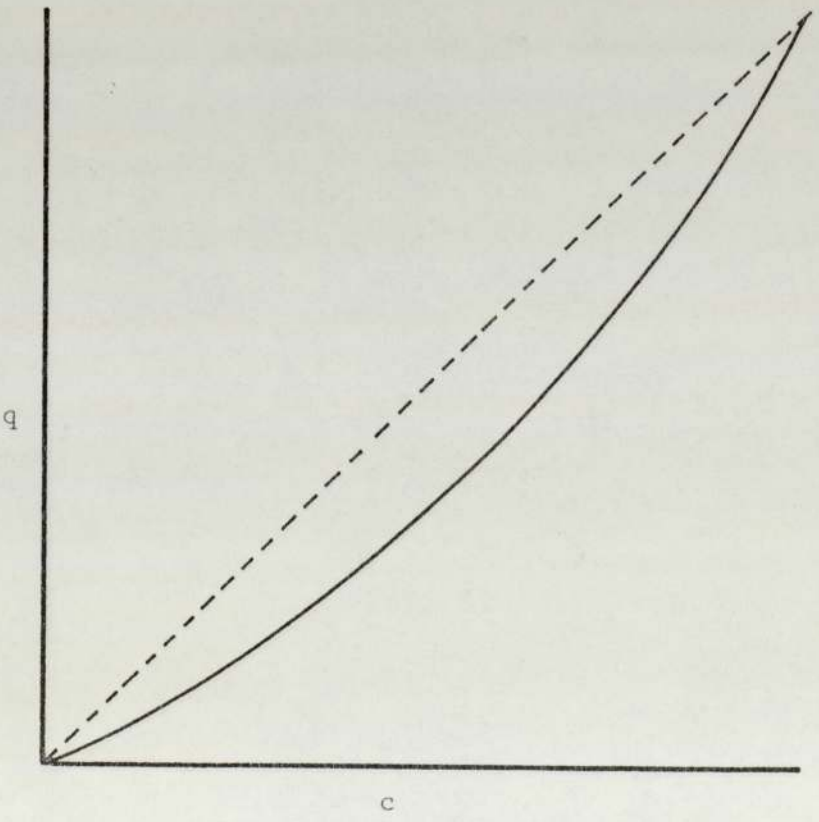


FIG. 2.9b UNFAVOURABLE ADSORPTION ISOTHERM (ANTI LANGMUIR)

have resulted in inefficient operation. A high efficiency, repeatable packing technique is therefore very desirable. This should exhibit a uniform packed density with no segregation of particles. This suggests a narrow particle size range.

Various techniques have been used to achieve these aims. Two basic methods have been used: dry packing and slurry or wet packing. Higgins and Smith (42) have reported 'bulk', 'snow' and 'mountain' techniques of dry packing - concluding that the 'mountain' method which was gradual addition of the dry material to the column gave the most favourable HETP values.

Fluidisation techniques have been reported by Guillemin (51) which produced a higher efficiency initially but the packed beds were prone to collapse with use. The classical 'shake turn and press' method was developed by Verzerle (52) who achieved good repeatability with dry packing.

Slurry techniques include bulk pouring, pouring with external vibration, pouring under vacuum and reservoir packing. In slurry packing techniques the bed is usually formed by sedimentation of a thick suspension. In such a suspension particles do not behave individually according to Stokes' Law, but interact with each other and the interparticle fluid. Thus they tend to flow downwards as a unit in some places forcing less concentrated suspension upwards. Thus segregation occurs. The best method is often found by trial and error for the particular column diameter and packing. The slurry packing with vacuum was the most repeatable and efficient for the zerolit 225 resin used in the SCCR equipment and was adopted as the packing technique for this research. A complete review of the packing techniques available has been published by Spencer and de Clerk (53).

2.4.2 The use of flow distributors

Even with an acceptable packing technique liquid generally enters a column from a narrow tube or pipe (to minimise remixing or backmixing). The diameter of this tube is generally far less than that of the chromatographic column. To create a uniform velocity profile throughout the bed area a flow distributor is necessary particularly on large diameter columns.

Various devices have been used. Musser and Sporks (54) investigated the use of inlet cones in gas chromatographic columns and found a significant increase in column efficiency. Other devices used have included perforated plates and screens. These provide a uniform resistance and thus a uniform profile results. It has been shown (55) that the packed bed itself may not create a uniform velocity profile until four or five bed diameters into the column have been traversed. Thus distributors play an important part in full bed usage. With long columns flow redistribution has proved necessary. Baddour (56) used a 'disc and doughnut' arrangement with the disc of a smaller diameter than the column forcing the mobile phase towards the column walls. When the fluid strikes the doughnut it it redistributed towards the central region. Abcor Inc. Massachusetts (57) report improved column efficiency using such baffles on large diameter columns.

2.4.3 The use of repeated feed injection

The analytical chromatographic column is injected with a slug of feed which is eluted through the column. During the elution, although some band broadening occurs only a small part of the bed length is being used to effect the components' separation at any one

time. The next step in the move to production scale chromatography was the repeated feed injection. This is a technique where feed slugs are regularly injected at set time intervals. This interval is calculated such that the leading edge of an injected feed sample elutes from the column just as the trailing edge of the previous injected sample elutes. Thus the whole column length is being used for resolution at all times, see Fig. 2.10. These fractions may then be collected separately and if contamination has occurred some may be recycled. Conder (58) has suggested that it is preferable to 'overlap' components with greater throughput rather than avoid the need for recycling by increasing the column length. He suggests a 60% recovery of the injected feed as pure product is an optimum with the contaminated 40% being recycled.

2.4.4 Continuous chromatography

So far, the scale up process has been taken from the small diameter analytical columns to larger diameter columns using the same batch mode. To make use of all available bed length for separation the repeated feed injection cycle was employed. The next step was to use a truly continuous system in which the feed entered continuously and two product streams were removed continuously as for example in a distillation column. To remove two products continuously most workers attempted to cause the stationary phase and the mobile phase to move counter-current to each other. Fig. 2.11 illustrates the type of concentration profile which would evolve if a two-component feed were separated using continuous counter-current chromatography.

FIG. 2.10 ILLUSTRATION OF REPETITIVE INJECTION TECHNIQUE FOR A TWO COMPONENT SYSTEM

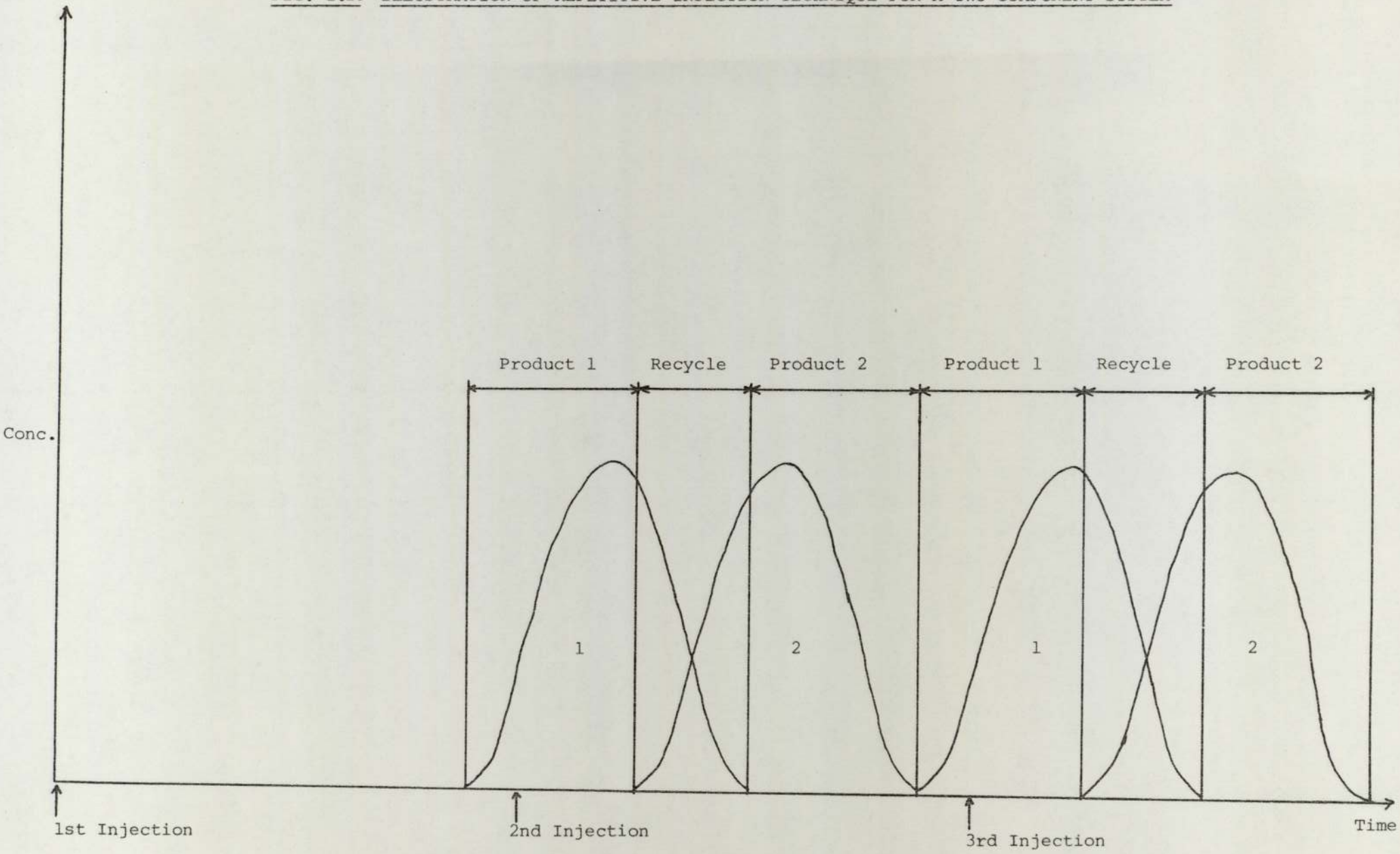
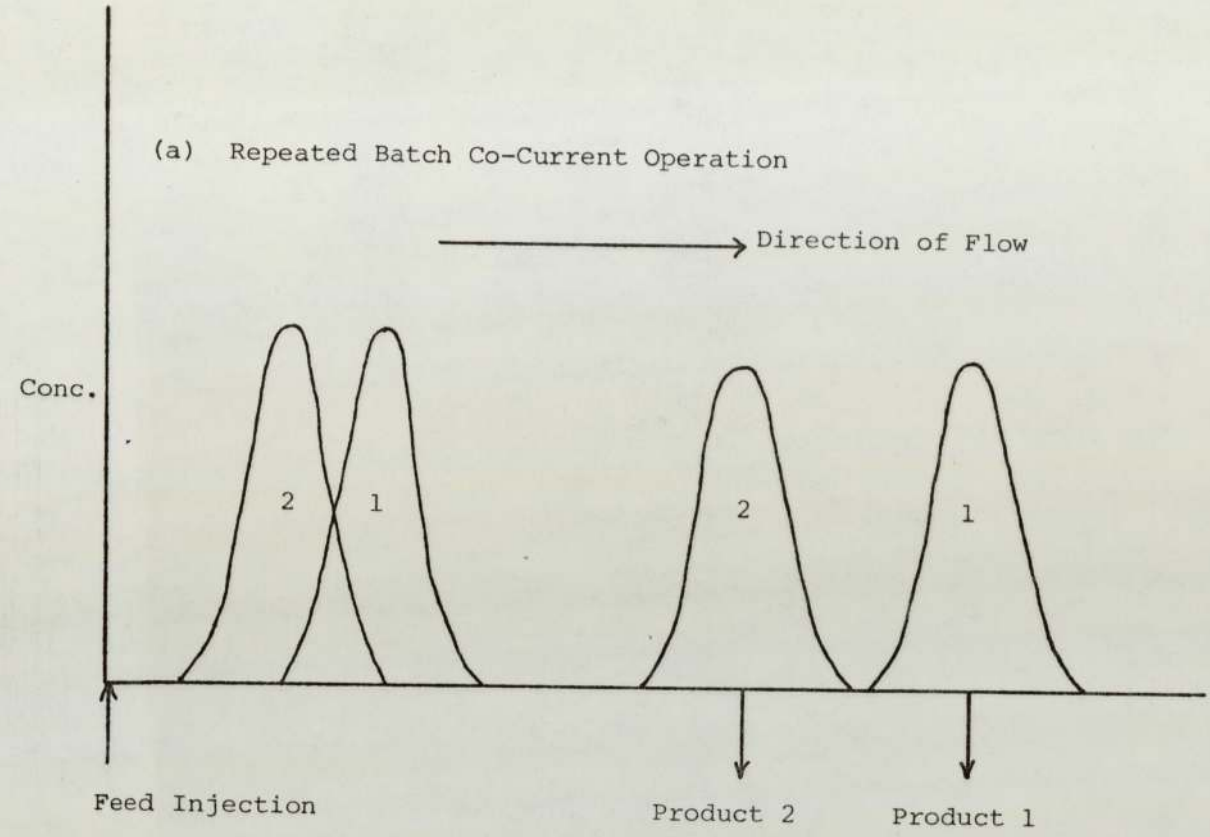
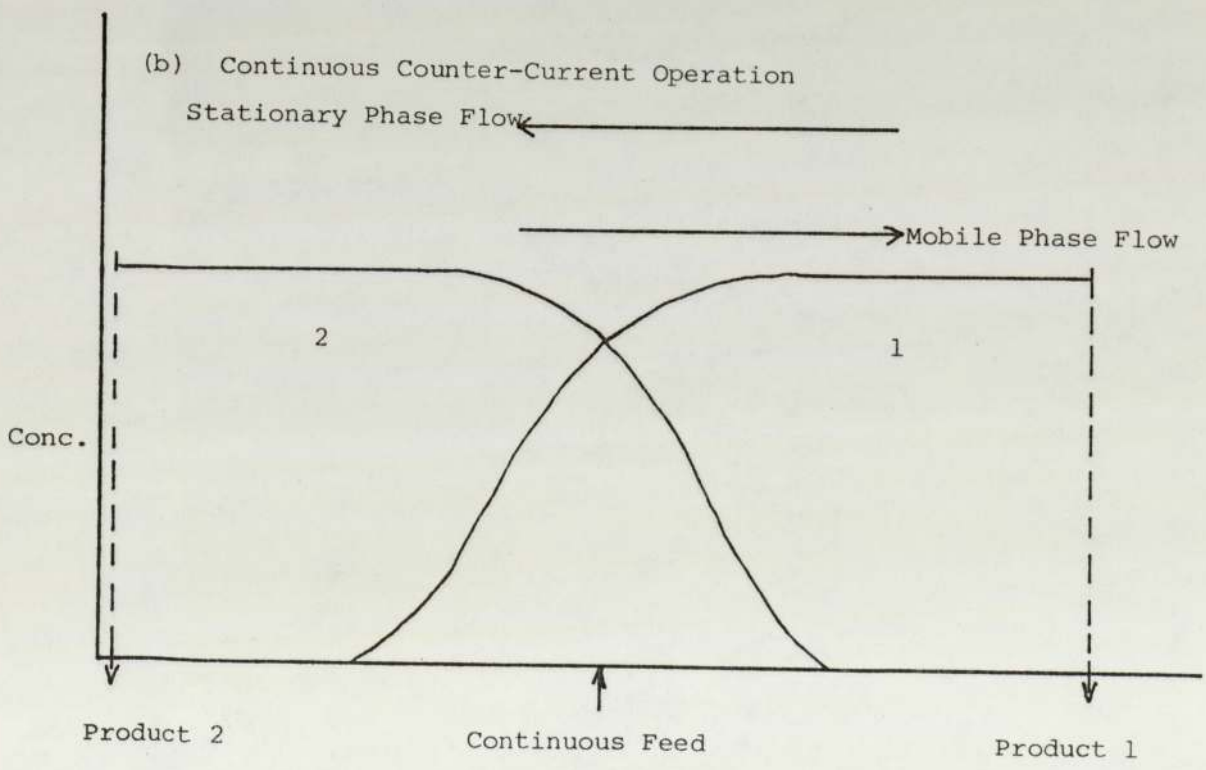


FIG. 2.11 CHROMATOGRAPHIC CONCENTRATION PROFILES OBTAINED FOR THE SEPARATION OF TWO COMPONENTS





Believing that greater throughputs and greater purity would result if this technique were applied many attempts were made to design chromatographs which would achieve this end. Three basic systems were developed: fixed bed, moving bed and simulated moving bed.

2.4.4.1 Fixed bed operation

Early work concentrated on GLC or GSC. Tilley (59-61) used a 2.5 cm column with a knitmesh packing. The liquid stationary phase flowed downward over the packing against the mobile phase gas flowing upwards. High product purities were obtained but low throughputs.

A similar scheme was used by Kuln (62,63) who exploited the effect temperature on the distribution coefficients. This was achieved by imposing a temperature gradient on to the column. Purities of 95% were achieved, but throughput was low.

Pigford (64,65) used temperature cycling on a fixed bed to achieve separation. In his studies a constant flow mobile phase gas was passed through the bed. The temperature cycling effect was achieved by the continuous introduction of the feed which was alternately heated and cooled.

Extensive studies have been undertaken using this technique by Wankat (66,67) and a survey of similar attempts is included in his review (67).

2.4.4.2 Moving bed systems

This group of developments may be divided into those in which the stationary phase physically moves counter-currently to

the mobile phase and those in which columns of stationary phase are made to move in a counter-current direction to the mobile phase.

The continuous movement of the stationary and mobile phases in opposing directions is the simplest mode of counter-current operation. Barker (68-70) and co-workers used a 2.5 cm diameter column which was 2.74 m in length. The packing was coated with the liquid stationary phase and allowed to flow under gravity against the rising mobile phase gas. The flow of packing was regulated by a variable orifice at the base of the column and as the packing was removed it was recycled to a hopper above the column. External vibration ensured a constant flow of packing. The least retained component was eluted in the mobile phase gas and the more retained component travelled with the packing. This was stripped off by a heated fresh gas stream at the column base.

Other researchers, Schultz (71), Scott (72) and Tilley (59) have reported successful operation using the same principle.

Moving bed systems had inherent disadvantages in that large quantities of solid packing need to be handled. This led to attrition of the particles. The flow rate of the mobile phase was also limited to a value below that of the minimum fluidising velocity of the particles. Physically as the bed moved down the column, packing characteristics varied and poor efficiency through backmixing often resulted.

In attempts to avoid these problems the moving column system was developed. This involved the rotation of a series of columns past fixed inlet and outlet ports against the direction of the mobile phase flow. Pickler and Schultz (73), Luft (74) and

Glasser (75) designed and operated equipment which allowed the mobile phase to enter the columns and then flow in either direction through a circular series of columns. With suitable outlet positions reasonable separations were achieved in terms of purity.

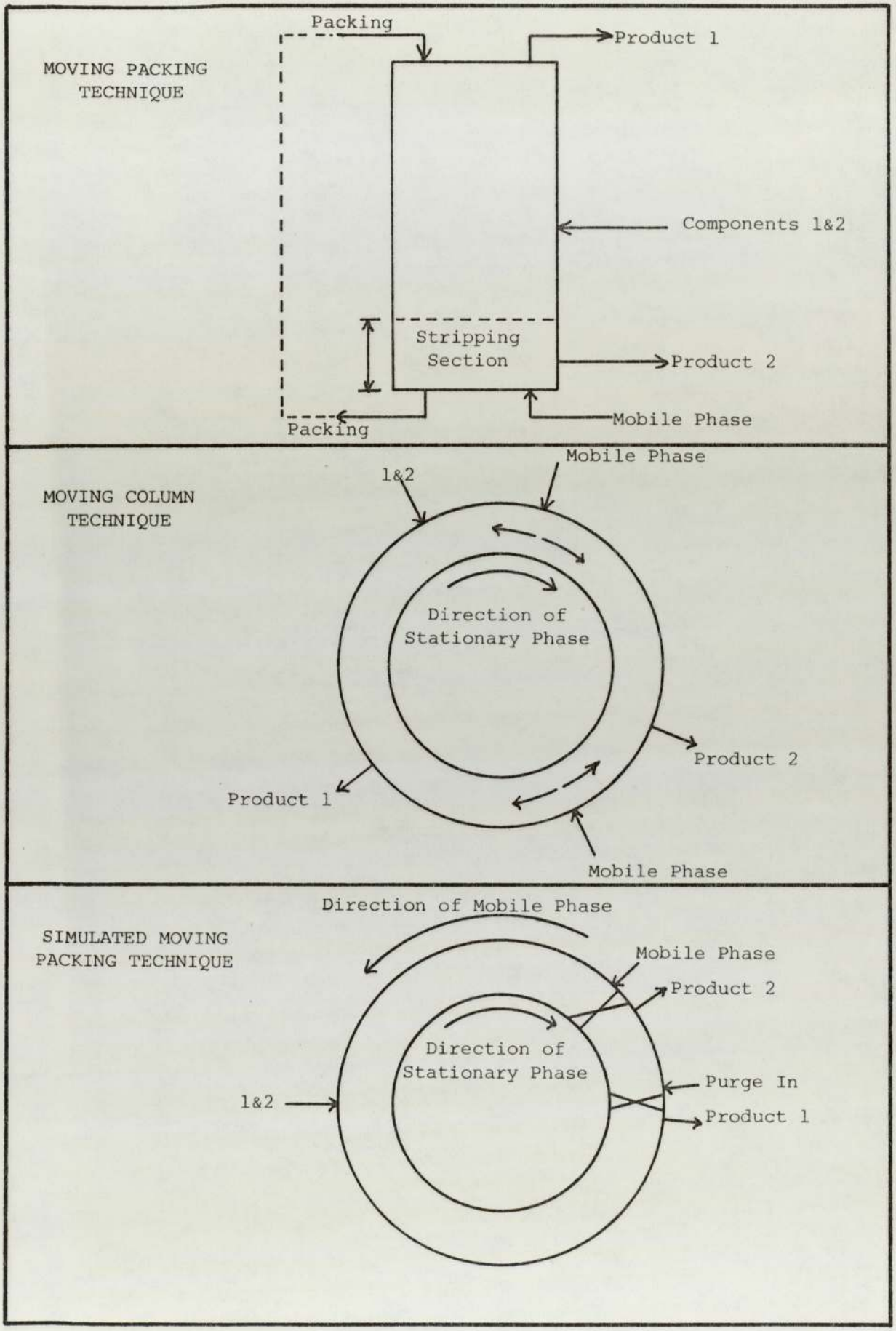
Barker (76) used valves to ensure that the mobile phase gas travelled in one direction only. This allowed more of the column to be used for separation. This was further developed by Barker and Al Madfi (77) by forming a torroid of columns each linked via a transfer valve, see Fig. 2.12.

To enable a more compact apparatus to be built Barker (76) and the Universal Fisher Group replaced the torroid with a tube bundle of 44 tubes arranged between two stainless steel annular rings. Spring loaded 'Graphlon' (PTFE-carbon composite) face seals were used to prevent leakage from the unit when the valves were open. The separations performed on this equipment (78-79) were better than the previous torroid due to the extra mass transfer length available. A more recent system developed by Barker, Hatt and Knoechelman (80) replaced the vertical bundle of columns by individual U-shaped columns. Hence the lower sealing face was eliminated. Fluid transfer between columns was achieved through holes in the upper ring.

2.4.4.3 Simulated moving bed counter-current systems

Moving seal problems incurred with the above systems encouraged the development of a simulated moving bed design. Szepesey (81) proposed a scheme in which a switching valve was centrally mounted on a rotary PTFE disc. Rotation of the valve altered the relative position of the inlet and outlet ports to a stationary series of columns. In such a manner counter-current

FIG. 2.12 CONTINUOUS FLOW SCHEMES



column movement is simulated.

The Szepesey design still relied upon the rotary seal. Barker and Deeble (82) designed an apparatus where all the moving parts were eliminated except the opening and closing of inlet and outlet valves which directed the various flows. Electrically operated solenoid valves were used in this SCCR. In this equipment twelve 7-6 cm diameter columns each 61 cm long were connected together to form a closed loop. Studies on this unit using binary halocarbon feed mixtures have proved successful with feed flow rates of up to 1400 $cm^3 hr^{-1}$ being used.

Ching (6) used a similar flow scheme in constructing a liquid-solid chromatograph for carbohydrate separations. Ten 2.54 cm diameter columns each 65 cm long were used to form the interconnected loop.

The apparatus designed in this research project uses a similar flow scheme and will be discussed in detail later.

CHAPTER 3

THE DESIGN AND CONSTRUCTION OF THE EQUIPMENT

3 Introduction

A semi-continuous chromatographic refiner had been built and commissioned by Ching (6) then modified and operated by Chuah (17). The packed columns were 2.54 cm in diameter. To assess scale-up effects in terms of throughput, product purity and product concentrations it was decided to design and construct a unit with columns whose diameter were 10.8 cm. This was the largest diameter column that the budget would allow based on material costs and resin costs to fill the columns. It was desirable that the rig should have the ability to operate in the batch mode as well as the continuous mode. As the columns were to be filled with an ion exchange resin, materials of construction were chosen so that no ionic contamination would occur.

3.1 Principles of Operation

It was found (1) that if a slug of feed mixture of fructose and glucose was eluted through a column with a packed bed of an ion exchange resin charged in the calcium form, separation of the sugars occurred. This happened because of a chemical complex which forms between the calcium ions and fructose molecules which retards the passage of this sugar through the column (see section 2.2.1).

Thus if the concentration of the eluted product is monitored after the injection of the feed slug comprising a 50% w/w mixture of the two sugars fructose and glucose, a chromatogram as shown in Fig. 2.11a results. If the stationary phase and the mobile phase were moved counter-current to each other at suitable flow rates than a concentration profile similar to that shown in Fig. 2.11b would result.

3.2 Method of Operation

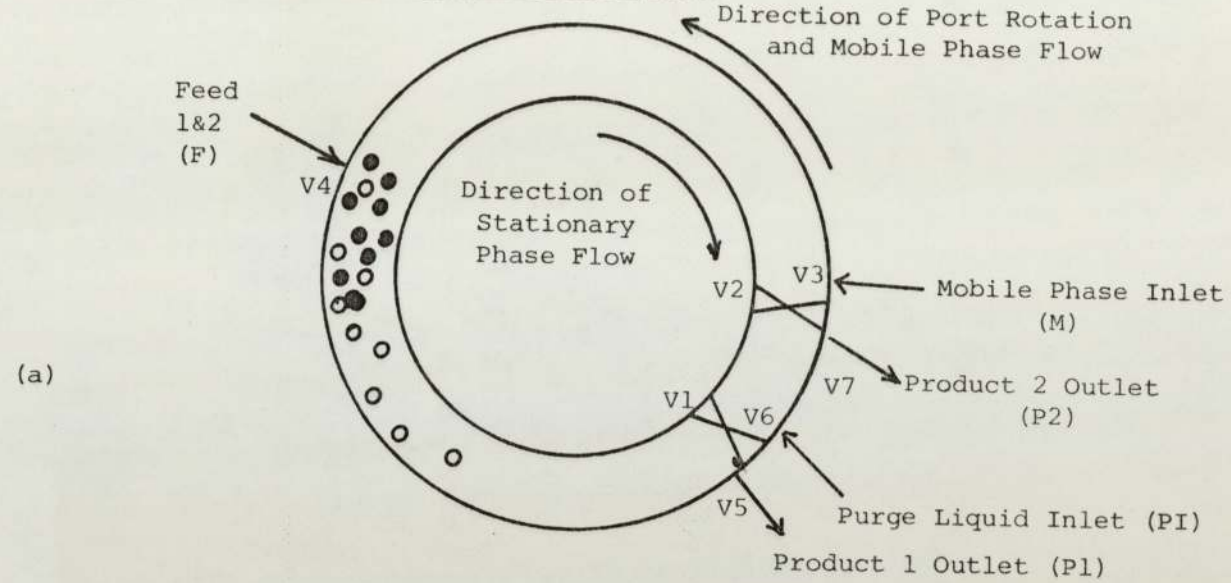
The separating section may be considered as a closed loop. Consider the start up operation (see Fig. 3.1a).

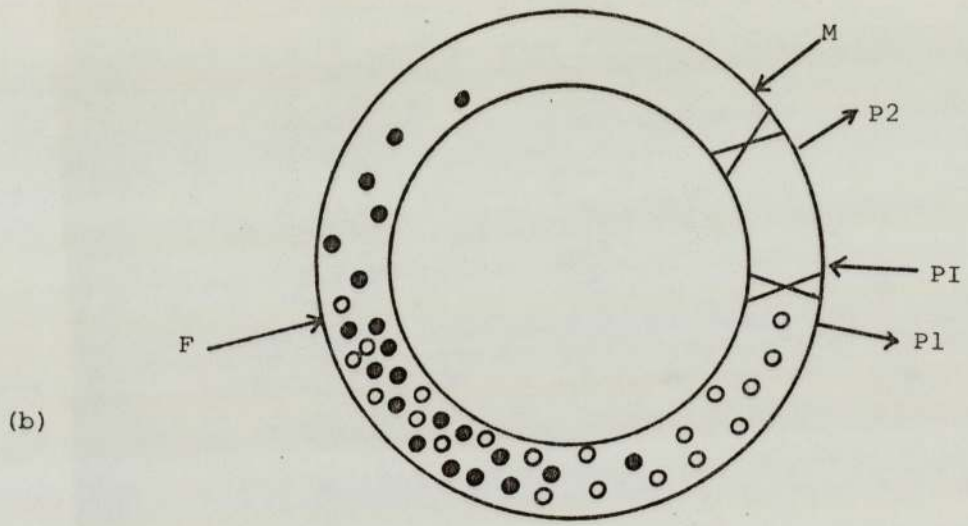
A portion of the loop is isolated by closing two valves V1 and V2. The mobile phase enters the loop through valve V3 immediately adjacent to the isolated section. The feed mixture enters the loop through valve V4. The glucose molecules tend to travel with the mobile phase and hence migrate through the column faster. If after a convenient 'switch period' all the liquid entry and exit points are advanced one position (or switch) as in Fig. 3.1b the glucose is seen to emerge in the eluent exit stream through valve V5 and fructose tends to migrate with the stationary phase towards the isolated section. If the rate of this port advancement is less than the velocity of glucose migration through the bed but greater than the velocity of the fructose through the bed two enriched products will emerge from each end of the separating section, Fig. 3.1c.

To remove the fructose from the isolated section of the packing the loose chemical complex is broken by the passage of an independent higher flow rate purge stream of water. This enters through valve V6 and leaves this section through valve V7. This not only removes the fructose but prepares the packing to receive the advancing glucose rich mobile phase when the next port advancement occurs.

Figs. 3.2a,b and c demonstrate how the simulated counter-current mode may be achieved without physically moving the packing and the SCCR was designed on these principles. Fig. 3.2 shows three successive switch periods to demonstrate the flow schemes.

FIG. 3.1 PRINCIPLE OF OPERATION OF THE SCCR





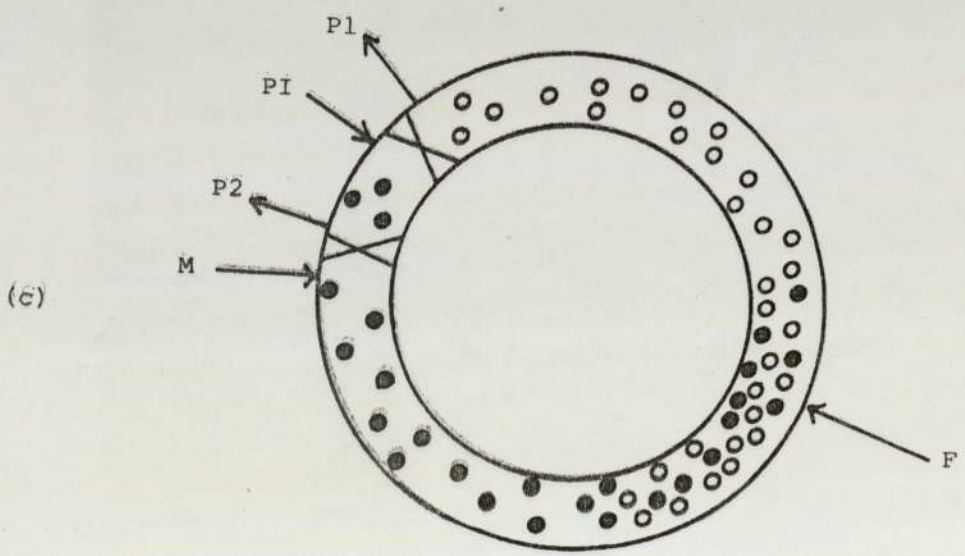
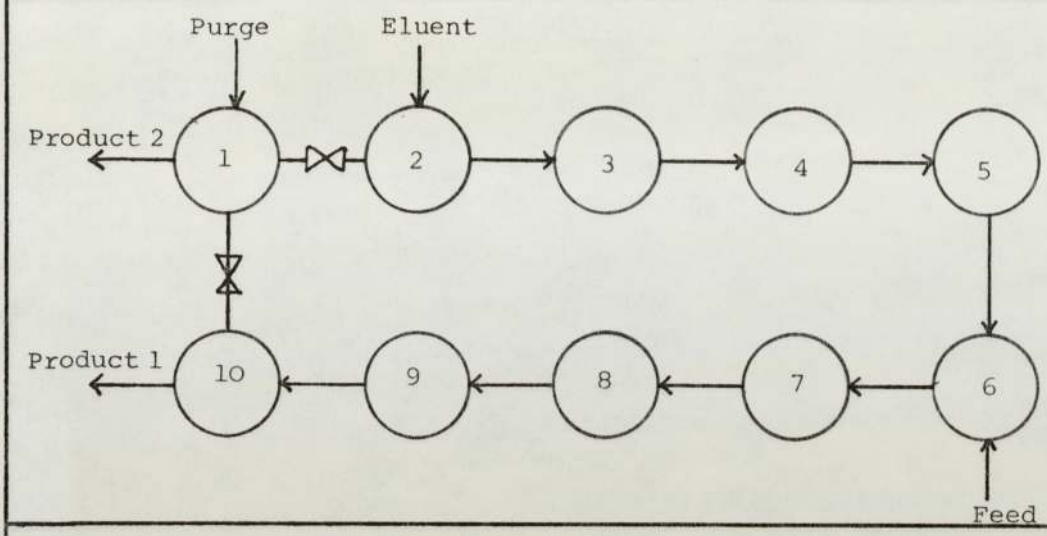
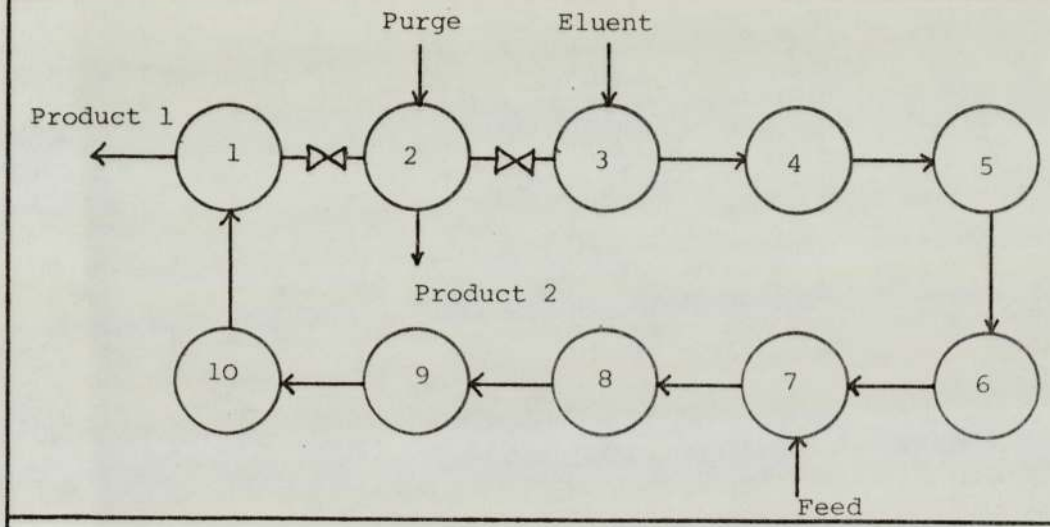


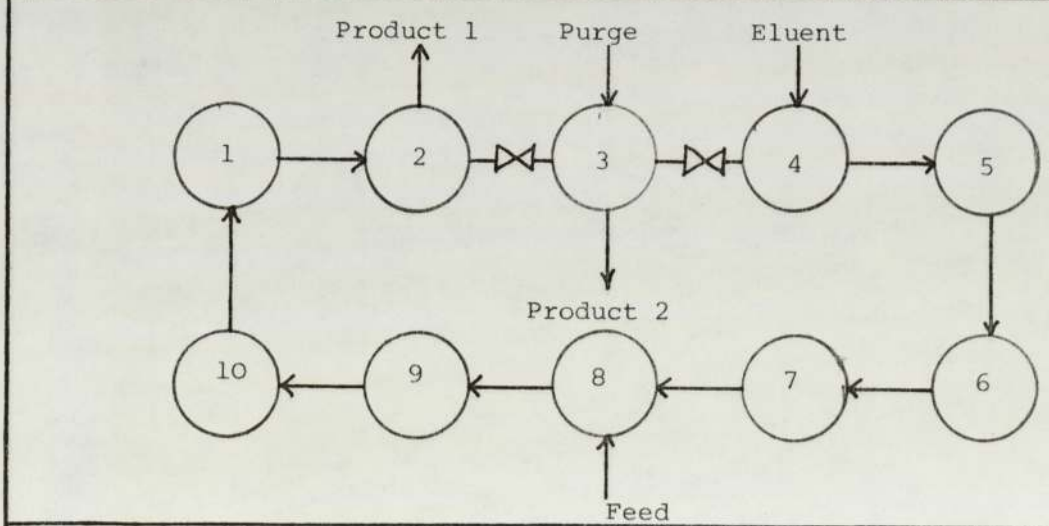
FIG. 3.2 DIAGRAMS SHOWING SEQUENCES OF OPERATION OF THE SCCR



Switch One: Column 1 Isolated Columns 2-10 Fractionating



Switch Two: Column 2 Isolated Columns 3-1 Fractionating



Switch Three: Column 3 Isolated Columns 4-2 Fractionating

3.3 The chromatographic refiner: overall view

Fig. 3.3 is a photograph of the apparatus. The separating section is a series of ten discrete packed columns linked at the top and bottom to form a closed loop. From Fig. 3.2 it can be seen that during any particular switch period seven valves will be activated, five to open namely: eluent inlet, feed inlet, purge inlet, glucose product outlet and fructose product outlet. Two valves will be activated to close namely the transfer valves at each end of the isolated column. These valves are double acting poppet valves energised by compressed air from the central control unit. Fig. 3.4 is a flow diagram of the complete system.

3.4 Detailed design and construction

3.4.1 The columns

As data already existed on the performance of the refiner built by Ching (6) it was decided to maintain approximately the same bed length i.e. 65 cm per column giving a total separating length of 6.5 m. Stainless steel was selected as the material of construction for two reasons: it would not contaminate the resin ionically and it would allow higher working pressures. Consequently seamless stainless steel pipes of length 76 cm and internal bore of 10.8 cm were used for each of the ten columns.

Sampling points constructed from stainless steel were welded to the sides of the columns radially and longitudinally. These were threaded and a silicone rubber septum was kept in place over the end by a $\frac{1}{8}$" B.S.P. simplifix nut. These septa allowed a hypodermic needle to be inserted through them and a small 'on column' fluid sample to be withdrawn into a syringe. This would allow on

FIG. 3.3 PHOTOGRAPH OF THE SCCR

A = PACKED COLUMN WITH SAMPLE POINTS

B = VALVES

C = CONTROL BOX

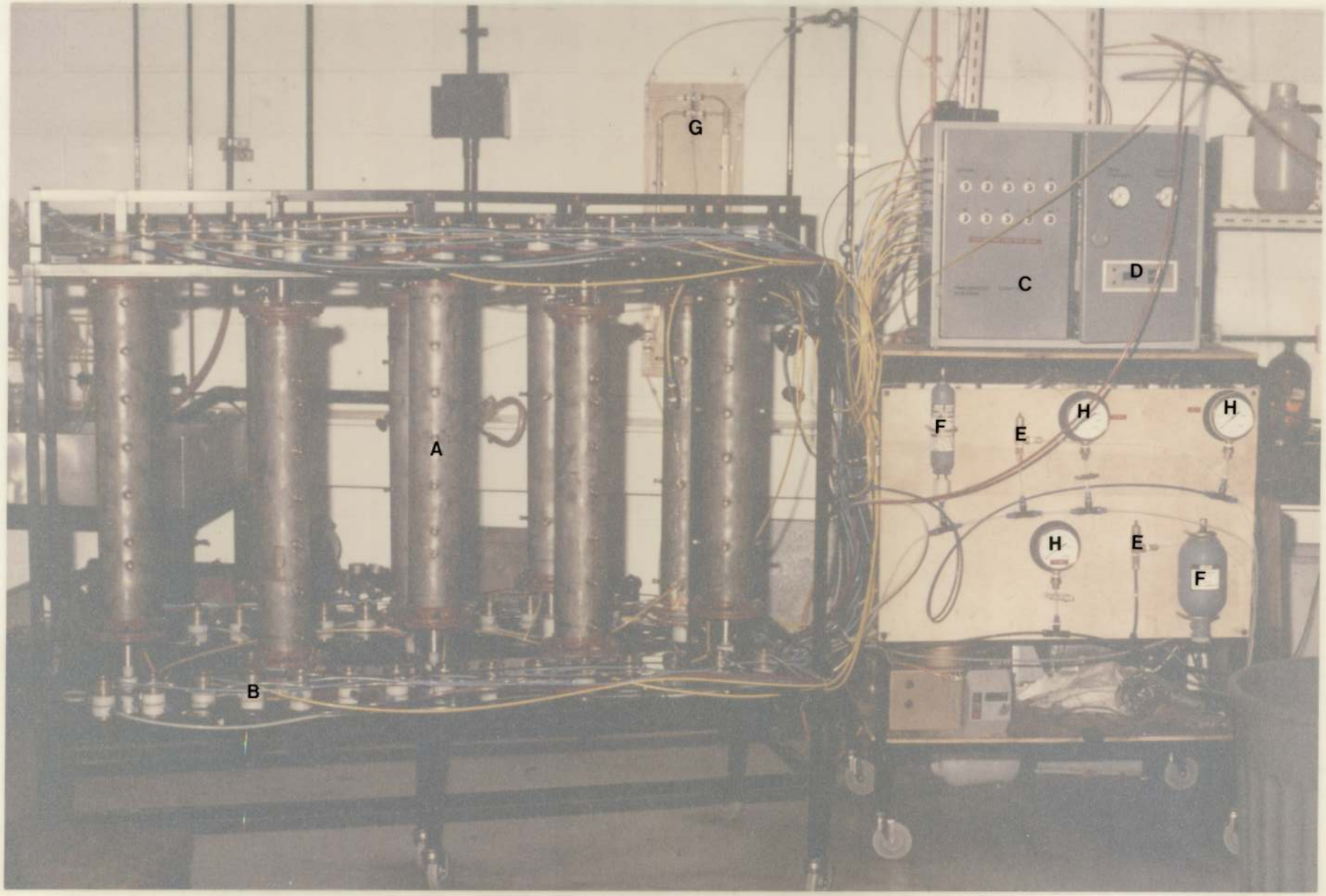
D = DIGITAL TIMER

E = PRESSURE RELIEF VALVES

F = PULSATION DAMPENERS

G = FLOW MEASURING DEVICE

H = PRESSURE GAUGES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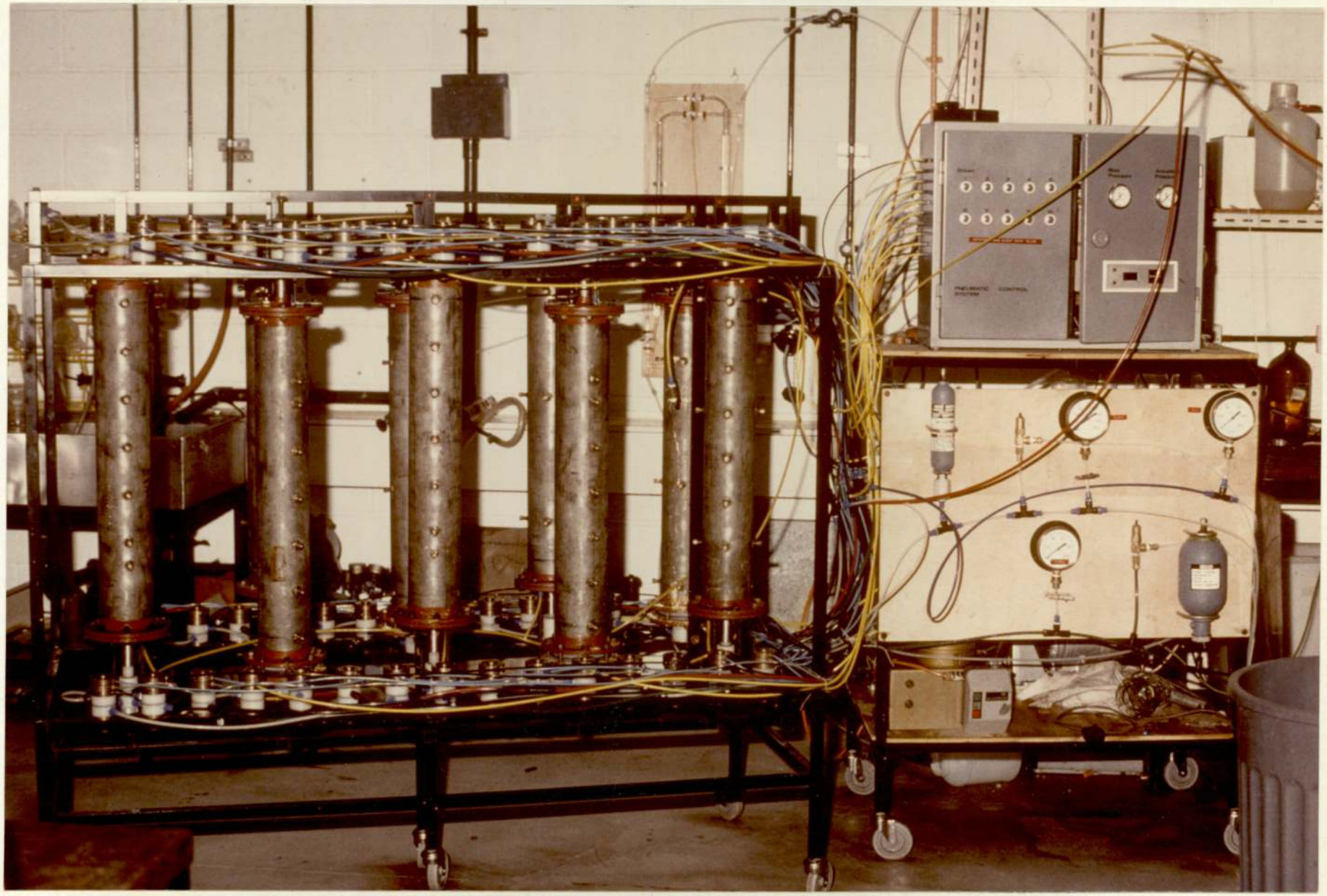
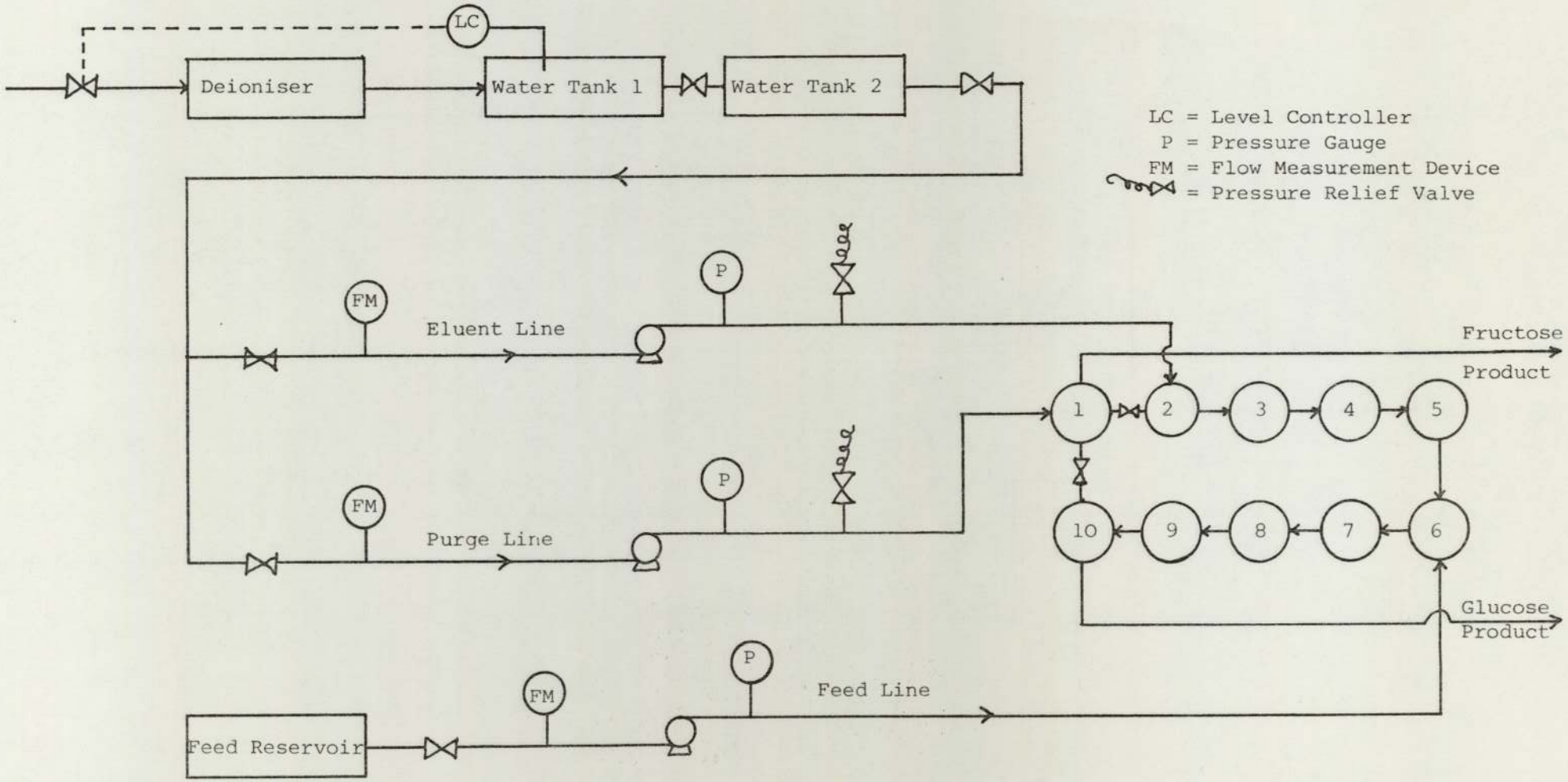


FIG. 3.4 COMPLETE FLOW DIAGRAM



column concentration profiles to be constructed from the data obtained.

Mild steel flanges were welded to the top and bottom of each column to allow inlet and outlet assemblies to be bolted into place. The outside diameter of each flange was 17.8 cm with 8 x 9.5 mm ($\frac{3}{8}$") bolt holes evenly spaced on a pitch circle of diameter 14.5 cm.

3.4.2 The column inlets

To counteract any packing expansion or contraction it was decided to have a floating end. This would allow a slight positive pressure to be kept on the beds at all times. A hydraulically operated piston constructed from polypropylene was designed to achieve this. The seal between the hydraulic fluid, which was deionised water and the packed bed was effected by rubber 'O' rings supplied by the Dowty Company from Cheltenham, Glos. These rings sealed on an accurately machined surface at the top of each column.

Process liquids enter the column through the piston via an inlet manifold and a connecting rod attached to it. To allow the necessary movement of the piston and still maintain the hydraulic seals, the connecting rod was also fitted with 'O' rings which sealed in a smaller cylinder welded to the top plate of each column, see Fig. 3.5. The liquid passageways in all the polypropylene fittings were 5 mm diameter.

It has been shown (83,84,85) that a uniform velocity profile through a packed bed is desirable for most efficient operation. As the column internal diameter is 10.8 cm and the liquid inlet duct 5 mm a distributor was included in the design. It is a removable

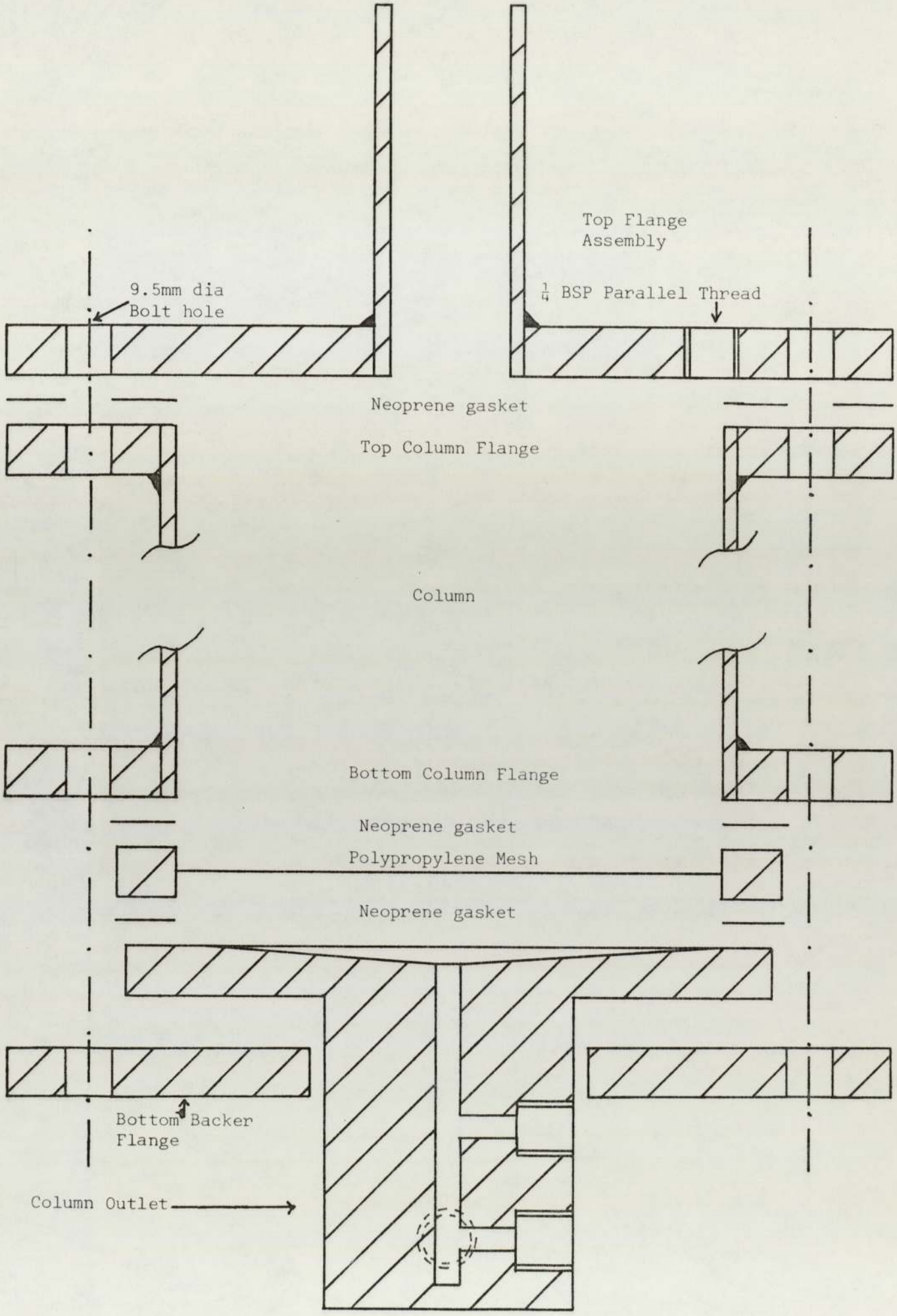


FIG. 3.5 SECTION THROUGH COLUMN ASSEMBLY

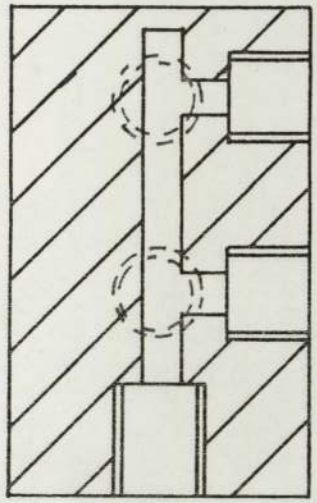
plate that fits into a recess on the underside of the piston (see Fig. 3.6). There is a 1 mm gap behind the plate to allow a uniform liquid reservoir. To prevent the packing from fouling the distributor a polypropylene mesh of 100 μm is fitted over it and kept in place by a polypropylene retaining ring and 8 x 6BA nylon screws. The holes in the plate are 3 mm in diameter and arranged on a square pitch of 1 cm.

3.4.3 The column outlets

To retain the packing in the column a polypropylene mesh of 100 μm aperture was welded between 2 polypropylene annuli (ID = 11.2 cm, OD = 13.3 cm) a coarser polypropylene mesh (1500 μm) was welded into the ring to give lateral support. The device was clamped between the bottom flange of the column and the outlet assembly block. This was also manufactured from polypropylene and had three outlet ports, see Fig. 3.5. To allow an easier escape of liquid from the bed through these outlets a 3^{o} cone was turned on the inside surface of the outlet block. These end fittings were kept in place by a mild steel backing plate which was bolted to the column flange. Sealing was effected by neoprene rubber gaskets.

3.4.4 The valves

Following the successful use of the double acting poppet valves by Ching (6) and Chuah (17) it was concluded after experimentation, using flow rate and pressure drop measurements, that with a slight enlargement of the inlet and outlet ducts of the valves, they would be suitable for use on the SCCR. Sixty valves were fabricated by Aston Technical Services Ltd. to the modified design used previously (6).



Inlet Manifold

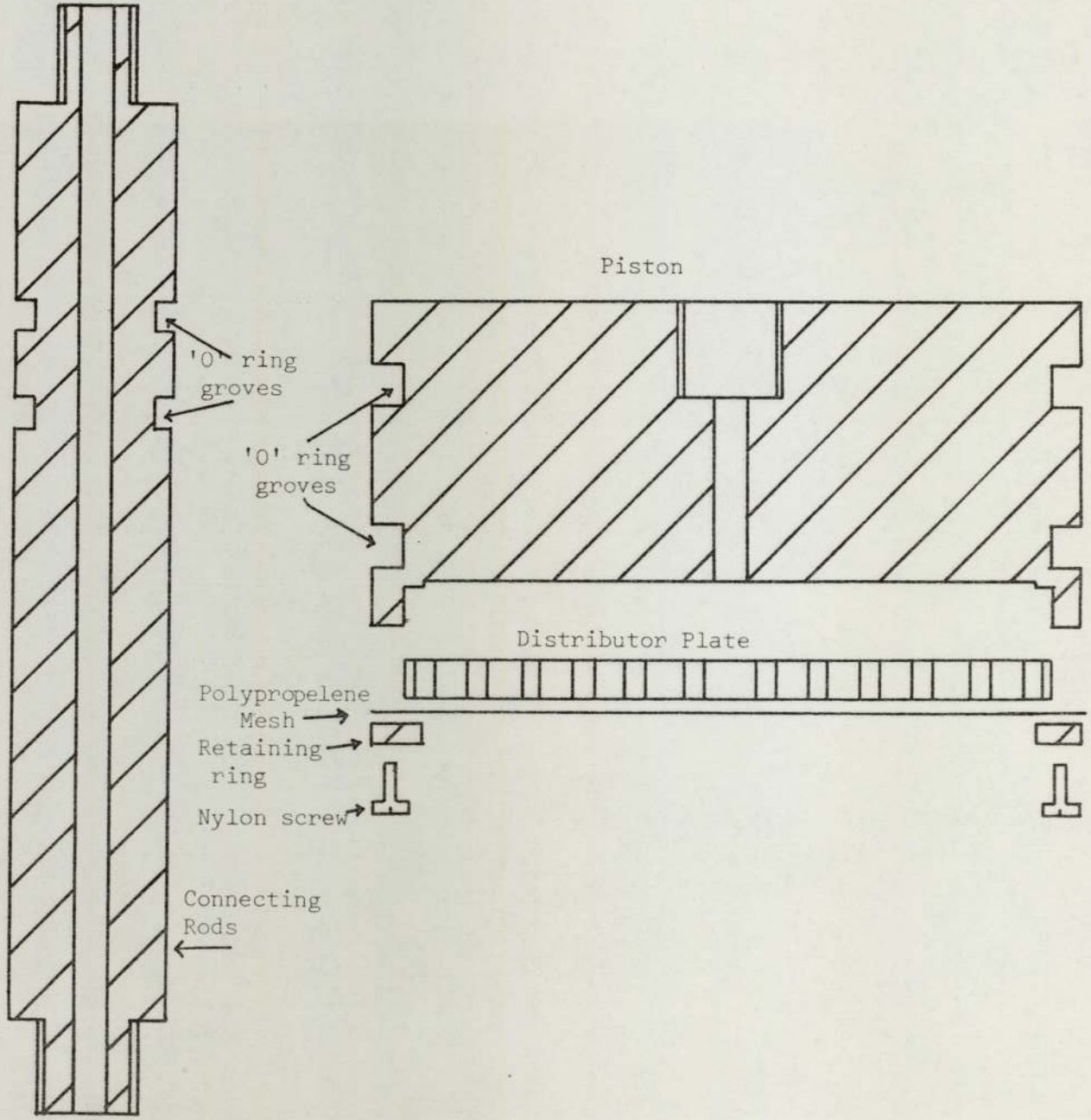


FIG. 3.6 SECTION THROUGH INLET ASSEMBLY

The double acting valves are operated pneumatically and are either open or closed. The moving component is the poppet valve (Part 10, Fig. 3.7) which in its open position (poppet lowered) allows fluid to flow past it through an annulus to the exit. In its closed position the poppet seats against the surface of the valve body and prevents flow. This movement is achieved by attaching a neoprene rubber diaphragm to the stem of the poppet. A bias pressure (constantly applied) acts on one side of this diaphragm and when the actuating pressure is applied which is higher than the bias pressure, the pressure differential causes the poppet-movement. The columns, valves and all the pipework were assembled on a mobile framework as shown in Fig. 3.3.

3.4.5 The control system

Compressed air is supplied to the refiner by a Broom and Wade portable compressor. This supply at 540 kNm^{-2} is divided into two streams namely the actuating and bias streams whose pressures are controlled using Spirax pressure regulators. The actuating air pressure is 540 kNm^{-2} and the bias air pressure is 270 kNm^{-2} - thus a differential pressure of 270 kNm^{-2} is obtained. The bias supply is linked directly to the bias side of the valves' diaphragm via a ring main and this pressure is maintained continuously throughout the entire operation of the refiner. The actuating air supply is fed to the central controlling unit, see Fig. 3.3, and this is relayed to the appropriate valve during a particular sequence. Table 3.1 shows which valves are in operation during each switch period and Fig. 3.8 shows the pneumatic control network.

FIG. 3.7 PHOTOGRAPH OF A VALVE

A = ASSEMBLED VALVE

B = UPPER DIAPHRAGM CHAMBER

C = LOWER DIAPHRAGM CHAMBER

D = NEOPRENE RUBBER DIAPHRAGM

E = INLET CHAMBER

F = OUTLET CHAMBER

G = POPPET AND STEM

H = VITON RUBBER GASKET

I = VALVE BASE

J = DIAPHRAGM BACKING PLATES

K = DIAPHRAGM ASSEMBLY COMPONENTS

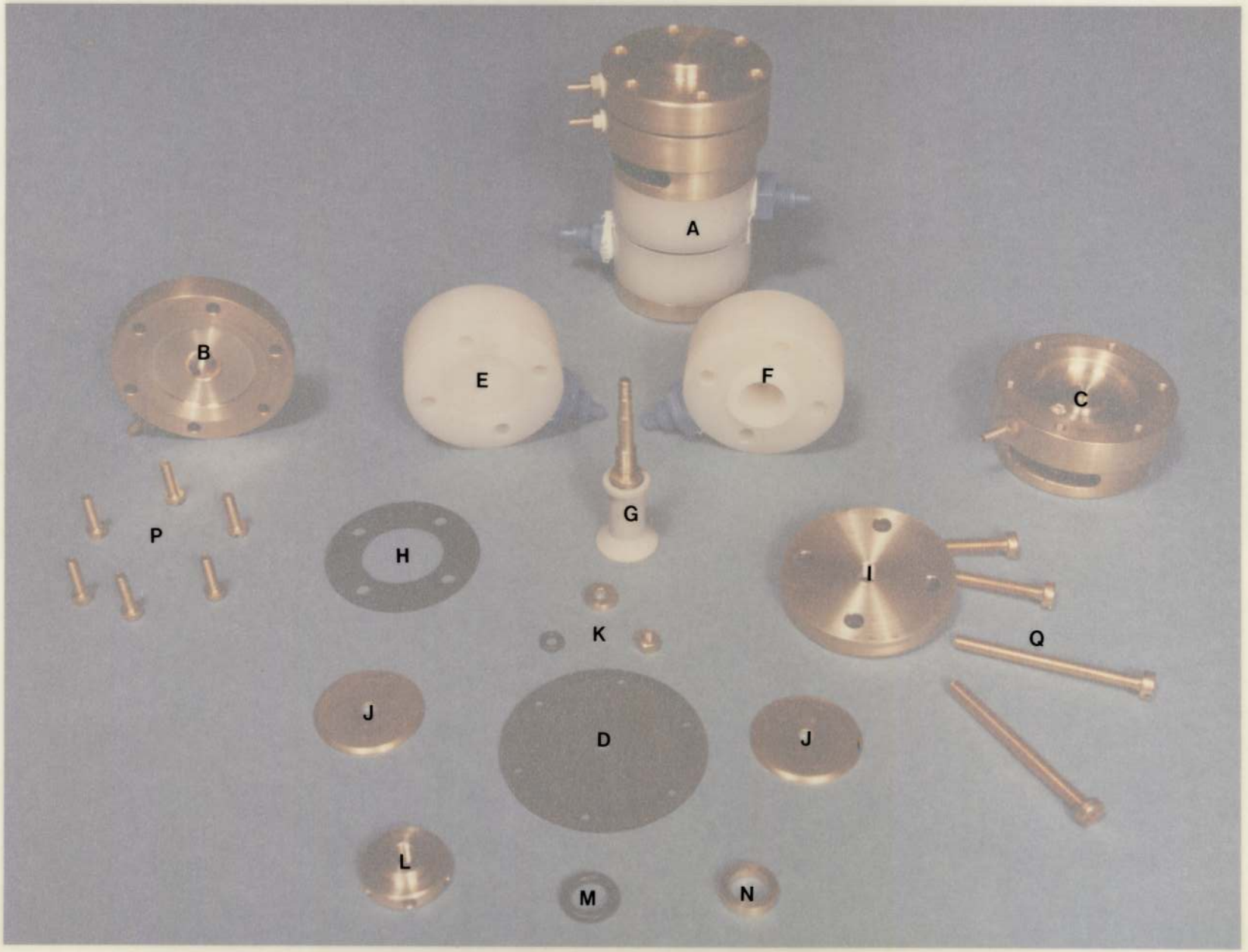
L = ADJUSTMENT NUT

M = VITON 'O' RING

N = THRUST WASHER

P = 6 x 4BA CAP SCREWS

Q = 4 x 2BA CHEESE HEAD BODY SCREWS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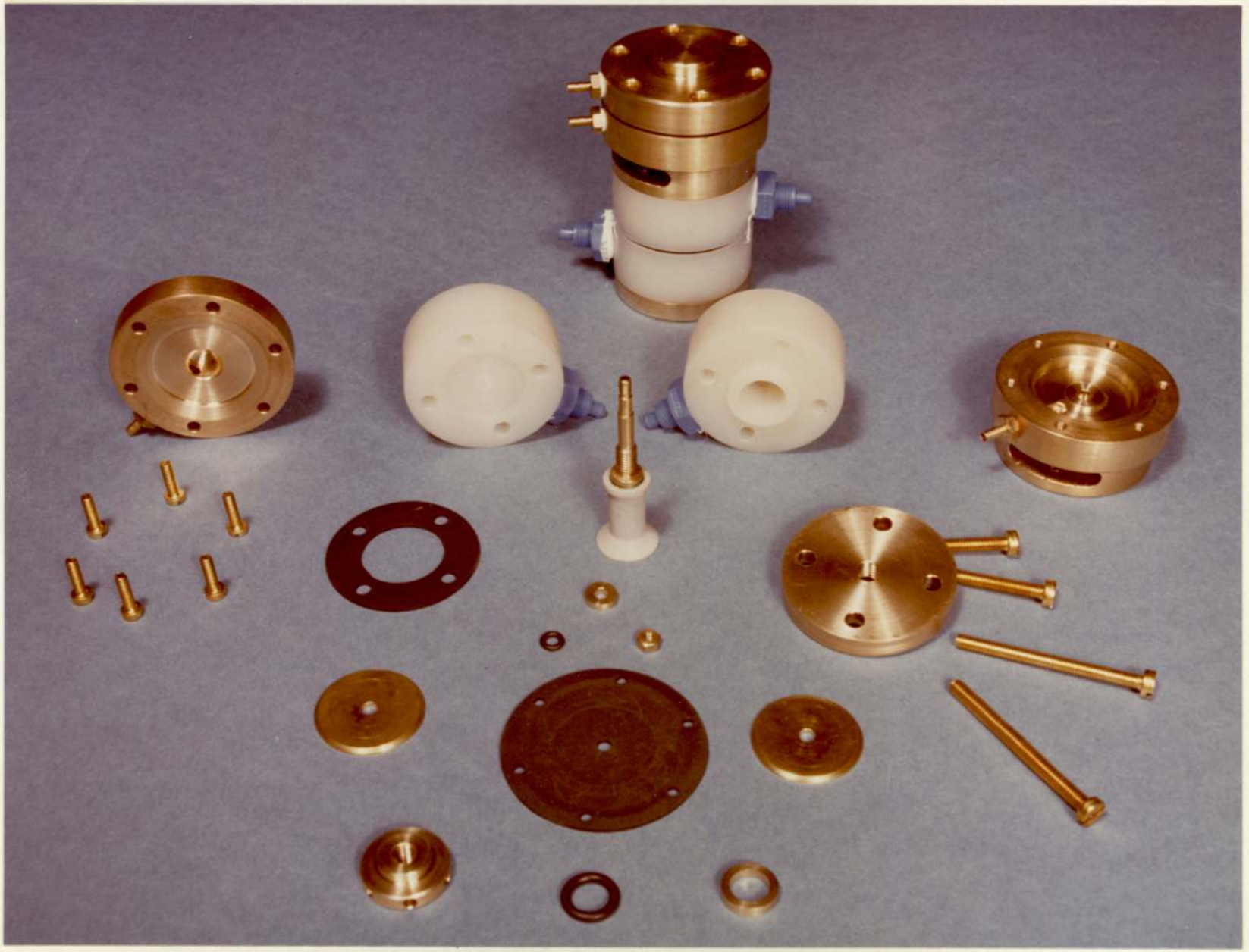


FIG. 3.8 PNEUMATIC NETWORK

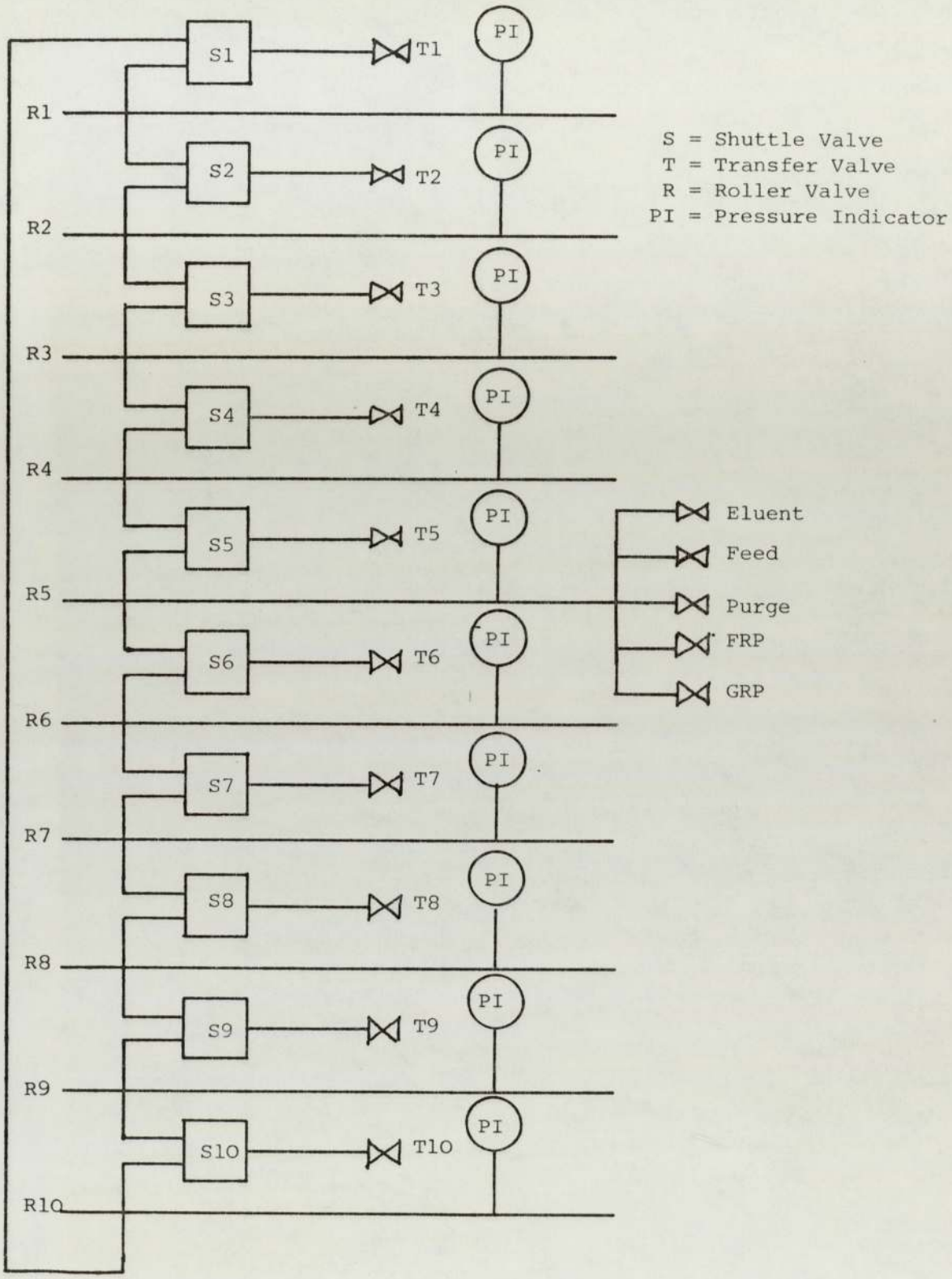


Table 3.1

SWITCH NUMBER	Valves Activated						
	To Open					To Close	
	Eluent	Feed	Purge	FRP	GRP	Transfer	Transfer
1	1	5	10	10	9	10	1
2	2	6	1	1	10	1	2
3	3	7	2	2	1	2	3
4	4	8	3	3	2	3	4
5	5	9	4	4	3	4	5
6	6	10	5	5	4	5	6
7	7	1	6	6	3	6	7
8	8	2	7	7	6	7	8
9	9	3	8	8	7	8	9
10	10	4	9	9	8	9	10

The control unit itself comprises an electric motor which drives a shaft with ten cams attached to it. These are set at 36^{o} and are out of phase with each other. Each single revolution of the shaft causes each of the ten cams in turn to operate on ten on-off roller valves. When a particular cam opens its roller valve the actuating air is allowed to flow through the valve and is then split into four streams; one stream enters a closed ring main which operates the five functional valves to open namely: eluent entry, feed entry, purge water entry, fructose product exit and glucose product exit. The second of the four streams is attached to a shuttle valve and closes one of the isolating transfer valves of the purged column, the third stream also flows to a shuttle valve and closes the other purge column isolating valve. The fourth stream flows to an indicator which shows visually which of the ten columns the eluent is entering, see Fig. 3.8. A further branch allows air to flow to a pneumatic/electric switch which stops the voltage to the motor. The switch also zeros and restarts a manually adjustable digital timer which permits the various sequencing intervals.

At the end of the sequence the timer activates the electropneumatic switch which rotates the camshaft another 36^{o} and the sequence repeats itself.

3.5 Ancilliary equipment

3.5.1 Pumps

Accurate and reliable pumps were necessary for successful operation of the refiner. Separate pumps were used for the eluent supply, feed supply and purge water supply. Over the ranges required, experiments have shown that the eluent and feed pumps were accurate to ±1% and the purge water pump to ±5%.

An L-series pump supplied by Metering Pumps Ltd. was selected for the purge water supply. This positive displacement pump operated at 150 strokes per minute and the flow range was 0-6.5 ℓmin^{-1}. This was guaranteed up to an inlet pressure of 1160 kNm^{-2}. The pump-head was constructed from polypropylene.

For the eluent supply a K-twin pump was used, again supplied by M.P.L. Its stroking speed was 96 strokes per minutes. Its flow range was 0-360 $cm^3 min^{-1}$ and operated up to 1360 kNm^{-2}.

The feed was delivered by a Hughes Micropump. Two No. 4 heads were attached to a common drive which operated 180° out of phase with each other to smooth the flow. The flow range was 0-100 $cm^3 min^{-1}$ up to an inlet pressure of 1700 kNm^{-2}.

Pressure gauges supplied by Bailey Mackay Limited and pre-set pressure relief valves were fitted to each process inlet line. Pulsation dampeners supplied by Fawcett Engineering Ltd. were also fitted to the eluent and purge supply lines to help smooth the liquid flow.

All process lines were made of nylon and piped so as to form a ring main for each function. The tubing for the eluent feed and glucose product lines was 0.6 cm internal diameter and the tubing used for the purge supply and the fructose product outlet was 0.8 cm internal diameter to carry the extra volume. All connectors, T-pieces and other fittings were manufactured from either nylon or polypropylene. All lines were colour coded.

3.5.2 Water, feed storage and product collection

All deionised water used in the experimental programme was produced from an Elgastat B224 deioniser supplied by Elga Products

Ltd. Town's water was supplied to the deioniser at approximately 340 KNm^{-2} and the conductivity of the deionised water produced remained below 50 μS. When the deionising resin in the cartridge became exhausted and this conductivity value was exceeded, a solenoid valve in the control head on the cylinder was automatically activated to prevent further flow of water. The water produced was stored in two large stainless steel tanks with a combined capacity of 750 ℓ. A feedback level controller was arranged to prevent the tanks from overflowing. These storage tanks were elevated and hence gave each pump a net positive suction head. The supplies to each pump were fed by a 2.54 cm diameter pipeline in a ring main from the reservoir.

The feed solution after being prepared in a 150 ℓ polythene tank was pumped as required to a 20 ℓ glass aspirator situated above the feed pump. This ensured a net positive suction head.

Both the feed and eluent flows entering the refiner could be diverted via a 3-way valve into calibrated flow measuring devices. The purge flow rate was calculated by weighing the product. Collection of both products was into plastic containers. All inlet pressures generated could be monitored on the relevant gauges.

CHAPTER 4

COMMISSIONING THE SCCR

4.1 Treatment of the packing

Zerolit 225 SRC 14 was selected as the ion exchange resin to effect the separation of the sugars to be used. It had been used successfully by Ching (6) in his work using a 2.54 cm diameter SCCR. 65 kg of the resin was purchased from Diamond Shamrock Polymers Ltd. of Middlesex. The particle size range was quoted by the manufacturer to be 52-100 mesh (150-300 μm). For a full sieve analysis see Table 4.1.

The resin is a sulphonated polystyrene with 8% crosslinking by divinyl benzene. This gives the resin mechanical strength to withstand pressures greater than 2000 kNm^{-2}. It may also be used over a temperature range of $-10^{o}C$ to $140^{o}C$ and is unaffected by pH variations in the range 1-14.

To allow the resin to retain between 5%-30% of free hydrogen ions on the resin, the conditioning process to the calcium form followed that outlined in Boehringer's patent (1). For the conversion from the sodium form in which the resin was supplied to the hydrogen form, a 10% w/v hydrochloric acid solution was used. Considering the quantity of resin and the dangers of handling the acid, two 30 ℓ batches of resin were treated in turn.

The calculation of the volume of reagent necessary to effect the conversion was based on the wet exchange capacity of the resin. This value was quoted by the manufacturer as 2 milliequivalents per cm^3 ($meq\,cm^{-3}$). Literature (86) suggests that the ratio of:

$$\frac{\text{meq regenerent}}{\text{meq resin}} = 4:1$$

B.S. Sieves meshes per inch	Size range (μm)	Mass Fraction
+ 52 mesh	> 295	0.03
- 52 + 60	251 - 295	0.48
- 60 + 72	210 - 251	0.21
- 72 + 85	178 - 210	0.14
- 85 + 100	152 - 178	0.11
- 100 mesh	< 152	0.03

TABLE 4.1 - Sieve Analysis for Ion Exchange Resin

The volume of 10% hydrochloric acid required for regeneration was calculated thus:

Wet exchange capacity of resin	=	2 meqcm^{-3}
Total exchange capacity of 30 ℓ resin	=	60,000 meq
Ratio of regenerant to resin	=	4:1
Required exchange capacity of acid	=	4 x 60000
	=	240,000 meq

As the regenerant is 10% w/v,

$$\text{Volume of acid necessary} = \frac{240000 \times 36.5 \times 100}{1000 \times 10}$$

$$= 87.6 \text{ ℓ}$$

The method recommended by the patent (1) for the conversion suggests a continuous flow of regenerant through the resin bed at a flow rate of 0.1 bed volumes per minute. This continuous technique was impractical due to lack of equipment for the volumes concerned. Instead the resin was placed in a 60 ℓ polythene tank and the acid added to it in 10 ℓ batches. After the addition the resin was agitated for an equivalent residence time which the continuous system would have provided. The liquor was then syphoned off. This process continued until all the acid had been used up.

The resin was then rinsed with deionised water until the pH of the effluent liquor was neutral indicating removal of all excess acid.

The conversion of the resin to the calcium form was achieved using 10% w/v calcium chloride solution. To allow the 5%-30% free hydrogen ions on the resin, the pattern suggested in the Boehringer patent (1) was followed. This states that all the hydrogen ions are only replaced with calcium ions if the regeneration is

carried out using a calcium chloride solution with a pH greater than 8.0 and at a temperature greater than 60^{o}C. Thus the conversion was carried out at ambient temperature and with a 10% solution which had a pH of 5.7.

The volume of regenerant was calculated as previously and added in the same manner. The resin was again rinsed with deionised water until a neutral pH of the effluent was obtained.

4.2 Column packing technique

As the packing had to remain wet, a slurry packing technique was necessary. Three methods of slurry packing were investigated. These were: random pouring, pouring with external vibration and pouring with an applied vacuum. To assess their efficiency and repeatability glucose chromatograms were performed and the height equivalent to a theoretical plate (HETP) measured. The results are shown in Table 4.2. From this it was concluded that pouring with applied vacuum gave the best results and this technique was adopted for all the columns. The resin to water ratio was approximately 1:1.

After packing the piston was positioned and secured and each packed column weighed to enable the weight of resin in each column to be calculated. The results are shown in Table 4.3. The columns were then fitted with a retaining bracket and mounted on the pre-constructed mobile framework. The hydraulic ring main was connected and attached to a small hydraulic hand pump. This was operated until a positive pressure of 200 kNm^{-2} was acting on the packed beds. The hydraulic system was then sealed.

H.E.T.P. For Glucose (cm)			
	Trial 1	Trial 2	Trial 3
Pouring only	3.10	2.96	3.04
Pouring with vibration	2.72	2.50	2.42
Pouring with vacuum applied	2.32	2.42	2.03

Table 4.2 - Comparison of Heights Equivalent to a Theoretical Plate (HETP) for Various Slurry Packing Techniques

4.3 The valves and pipework

To prevent any air entering the packed beds while the piping up was taking place a continuous stream of deionised water was allowed to flow through the bed. In the process line networks, the piping between a column and its transfer valve was kept as short as possible so as to minimise the dead volume and keep any back-mixing and diffusion effects as small as possible.

All the colour coded process ring mains were connected to their respective pumps. The airline networks to the inlet and outlet valves also formed ring mains and were connected to their respective control valves. Deionised water was then pumped around the SCCR while the control box sequenced the valves every 30 minutes. This allowed a check for correct valve sequencing, leaks, correct flow directions of the various streams and satisfactory control box operation. Mass balances were performed to check the calibration of the flow measuring devices. After minor adjustments the rig was considered ready for use.

4.4 Characterisation of the columns

4.4.1 A theoretical basis for comparison

Packed density is often used to characterise a packed bed. Generally the greater the packed density the more efficient the packing. The packed density can be calculated if the voidage can be measured. This may be done by running a chromatogram through the bed using a totally excluded species. This means that the solute only travels in the interstitial or void volume and does not enter the intra-particle volume. The elution volume necessary for this solute to be removed from the column is the void volume, V_o, of that column.

In the measurement of this value, a polymer, dextran with a molecular weight greater than 5000, was selected as the totally excluded species.

The criterion for comparing the performance of a chromatographic column is generally the height equivalent to a theoretical plate HETP. This is calculated by dividing the bed height L by the number of theoretical plates N in a column. The number of plates in a column is calculated directly from a chromatogram using Glenckauf's equation (12) as demonstrated in section 2.1.4. Consequently to assess this parameter for both fructose and glucose, chromatograms were developed for each sugar.

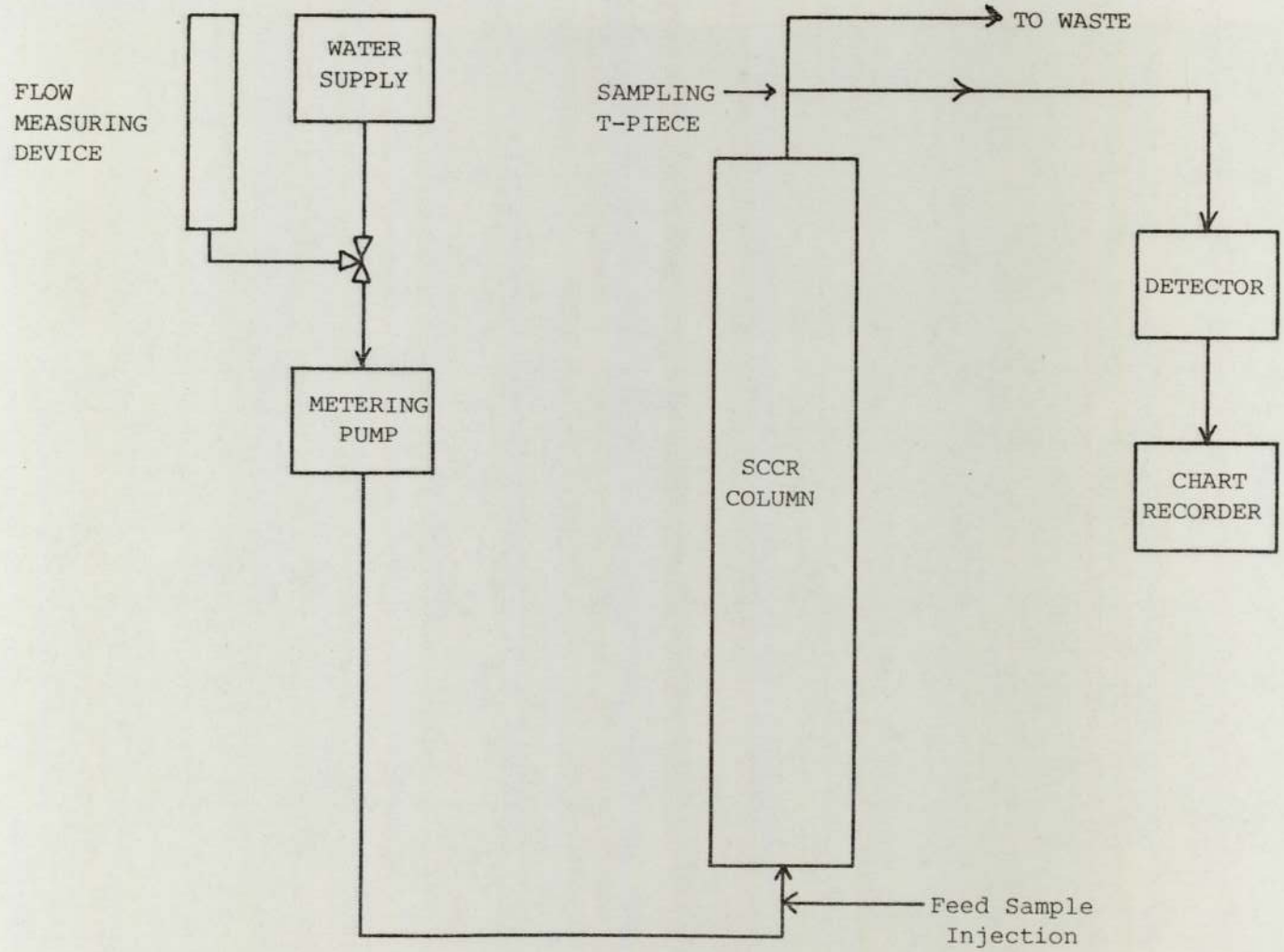
4.4.2 Experimental techniques

Three chromatograms per column were run: dextran, fructose and glucose. These were performed in the batch mode.

T-pieces with injection and sampling points were fitted as close as possible to the inlet and outlet of the column under test. This minimised any extra-column dispersion. A sampling needle was inserted into the rubber septum at the outlet which allowed a stream of the outlet liquid to pass along very narrow bore tubing to a refractive index detector. This was linked to a chart recorder, see Fig. 4.1.

The eluent pump was set at 105 $cm^3 min^{-1}$ and the flow started. This flow rate was used for all chromatograms as it was in the predicted operating range of future experimental work with the refiner. When a steady base line had been established on the chart recorder, a 5 cm^3 slug of the feed at a concentration of 10% w/v solids was injected into the inlet line through the rubber septum.

FIG. 4.1 FLOW DIAGRAM FOR INDIVIDUAL COLUMN COMPARISON



The peak was shown on the chart recorder as the sugar was eluted. All the chromatograms produced were analysed and the information determined shown in Tables 4.3 and 4.4.

4.4.3 Discussion of results

Calculation of the retention volume of dextran V_d gives the interstitial or void volume V_o. This in turn allows the packed density to be found knowing the total column volume and the weight of resin therein. The wide range of the void volumes for the ten columns was thought to be the result of column packing. As has been discussed previously (section 2.3.1.1) repeatable packing becomes more difficult with increase in column diameter.

If the voidages, ε, the ratio of the void volume V_o to the total bed volume are studied it may be seen that they are considerably lower than those quoted in the literature (87) for spherical particles. This is thought to be so for two reasons: (i), the particle size range, (ii), the compression by the piston. The size range of the particles is 150-300 μm. Thus the smaller ones may well be dispersed to fill the volume between the larger ones more efficiently.

The retention volume for glucose V_g varied in the columns. This is believed to reflect the differences in packing efficiency. The glucose molecules are small enough to diffuse in and out of the intra-particle volume. This is why the V_g values are greater than the V_o values. The non-uniformities in the packing give rise to variations in the flow velocities across the bed. This in turn gives rise to band spreading of the solute as it travels through the column.

Col. No.	wt. of wet resin g	Bed Volume cm^3	Bed height cm	Average Packed Density $g\ cm^{-3}$	Void Volume (V_o) cm^3	Voidage
1	7604	6030	65.9	1.26	1866	0.31
2	7832	6200	67.7	1.26	1667	0.27
3	7832	6210	67.9	1.26	1743	0.28
4	7720	6240	68.2	1.24	1963	0.31
5	7832	6210	67.9	1.26	1564	0.25
6	7832	6150	67.2	1.27	1932	0.31
7	7491	6120	66.9	1.23	1840	0.30
8	7491	6060	66.2	1.24	2070	0.34
9	7720	6210	67.9	1.24	1840	0.29
10	7720	6100	66.6	1.27	2036	0.33

Table 4.3 Individual Column Properties

Table 4.4 Individual Column Properties (Continued)

Col. No.	Elution Volumes		H.E.T.P.'s		Dist. Coeffs	
	Fructose (V_F) cm^3	Glucose (V_G) cm^3	Fructose cm	Glucose cm	K_{D_f}	K_{D_g}
1	3976	2679	3.66	2.64	0.51	0.20
2	3335	2415	2.94	2.33	0.41	0.17
3	4011	2968	3.39	2.19	0.51	0.15
4	3874	2646	3.78	2.13	0.45	0.16
5	3492	2421	3.77	2.226	0.42	0.16
6	3769	2824	3.53	1.92	0.44	0.21
7	3528	2763	4.18	3.34	0.40	0.20
8	3910	2760	2.87	1.29	0.46	0.17
9	4000	3105	2.82	2.43	0.49	0.21
10	4140	2875	2.77	1.62	0.52	0.21

Giddings (15) has cited five types of velocity inequality which may arise:

1. Transchannel - in each interstitial channel a higher velocity exists in the centre of the flow pattern than at the walls of the channel.
2. Transparticle - the mobile phase within the particle is essentially stagnant but is surrounded by moving mobile phase thus creating a velocity difference.
3. Short range interchannel - small tightly packed regions of the bed joined together by rather loosely filled space composed of large channels in which lower velocities occur.
4. Long range interchannel - the above pattern in (3) of large channels alternating with small ones gives rise to velocity variations with each indulation leading to a long range interchannel effect.
5. Trans column effect - velocity differences between the outer regions i.e. near the wall and the centre of the packed bed.

The velocity inequalities are dependant on the packing of each individual column and it is thought that these differences are reflected in the variation of V_g.

The retention volumes for fructose V_f vary both for the same reasons as for glucose but the band is broadened even more because of the chemical complexing which occurs with the calcium ions. The variation may also be a result of different amounts of calcium charge in each column.

The HETP values vary in some cases. The method used to calculate their value basically compares the elution time in relation to the bandwidth. As the elution times vary somewhat, as shown in the elution volume measurements, some variation will automatically be incurred. However the width of the solute band and its variation from column to column is a function of the velocity inequalities, the distribution within the column and possibly kinetic effects which occur as the band travels down the column.

4.4.4 Determination of distribution coefficients

The fundamental retention equation for a solution in chromatography is defined (9) as:

$$V_R = V_M + K_d V_S \qquad 4.1$$

Where V_R = retention volume of a component

V_M = volume of mobile phase in column

V_S = volume of stationary phase in volumn

K_d = distribution coefficient

K_d is defined as

$$K_d = \frac{\text{concentration of solute in the stationary phase}}{\text{concentration of solute in the mobile phase}}$$

From equation 4.1 the K_d can be defined as a function of the retention volume of a solute

$$K_d = \frac{V_R - V_M}{V_S} \qquad 4.2$$

Consequently the distribution coefficient of glucose may be defined as:

$$K_{dg} = \frac{V_g - V_d}{V_T - V_d} \qquad 4.3$$

Where V_T = Total column volume

V_d = Mobile phase volume or void volume V_o

and for fructose

$$K_{df} = \frac{V_f - V_d}{V_T - V_d}$$

The values for K_{dg} and K_{df} can be seen in Table 4.4.

From the elution volumes measured it may be seen that if a three component feed mixture of dextran, glucose and fructose were eluted through a column, dextran would emerge first followed by glucose followed by fructose. This is because the dextran molecules are too large to enter the stationary phase: thus there is no distribution and it elutes in the mobile phase volume of the column.

Glucose is retained in the stationary phase because the molecules are physically small enough to enter the intra-particle volume of the stationary phase and a gel permeation effect results giving a distribution of glucose between the two phases.

The fructose molecules are retained for longer than the glucose because not only are they small enough to enter the stationary phase and be retained by the gel permeation effect but also they form the complex with the calcium ions and are consequently retained to a

greater extent. As migration of solute molecules only occurs when they are in the mobile phase they take longer to travel through the column. Thus fructose has a greater K_d value than glucose.

CHAPTER 5

OPERATION OF THE SCCR

5 Introduction

Prior to any experimental run using the rig, theoretical calculations were performed to fix various parameters - particularly flow rates.

The main objectives of the research programme were as follows:

1. To assess the rigs' capability to separate or refine fructose-glucose and fructose-glucose-dextran feed mixtures of varying concentrations under various operating conditions.
2. To obtain information about how product qualities, concentrations and throughputs were affected by changing various parameters.
3. To study the scale-up effects in comparison with similar data obtained from an SCCR with 2.54 cm diameter columns.
4. To simulate the performance using a mathematical model which has been developed and adapt it to the larger system.

An item of importance in the first run was to establish how many cycles were necessary for the SCCR to reach a state of pseudo equilibrium.

5.1 Theoretical operating conditions

In the chromatographic unit which has been constructed glucose travels preferentially with the mobile phase and fructose with the stationary phase.

These two phases move counter-current to each other in a semi-continuous fashion while a feed stream, an eluent stream and a purge stream enter the column and two product streams leave it continuously. This is shown in Fig. 5.1.

A model can be constructed relating flow rates of the mobile phase and stationary phase and component separation. A material balance on glucose about the feed point gives:

$$F_g = L_e y_g + P x_g \qquad 5.1$$

Where

F_g = mass rate of glucose input at feed point (gm s^{-1})

L_e = eluent phase flow rate ($cm^3 s^{-1}$)

P = stationary phase flow rate

y_g = concentration glucose in eluent phase (gm cm^{-3})

x_g = concentration glucose in stationary phase (gm cm^{-3})

For a glucose molecule to move preferentially with the eluent phase then

$$L_e y_g > P x_g \qquad 5.2$$

Rearranging

$$\frac{L_e}{P} > \frac{x_g}{y_g} \qquad 5.3$$

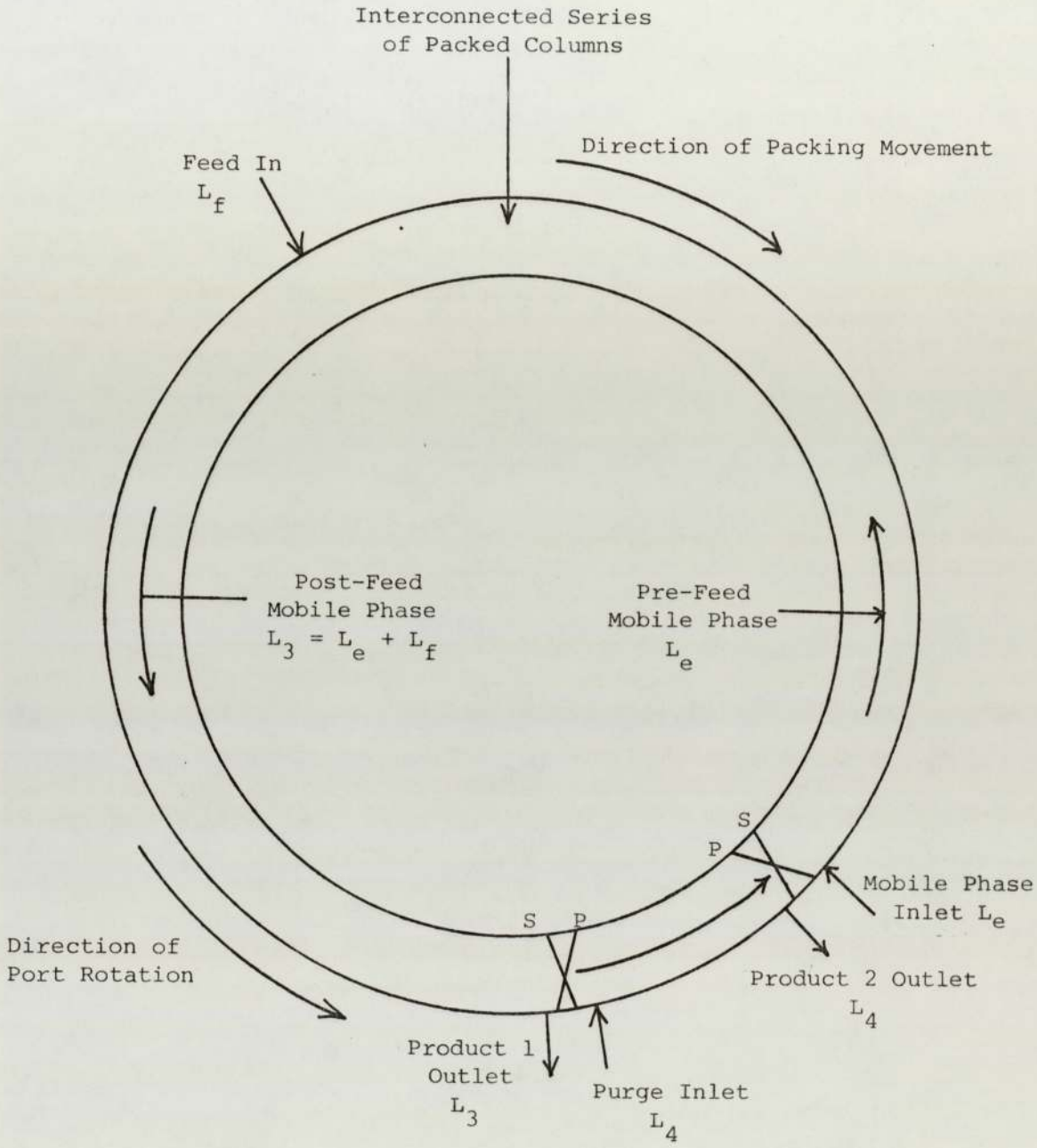
and since

$$\frac{x_g}{y_g} = K_{dg} \qquad 5.4$$

then

$$\frac{L_e}{P} > K_{dg} \qquad 5.5$$

FIG. 5.1 SCHEMATIC DIAGRAM OF SCCR UNIT



S-S Separation Section

P-P Purge Section

Product 1 - Glucose-Rich

Product 2 - Fructose-Rich

Similarly for fructose to move with the stationary phase

$$\frac{L_e}{P} < K_{df} \qquad 5.6$$

Combing equations 5.5 and 5.6 we obtain the theoretical limits of eluent and stationary phase flow rates to achieve separation of the fructose and glucose:

$$K_{dg} < \frac{L_e}{P} < K_{df} \qquad 5.7$$

In the refiner the stationary phase flow is achieved by the sequencing action at the end of every switch period. As each column contains eluent phase in the void volume the actual eluent phase flow rate is effectively reduced so:

$$L_e = L_i - \frac{V_o}{S} \qquad 5.8$$

Where L_i = measured eluent phase flow rate $cm^3 s^{-1}$

V_o = void volume (cm^3)

S = switch period (S)

The equilibrium distribution coefficients K_d's were measured as described in section 4.4.2 under dilute conditions.

To completely purge all the fructose from the isolated column it is therefore necessary for

$$\frac{L_4}{P} > K_{df} \qquad 5.9$$

where L_4 = Purge stream flow rate

5.2 The practical case

The above model is very simplistic and assumes ideal conditions - a small feed flow rate, low feed concentrations a truly continuous mode of operation. In practice these factors are significant and can cause a departure from ideality. They are discussed in detail below.

5.2.1 Band broadening

This is the effect of velocity inequalities caused by non-uniform packing, kinetic effects, and adsorption-desorption in the stationary phase. The extent of broadening is a function of the rate of flow relative to the rate of mass transfer. As a result of this broadening the component is eluted with K_d values varying between $K_d-\delta$ to $K_d+\delta$ where δ is related to the peak width on the chromatograph. When the continuous system is considered this effect would increase the length of column necessary to attain the resolution necessary. Thus to achieve separation the L_e/P ratio must be narrowed to within these imposed limits:

$$(K_{dg} + \delta_1) < L_e/P < (K_{df} - \delta_2) \qquad 5.10$$

5.2.2 Finite feed flow rate

The feed enters the refiner continuously at about the midpoint of the separating section. This gives rise to an increase in mobile phase flow rate at and beyond the feed point. The separating length may therefore be considered as two sections, the prefeed section and the post feed section. Two sets of operating conditions will consequently exist. The difference between the two

is dependent on the magnitude of the feed flow rate. Thus equation 5.10 becomes:

$$(K_{dg} + \delta_1) < \frac{L_e}{P} < \frac{L_e'}{P} < (K_{df} - \delta_2) \qquad 5.11$$

Where $L_e' = L_e + L_f$

and L_f = feed flow rate ($cm^3 s^{-1}$)

To minimise the effect of flow discontinuity $L_e >> L_f$. However it is desirable to maintain a high L_f to increase the mass throughput. A compromise is therefore necessary.

5.2.3 Effect of a finite feed concentration

Equation 5.11 uses a distribution coefficient which has been measured as described previously under dilute conditions for a single slug of feed. Changes then are expected to occur in the distribution and as shown in section 2.3.1.2 the shape of the distribution isotherm will change. The exact nature of this concentration effect on the K_d values is uncertain as yet for no experimental data are available. When these do become available, modifications will be necessary in equation 5.11 to account for this effect and its limitations on the operating conditions of the refiner.

It is apparent that if both distribution coefficients increase in value, i.e. both components become more attracted to the stationary phase, a wider range of operational flow rates will be allowable. If they both decrease, then a narrower range will be available. The most significant effect would be caused if the results

showed a decrease in the glucose distributions coefficient and an increase in the fructose distribution coefficient. This would allow a very wide range of flow rates to be used. A term δ', must therefore be included in the equation:

$$(K_{dg} + \delta_1 + \delta_1') < \frac{L_e}{P} < \frac{L_e'}{P} < (K_{df} - \delta_2 - \delta_2') \qquad 5.12$$

5.2.4 Effect of the semi-continuous mode of operation

During a switch period the unit is operating as an elution batch chromatograph. The counter-current movement of the stationary phase is imposed by the discontinuous 'stepping' of the inlet and outlet valves around the ten linked columns. This fact is of particular importance when selecting the purge stream flow rate. Within the switch period all fructose must be removed from the isolated column to avoid contamination of the glucose product in the next switch period. To ensure this, L_4, the purge flow rate was set higher than the theoretical limit as calculated from equation 5.9 thus the concentration will decrease.

$$\text{Thus } \frac{L_4}{P} >> (K_{df} - \delta_2' - \delta_2'') \qquad 5.13$$

The isolated column in one switch becomes the glucose product outlet column in the next. As explained above this column contains no sugar but the void volume is full of water. This has to be passed out in this product stream before any emergence of the glucose is detected. Thus this product stream is much diluted.

5.2.5 General comments on operating conditions

Equations 5.7 and 5.9 give the theoretical values for

$\frac{L_e}{P}$ and $\frac{L_4}{P}$. These are revised to equations 5.12 and 5.13.

To evaluate the 'δ' terms and how these practical limitations affect the performance requires considerable experimental work and would involve batch and continuous chromatography. Applying them to a semi-continuous system would contain inherent error. In selecting actual operating conditions for initial experiments these equations were used as a guide. Subsequent experience in operation of the rig and analysis of the results obtained often influenced subsequent operating conditions.

5.3 Experimental procedure

5.3.1 Feed preparation

The average volume of feed necessary for each run was 120 ℓ. Consequently the fructose and glucose were purchased in 25 kg bags from wholesalers. They were both of food grade purity. The glucose was only available as the monohydrate.

The feed solution of the required concentration was prepared by dissolving the calculated weight of the sugars in deionised water. This water was preheated if the required concentration was greater than 40% w/v. After dissolution a sample was analysed to determine the exact concentration of the sugars.

In order to prevent any biological growth in the pipework or on the columns an inhibitor, sodium azide, NaN_3, was used. The dose recommended by Fisons Ltd. was 0.02% w/v. This was selected because sodium ions did not displace calcium ions on the resin.

5.3.2 Preliminary checks

The considerable quantities of deionised water necessary during operation (40 ℓhr^{-1}) were produced by an Elgastat B224 deioniser. This with its level controlling device in the holding tanks was switched on. The conductivity of the produced water was checked to ensure it was below the level of 50 μS. The portable air compressor was activated and the bias and activating pressures checked.

5.3.3 Start up procedure

1. The eluent, feed and purge pump were set to the required flow rates.
2. The digital timer for controlling the switch period was set to the required interval.
3. All valves were opened and collectors positioned.
4. Pumps were turned on and the timer started.

5.3.4 Procedures during operation

5.3.4.1 Liquid flow rates

Immediately after start up the eluent inlet flow rate and feed inlet flow rate were checked using the calibrated measuring devices and adjusted if necessary. The purge column flow rate was measured by monitoring the outlet stream. These flow measurements were generally made once during each switch period.

At the end of a cycle, the product collecting vessels were weighed and a mass balance carried out.

5.3.4.2 Pressure drop data

The pressure drops of the eluent, feed and purge streams through the rig were recorded every switch. It was observed that as the switch progressed pressure drops in the eluent and feed lines increased as on column concentrations changed. It was also observed that the feed pressure was just over half the value of the eluent pressure. This was because the feed entered the rig $\frac{4}{9}$ of the way along the separating column. The purge pressure drop was high at the beginning of a switch but soon decreased as the fructose was removed.

5.3.4.3 Sample collection

At the end of each cycle a sample was taken of the bulked switch period products of that cycle. These were analysed for purity and concentration as the run progressed. When designing the columns, sample points were included along the length of each column. This allowed "on column" samples to be withdrawn by inserting a hypodermic needle through the silicone rubber septum and filling a 2 cm^3 syringe. These samples were removed at the same time, from the same point on the same column in each switch period. These were immediately analysed and from the data obtained, an "on column" concentration profile constructed. This was possible because during the ten switch periods of a cycle, each and every column served every function, i.e. as the purge column, the eluent entry column, the feed column, the glucose exit or a part of the separating length. Hypodermic needles of various lengths were made to order by Coopers Needleworks of Birmingham. These allowed traverses of the bed to be made for radial concentration profiles to be determined.

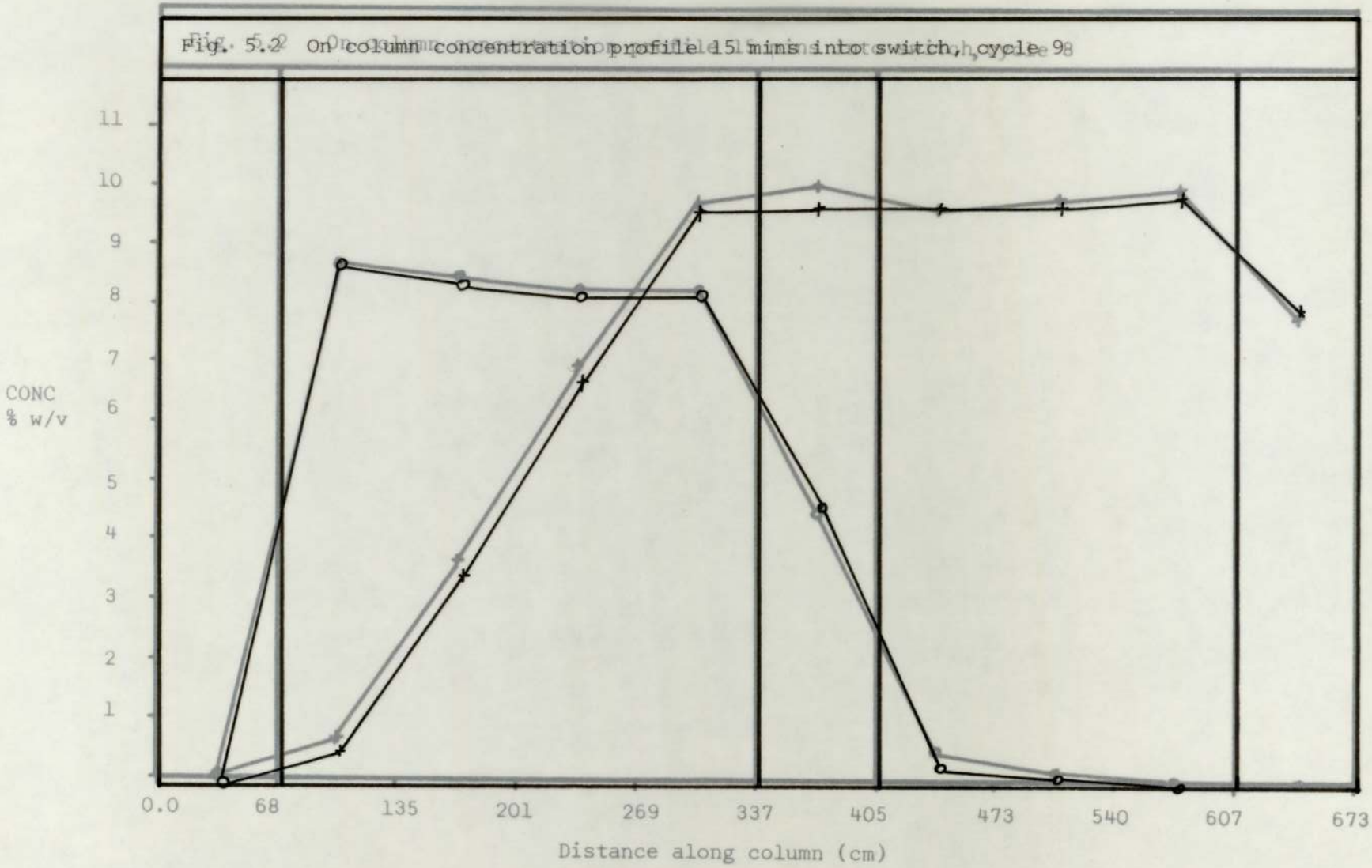
5.3.4.4 Establishing pseudo-equilibrium

In a distillation column which is a truly counter-current mass transfer process, when equilibrium is established the tray to tray compositions remain constant along the column. In a semi-continuous process the same dynamic equilibrium is not possible because of the stepwise nature of operation. During a switch period of say 30 minutes the concentration profile will be gradually changing along the column as it would in a batch chromatograph. This is because only the mobile phase is physically moving. At the end of the switch period the stepwise counter-current movement of the stationary phase takes place and the concentration profile is consequently displaced by one column length.

Therefore only a "pseudo equilibrium" can be achieved. This occurs when the displacement of the profile is constant from switch to switch i.e. the concentration profile is identical over the whole column length at the same time point in every switch.

To assess this, actual samples of 'on column' fluid were withdrawn using the sample points. This took place from the same point on the same column at the same fixed point in time into each switch while the experiment was in progress. These were then analysed on line and the results plotted graphically on tracing paper as shown in Fig. 5.2. When two profiles were the same from one cycle to the next it was assumed that the pseudo equilibrium state had been reached. To check this the product concentrations were measured and their weights recorded. This enabled a mass balance on each of the sugars to be carried out. When these were approximately 100% equilibrium was assumed.

Fig. 5.2 On column concentration profile 15 mins into switch, cycle 98



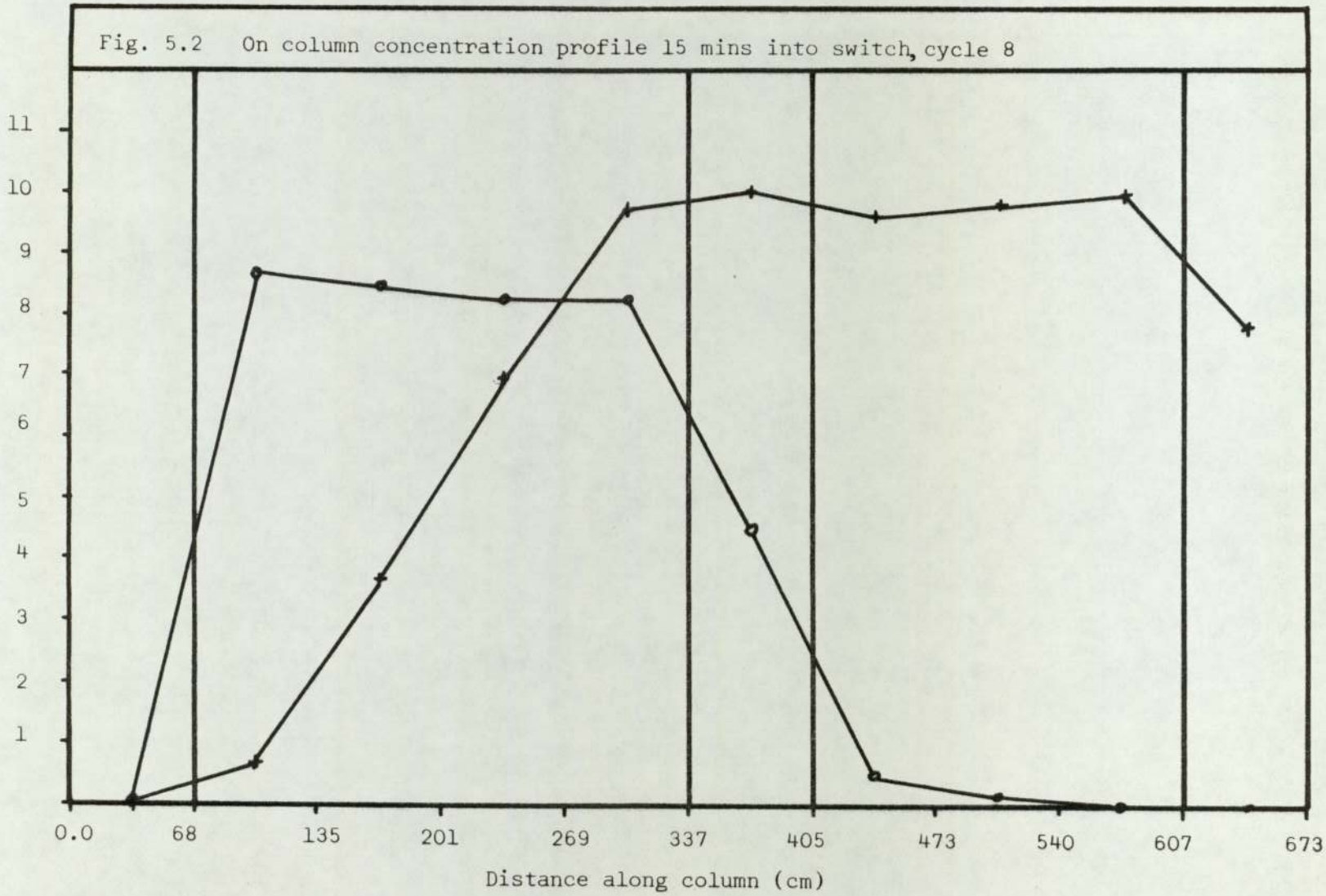
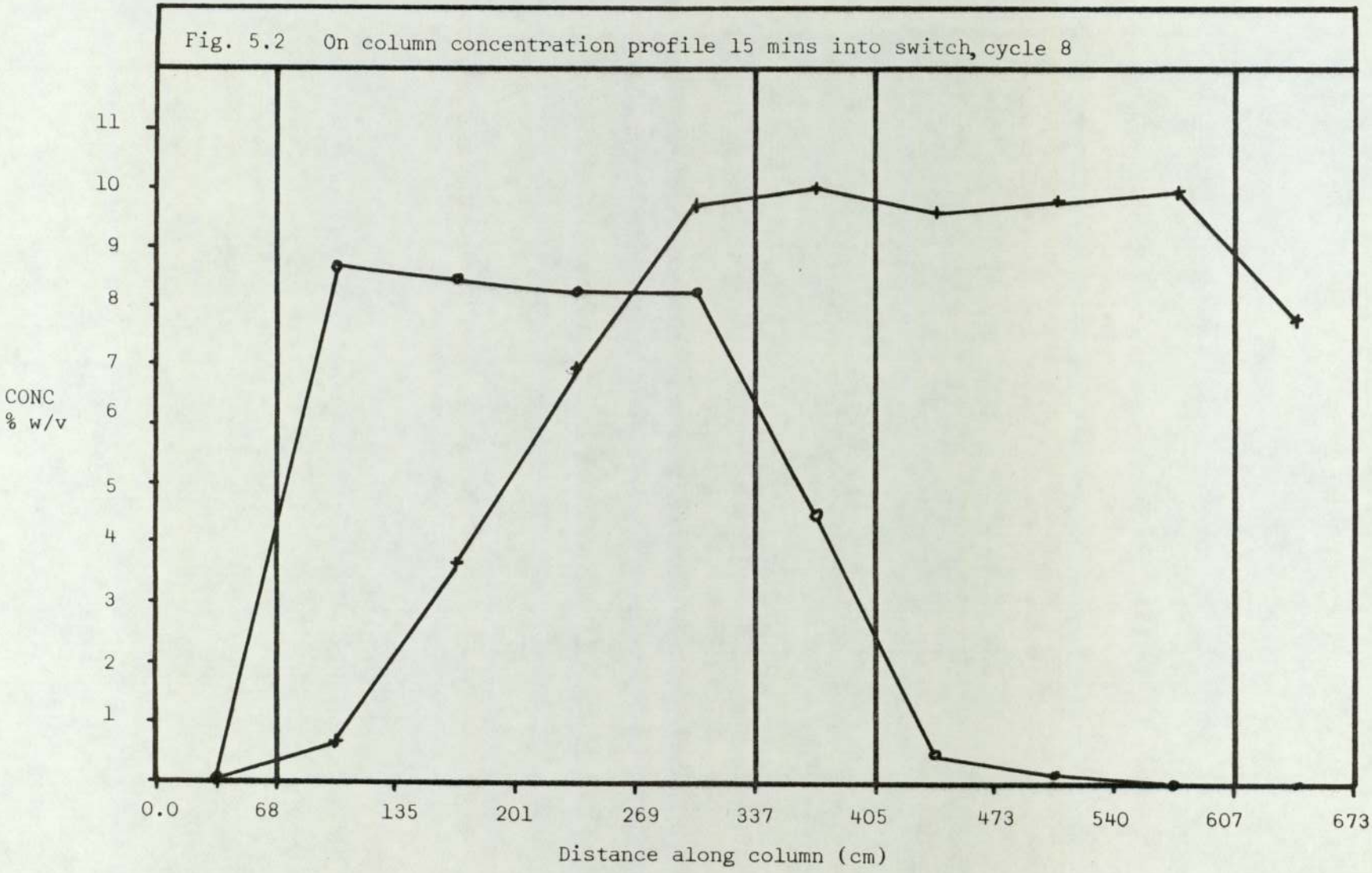


Fig. 5.2 On column concentration profile 15 mins into switch, cycle 8

Fig. 5.2 On column concentration profile 15 mins into switch, cycle 8



5.3.5 Purging the columns and shutdown procedure

As the last cycle was due to finish all pumps were turned off. The feed and eluent inlet valves were closed. The control system was switched to manual and moved one position. The purge pump was again activated and the product collected in a separate container. This was weighed and analysed. The procedure was repeated until all ten columns had been purged out and their products weighed and analysed. The analysis and weight enabled the mass of each sugar on each column to be calculated. From these data an end of run average concentration profile could be constructed.

5.3.6 The analytical system

During and after an experimental run a large number of analyses were necessary. The equipment therefore needed to be fast, accurate and reliable. A carbohydrate analysis column manufactured by Bio-Rad Ltd. was purchased. This was a high performance liquid chromatographic column (HPLC). It was 250 mm in length and 4.0 mm in diameter. It was packed with Animex HPX87 a polymer based ion exchange resin charged in the calcium form. The average particle size was 10 μm. The plate number as quoted by the manufacturer was 2700 giving a HETP of 9.25×10^{-3} mm.

The eluent used was deionised water which before entry to the column was degassed at $65^{o}C$ by placing the reservoir in a constant temperature water bath. To protect the column against any particulate matter, the eluent was filtered through a Whatman Grade 10 inline filter.

The column itself was operated at 85°C by means of a waterjacket and heater circulator unit supplied by Techam Ltd. of Cambridge. A Specac six port sample injection valve with a 25 µl external loop was used to load the samples on to the column. Samples were filtered through a 0.45 µm filter before application. Eluent was supplied to the column by a micropump supplied by Metering Pumps Ltd. of Ealing. During its use the flow rate was set at approximately 0.3 $cm^3 min^{-1}$ which gave a pressure drop through the column of approximately 7800 kNm^{-2} (1000 psi). Under these conditions glucose was eluted in approximately 190 seconds and fructose in approximately 270 seconds.

To detect the sugars as they emerged, a differential refractometer supplied by L.D.C. of Stone was used. Its mode of operation was to compare the refractive index of the outlet stream of the HPLC column (containing the separated sugars) with that of the pure eluent, deionised water. The detector was linked electrically to a Servoscribe Chart Recorder and a Hewlett Packard 337B Integrator, see Fig. 5.3. This arrangement gave a chromatogram of the separation on the chart together with a printout of the area under the curve by the integrator. The actual sample concentrations were calculated by comparing the areas obtained with those of known standard samples. The relationship was linear over the range used.

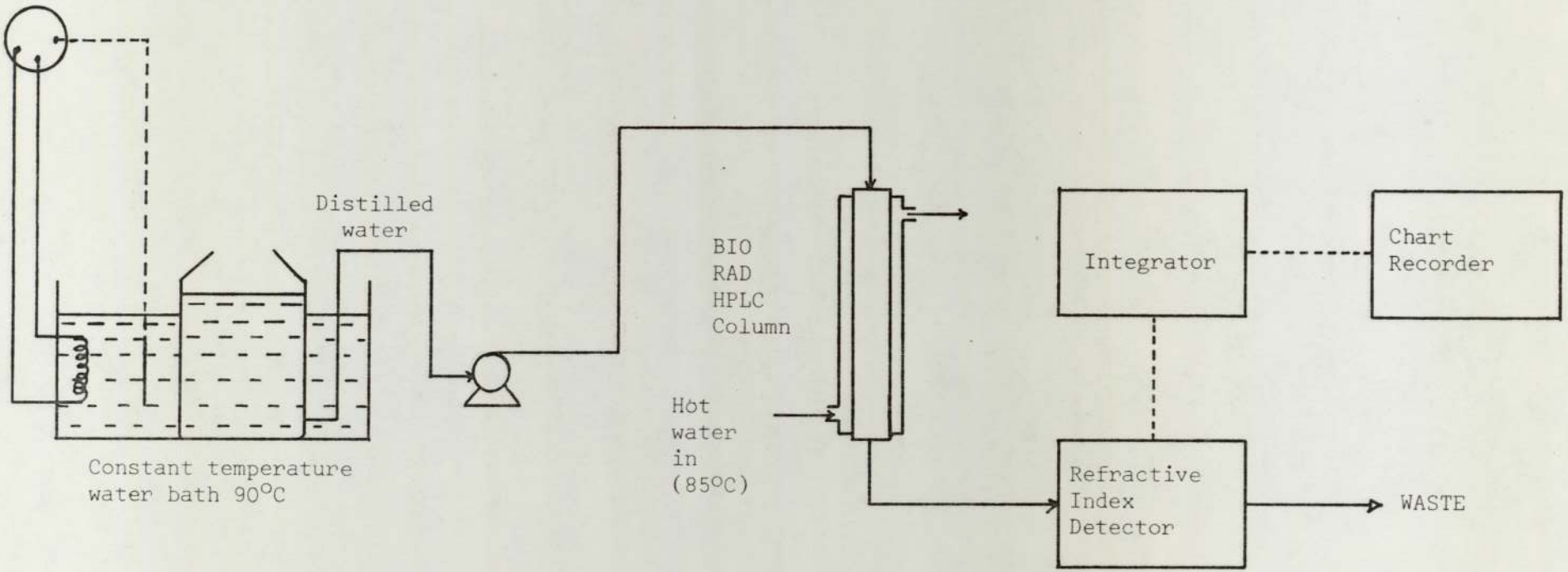


FIG. 5.3 THE ANALYTICAL SYSTEM

CHAPTER 6

THE CONTINUOUS REFINING OF FRUCTOSE AND GLUCOSE

6.1 Commissioning run

Using equation 5.7 as a guide

$$K_{dg} < \frac{L_e}{P} < K_{df} \qquad 5.7$$

and the average values of K_{dg} and K_{df} as seen in Table 4.4 the flow rates and conditions selected for the commissioning run were as shown in Table 6.1.

All the experimental work in the past reported by Ching (6) on the 25.4 mm diameter rig had been performed using a feed concentration of 50% solids. It was decided because of the high cost of fructose, to conduct most of the experimental runs in this research programme at a concentration of 20% solids. This was prepared as a 10% w/v solids fructose and 10% w/v glucose solution. The switch period was selected to conform with the guide lines of equation 5.7 and was 30 minutes in length. The run was consequently defined in the following manner:

Run 20-35-105-30

where 20 = feed concentration (% w/v solids)

35 = feed flow rate ($cm^3 min^{-1}$)

105 = eluent flow rate ($cm^3 min^{-1}$)

30 = switch period (mins)

6.1.2 Results and discussion

Using the previously mentioned criteria (see section 5.3.4.4) it was found that pseudo-equilibrium was reached after 7 cycles had been completed i.e. 70 switch periods of 30 mins duration = 35 hrs.

Table 6.1 - Run Conditions for the Commissioning Run

Run Number	Average Flow Rates			Feed Conc % w/v		Switch Period min	Average Pressure Drops			L/P Ratios		
	Eluent cm^3/min	Feed cm^3/min	Purge cm^3/min	F	G		Eluent kNm^{-2}	Feed kNm^{-2}	Purge kNm^{-2}	Pre-Feed	Post-Feed	Av
20-35-105-30	105	35	550	10.0	10.0	30	190	54	61	0.165	0.473	0.320

Results for the Commissioning Run

Run Number	Fructose Product			Glucose Product			Feed T'put kg/hr
	Purity %	Mass Balance %	Product Conc % w/v	Purity %	Mass Balance % w/v	Product Conc % w/v	
20-35-105-30	99.9	98	0.64	99.9	101	2.37	0.42

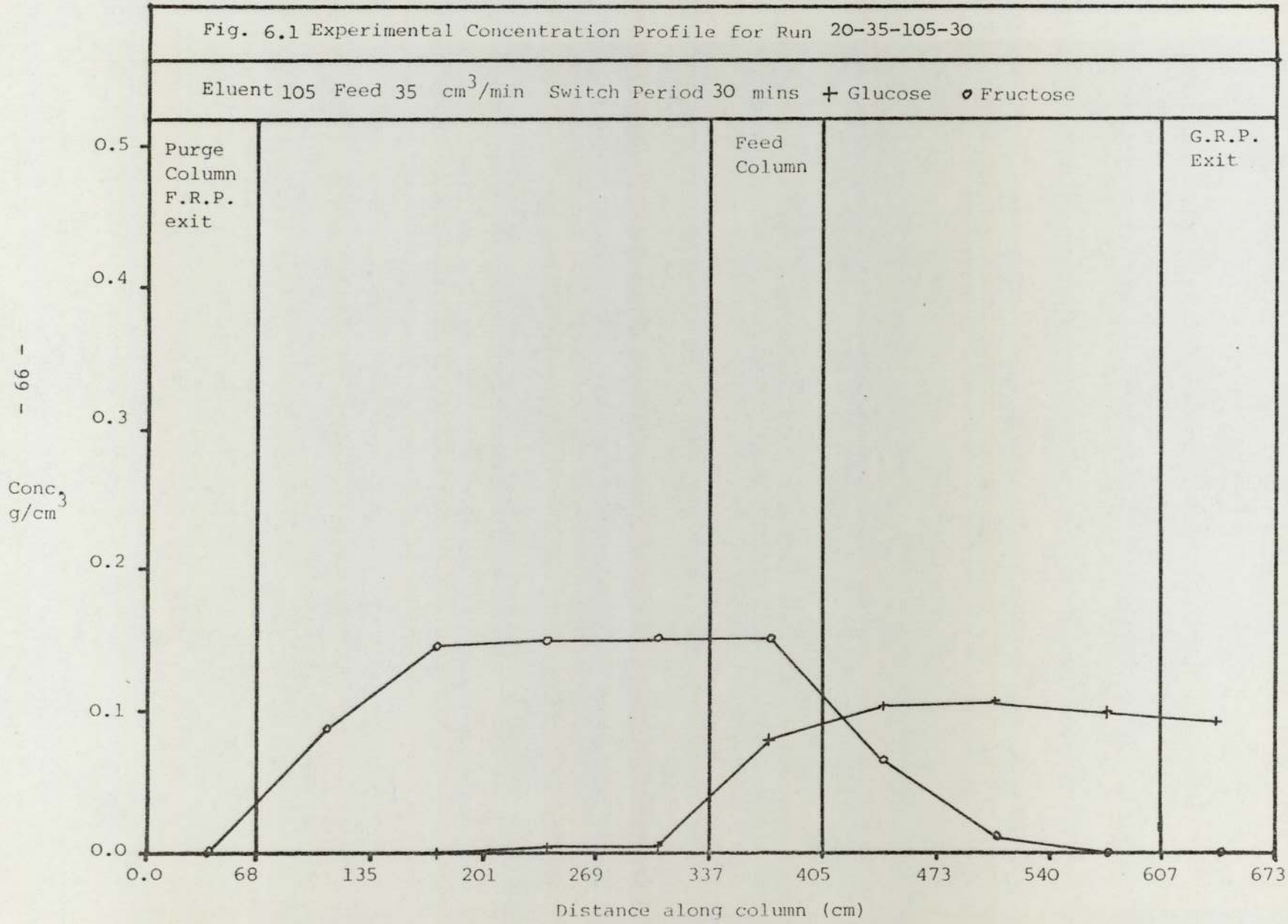


Fig. 6.1 Experimental Concentration Profile for Run 20-35-105-30
Eluent 105 Feed 35 cm3/min Switch Period 30 mins + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
G.R.P. Exit
Conc. g/cm3
0.0
0.1
0.2
0.3
0.4
0.5
0.0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

Two 99.9% pure products were obtained. The concentration of the glucose product was 2.4% solids and the concentration of the fructose product was 0.64% solids. As seen from Fig. 6.1, the end of run purged profile of the concentrations of sugar on each column, the 'cross over' point of the two profiles was just after the feed column. The leading edge of the fructose profile dropped sharply in the post-feed section. This indicated that very little fructose was travelling preferentially with the mobile phase. Thus a substantially fructose-free glucose product was obtained. In the pre-feed section the trailing edge of the glucose profile dropped sharply indicating that very little glucose was travelling preferentially with the stationary phase resulting in a substantially glucose-free fructose product. It is seen that in the eluent entry column the mass of fructose present was significantly lower than in the rest of the pre-feed section. This is because fresh eluent (deionised water) enters this column and even though the fructose has complexed with the calcium ions, some weak bonds are being broken and some fructose is travelling with the mobile phase.

Studying the profiles generally, it is seen that only four column lengths are necessary for separation, the rest containing almost totally resolved products.

This run was considered very successful and believed to be the first ever continuous separation of a fructose-glucose mixture giving pure products without any recycle taking place. It was also considered that in terms of throughput the apparatus offered much greater potential, since only low concentrations of feed had been used.

6.2 Effect of changing the feed concentration

To investigate this parameter two further experimental runs were conducted.

The conditions under which these were performed were identical to run 20-35-105-30 but the concentration of the feed was 40% w/v solids in the first and 60% w/v solids in the second. They were given the identification numbers 40-35-105-30 and 60-35-105-30 respectively.

The summaries of the conditions and results are given in Table 6.2. To highlight the effects of concentration change, the glucose profiles and fructose profiles have been plotted separately, Figs. 6.5 and 6.6 together with the individual profiles Figs. 6.2, 6.3 and 6.4.

6.2.1 Results and discussion

As concentration of the feed increases the profile crossover point moves to the left towards the fructose product offtake. If the fructose profiles alone are studied, Fig. 6.6, it may be seen that only very small quantities of fructose travel preferentially with the mobile phase into the post-feed section. As the concentration of the feed increases so the amount of fructose retained in the columns and consequently product concentration increase.

When the glucose profiles are studied, Fig. 6.5, a marked effect is observed. At 20% solids, i.e. 10% fructose and 10% glucose, only a negligible amount of glucose travels preferentially with the stationary phase - this is likely to be the glucose present in the feed column when the sequencing action occurs. At 40% and

TABLE 6.2 RUN CONDITIONS FOR VARYING FEED CONCENTRATION EXPTS.

RUN NUMBER	AVERAGE FLOW RATES			FEED CONC. % w/v		SWITCH PERIOD min	AVERAGE PRESSURE DROP			L/P RATIOS		
	ELUENT cm^3/min	FEED cm^3/min	PURGE cm^3/min	F	G		ELUENT kNm^{-2}	FEED kNm^{-2}	PURGE kNm^{-2}	PRE FEED	POST FEED	AV
20-35-105-30	105	35	550	10.0	10.0	30	190	55	61	0.165	0.475	0.32
40-35-105-30	105	35	550	19.5	21.0	30	279	95	61	0.165	0.475	0.32
60-35-105-30	105	35	550	29.7	29.2	30	293	136	68	0.165	0.475	0.32

RESULTS FOR VARYING FEED CONCENTRATION EXPTS

RUN NUMBER	FRUCTOSE PRODUCT			GLUCOSE PRODUCT			FEED T'PUT kg/hr
	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	
20-35-105-30	99.9	98	0.64	99.9	101	2.37	0.42
40-35-105-30	88	97	1.29	99.9	96	3.60	0.84
60-35-105-30	77	86	1.79	99.9	93	5.00	1.26

Fig. 6.2 Experimental Concentration Profile for Run 20-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins Glucose + Fructose o

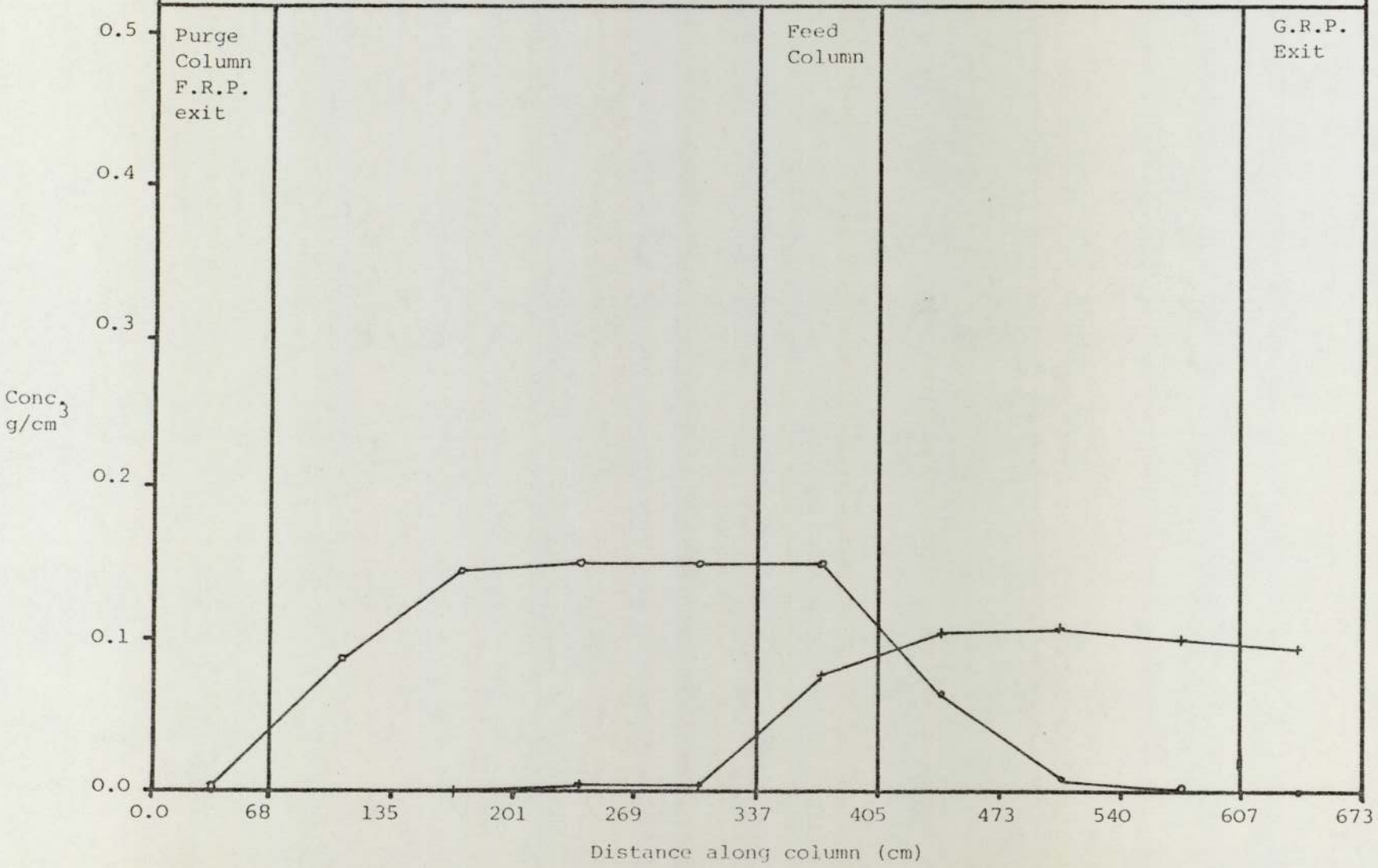


Fig. 6.3 Experimental Concentration Profile for Run 40-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins +Glucose oFructose

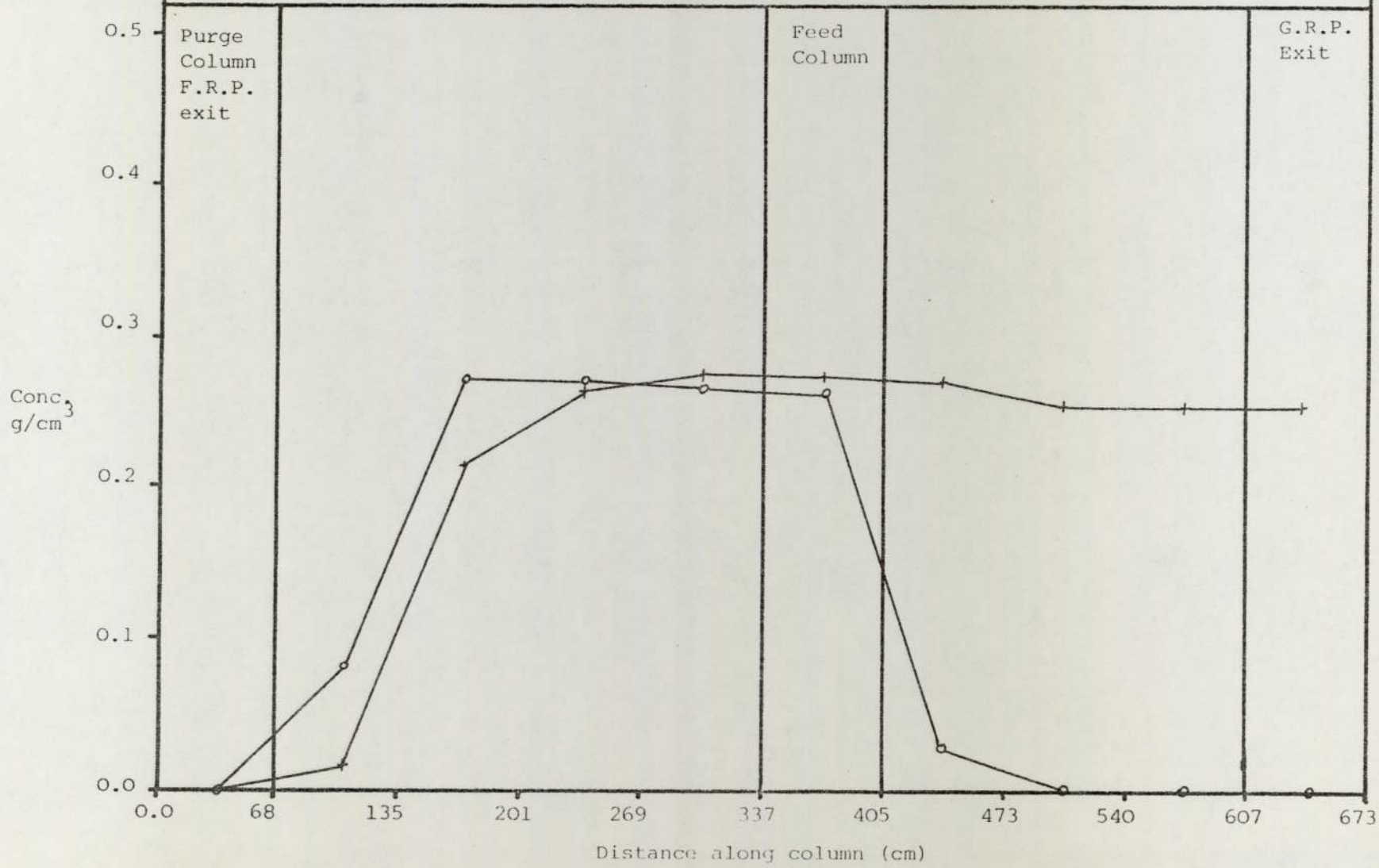


Fig. 6.4 Experimental Concentration Profile for Run 60-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins +Glucose oFructose

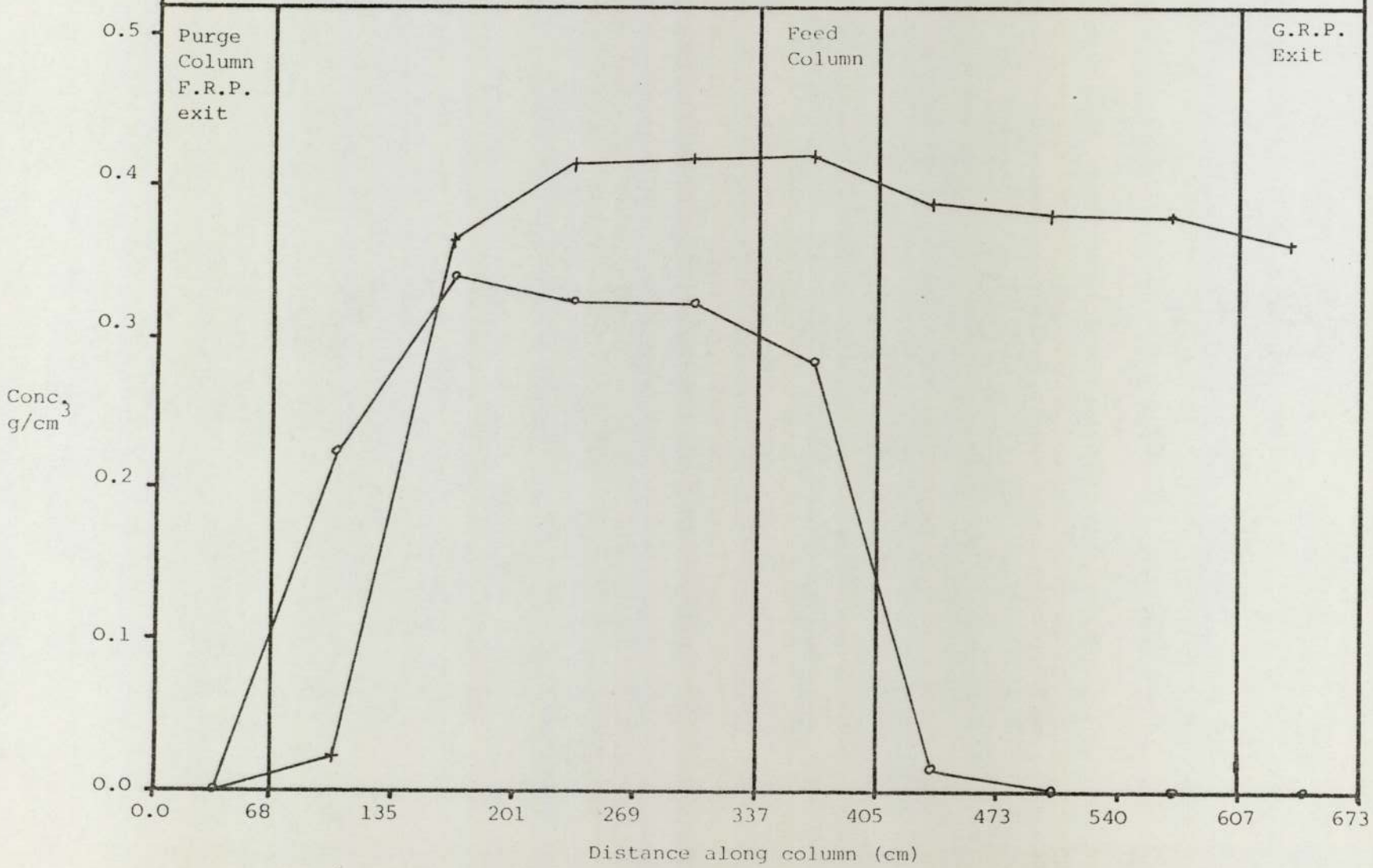


Fig. 6.5 Individual glucose profiles for varying feed concentration experiments

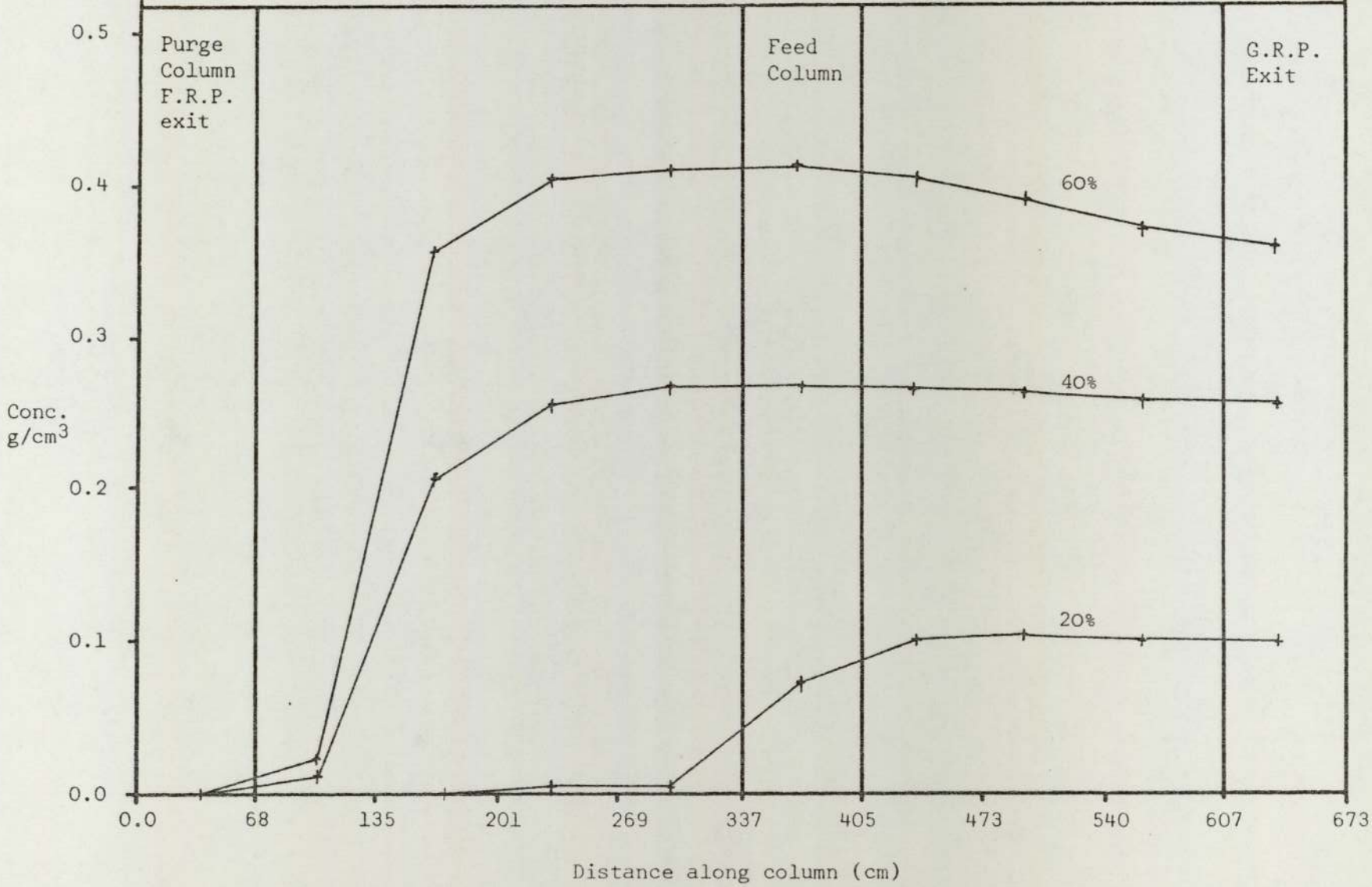
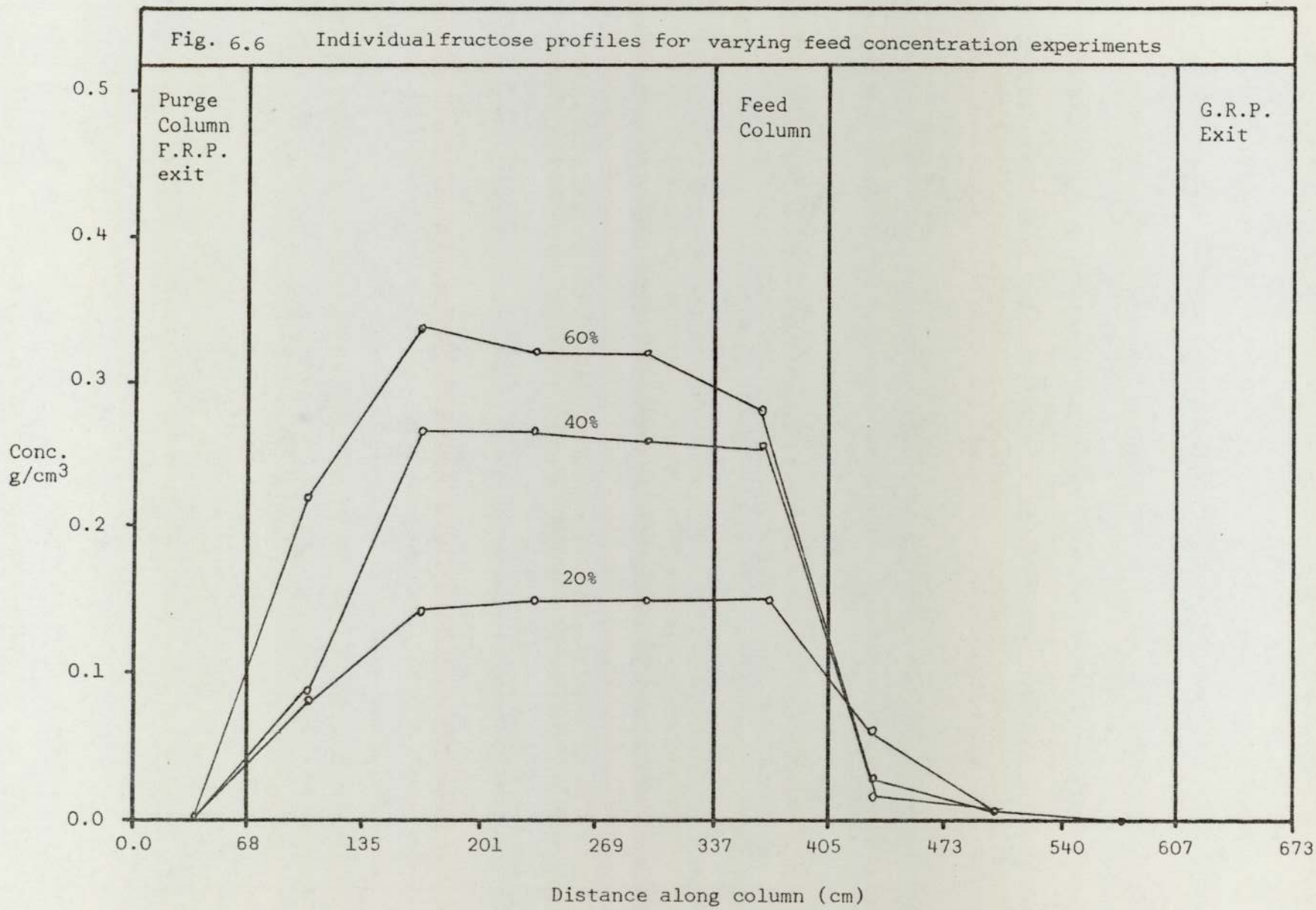


Fig. 6.6 Individual fructose profiles for varying feed concentration experiments



60% solids feed concentration significant amounts of glucose travel preferentially with the stationary phase into the prefeed section.

It is thought that this change in the distribution of the glucose between the mobile and stationary is due to a number of factors.

In dilute solution there are two 6-membered ring structures of the glucose molecules which are in equilibrium. These are the α-D-glucopyranose and the β-D-glucopyranose - the difference being the spatial position of the hydroxyl group next to the oxygen atom, see Fig. 2.7.

The equilibrium ratio of the two structures is approximately $\frac{1}{3} : \frac{2}{3} :: \alpha : \beta$ in dilute solution (see Table 2.1). When in dilute solution to attain the lowest energy of conformation and thus gain the greatest thermodynamic stability each glucose molecule has between fifteen and twenty molecules of water associated with it. These water molecules hydrate or solvate the sugar molecules. They may be linked through electrostatic dipole interaction or by hydrogen bonds. Under these dilute conditions the glucose structure is not sterically favourable for complexing with calcium ions, see section 2.3.1. Under conditions of higher concentration the number of water molecules in solution which are available for hydration decreases. At 20% solids the ratio of water molecules to sugar molecules is 40:1, at 30% solids it is 24:1, at 40% it is 15:1 and at 60% it is 7:1. This suggests that at concentrations of greater than 30% there will be competition between the sugar molecules for water molecules in an attempt to totally hydrate and achieve the lowest conformational energy.

In this competitive situation various changes are thought possible to occur to the molecular structure of the glucose molecules. These are:

1. Weak bonding may occur between adjacent groups on the same glucose molecule - two hydroxyl groups "sharing" the same water molecule.

2. The calcium ions on the resin are themselves hydrated with six water molecules per ion, and these may also be used by the sugars for hydration purposes.

3. Water molecules already associated with a complexed fructose molecule may also be "shared" with a glucose molecule.

4. As a result of 1,2 and 3 above, the stereo chemistry of glucose may be altered to produce favourable structures of hydroxyl groups for complexing with the calcium ions thus causing the glucose to move with the stationary phase.

A gel permeation effect is also likely to play a part in the retardation of all sugar molecules. This happens as the 'active sites' are used for chemisorption the physical volume available for free flow within the resin will decrease. A diffusion effect will also be in operation in attracting incompletely hydrated sugar molecules towards the eluent input column where free water is being pumped in.

The level of glucose in the eluent entry columns is significantly lower than that of the fructose. This is because the fructose is still able to form the complex with the calcium ions and be retarded, but because of the abundance of water molecules from the eluent stream the glucose becomes totally hydrated and the structures are not favourable for complexing. Thus it travels preferentially with the mobile phase.

In conclusion it can be stated that at 20% solids feed concentration total continuous separation is achieved. As the feed concentration increases it is thought that the structure of the glucose molecules changes due to the lack of free water molecules available for total hydration. These structures so formed are more likely to form complexes with the calcium ions and hence travel into the pre-feed section with the stationary phase. Thus contamination of the fructose product results.

6.3 Effect of changing the feed rate

Three experimental runs were conducted to assess this parameter: Run 20-35-105-30, Run 20-45-105-30 and Run 20-55-105-30. The conditions and results of this set of experiments are shown in Table 6.3. The end of run concentration profiles of all column sugars are seen in Figs. 6.7, 6.8 and 6.9. The glucose and fructose profiles have also been plotted individually to allow comparison, Figs. 6.10 and 6.11.

6.3.1 Results and Discussion

It is observed from Table 6.3 that when the feed rate increases, the $\frac{L_e}{P}$ ratio in the post-feed section increases. Thus the linear velocity of the mobile phase in that section increases as the stationary phase flow rate remains constant. A direct result of the feed rate increasing is to increase the total amount of sugar being fed into the equipment. It is apparent that as the feed rate increases the 'crossover point' of the profiles moves towards the glucose product outlet.

TABLE 6.3 RUN CONDITIONS FOR VARYING FEED RATE EXPTS

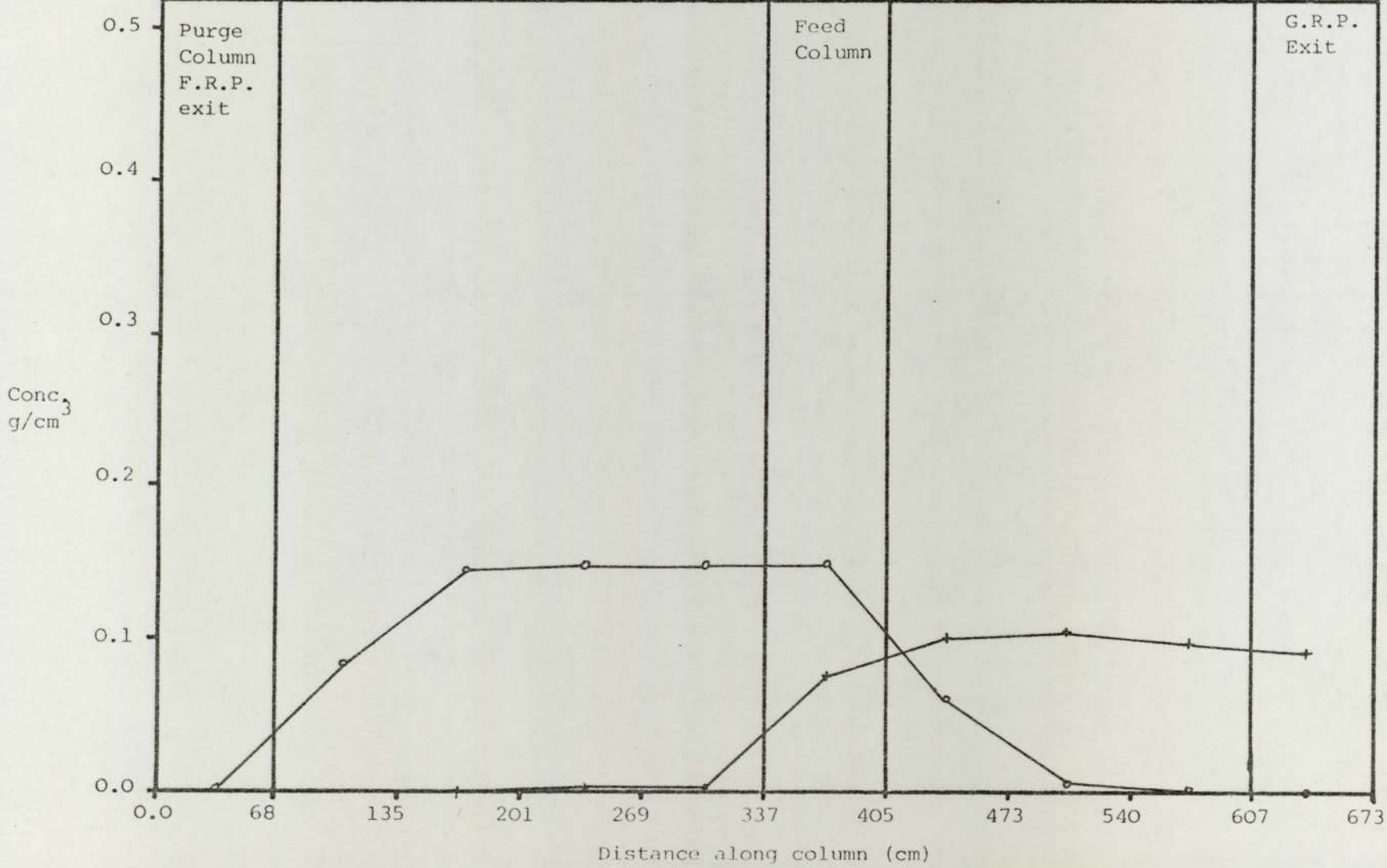
RUN NUMBER	AVERAGE FLOW RATES			FEED CONC. % w/v		SWITCH PERIOD min	AVERAGE PRESSURE DROP			L/P RATIOS		
	ELUENT cm^3/min	FEED cm^3/min	PURGE cm^3/min	F	G		ELUENT kNm^{-2}	FEED kNm^{-2}	PURGE kNm^{-2}	PRE FEED	POST FEED	AV
20-35-105-30	105	35	550	10.0	10.0	30	190	54	61	0.165	0.475	0.320
20-45-105-30	105	45	550	9.6	10.2	30	231	82	61	0.165	0.564	0.364
20-55-105-30	105	55	550	9.8	9.9	30	259	109	61	0.165	0.630	0.400

RESULTS FOR VARYING FEED RATE EXPTs

RUN NUMBER	FRUCTOSE PRODUCT			GLUCOSE PRODUCT			FEED T'PUT kg/hr
	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	
20-35-105-30	99.9	98	0.64	99.9	101	2.37	0.42
20-45-105-30	99.9	102	0.87	99.9	108	3.0	0.54
20-55-105-30	99.9	96	0.94	94	97	3.1	0.66

Fig. 6.7 Experimental Concentration Profile for Run 20-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins + Glucose o Fructose



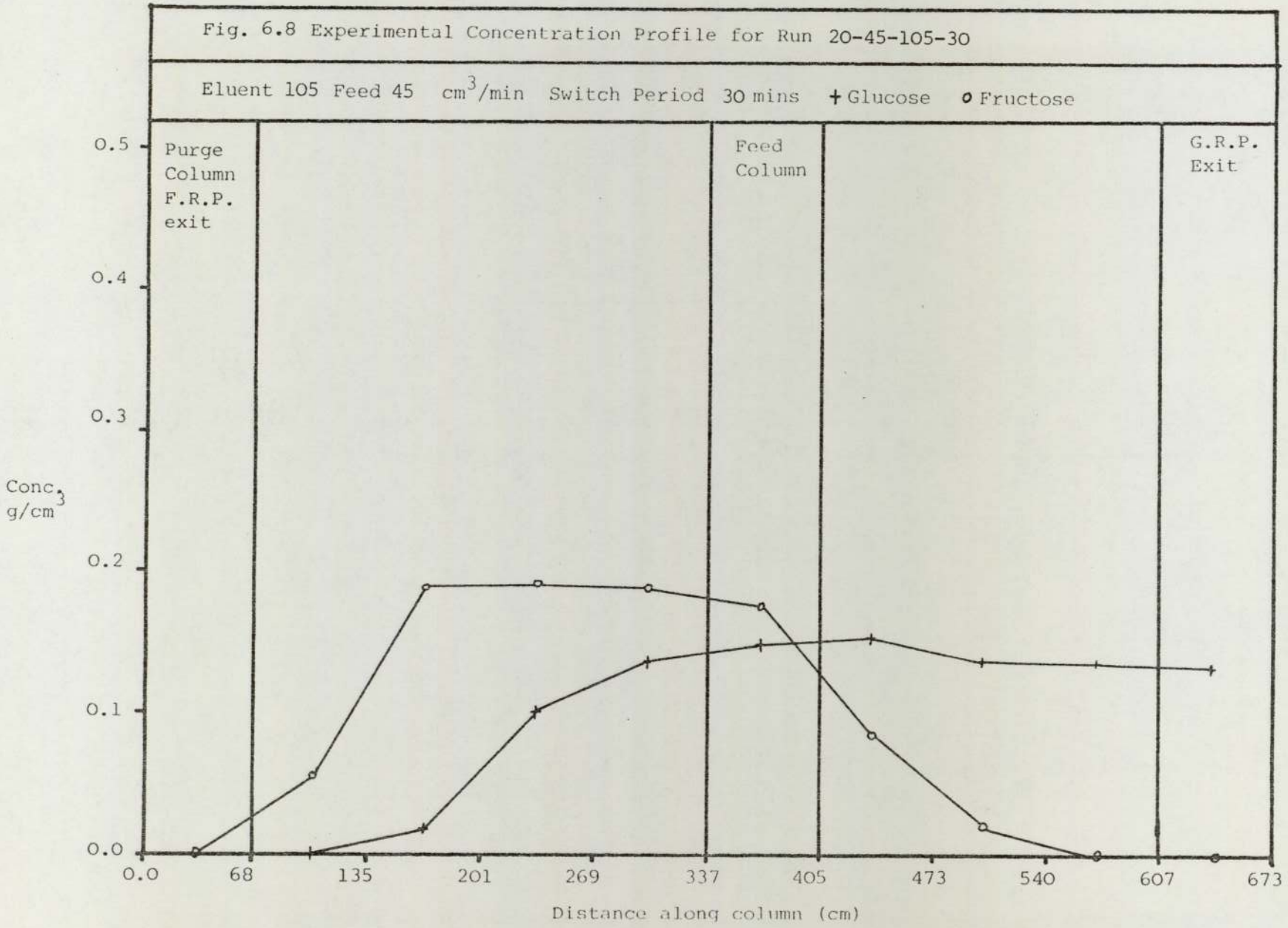


Fig. 6.8 Experimental Concentration Profile for Run 20-45-105-30
Eluent 105 Feed 45 cm3/min Switch Period 30 mins + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
G.R.P. Exit
Conc, g/cm3
0.5
0.4
0.3
0.2
0.1
0.0
0.0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

Fig. 6.9 Experimental Concentration Profile for Run 20-55-105-30

Eluent 105 Feed 55 cm^3/min Switch Period 30 mins + Glucose o Fructose

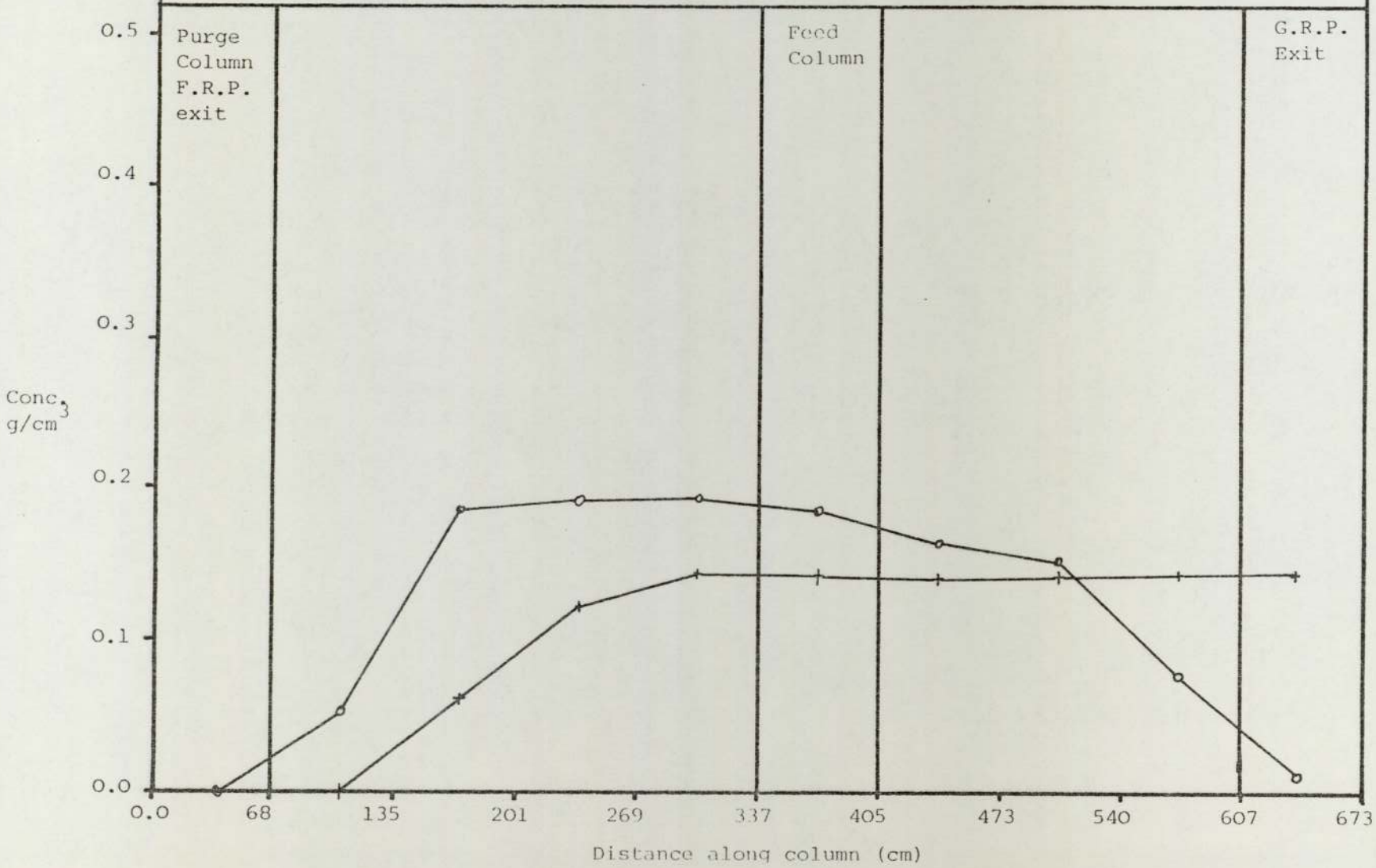


Fig. 6.10 Individual glucose profiles for varying feed flow rate experiments

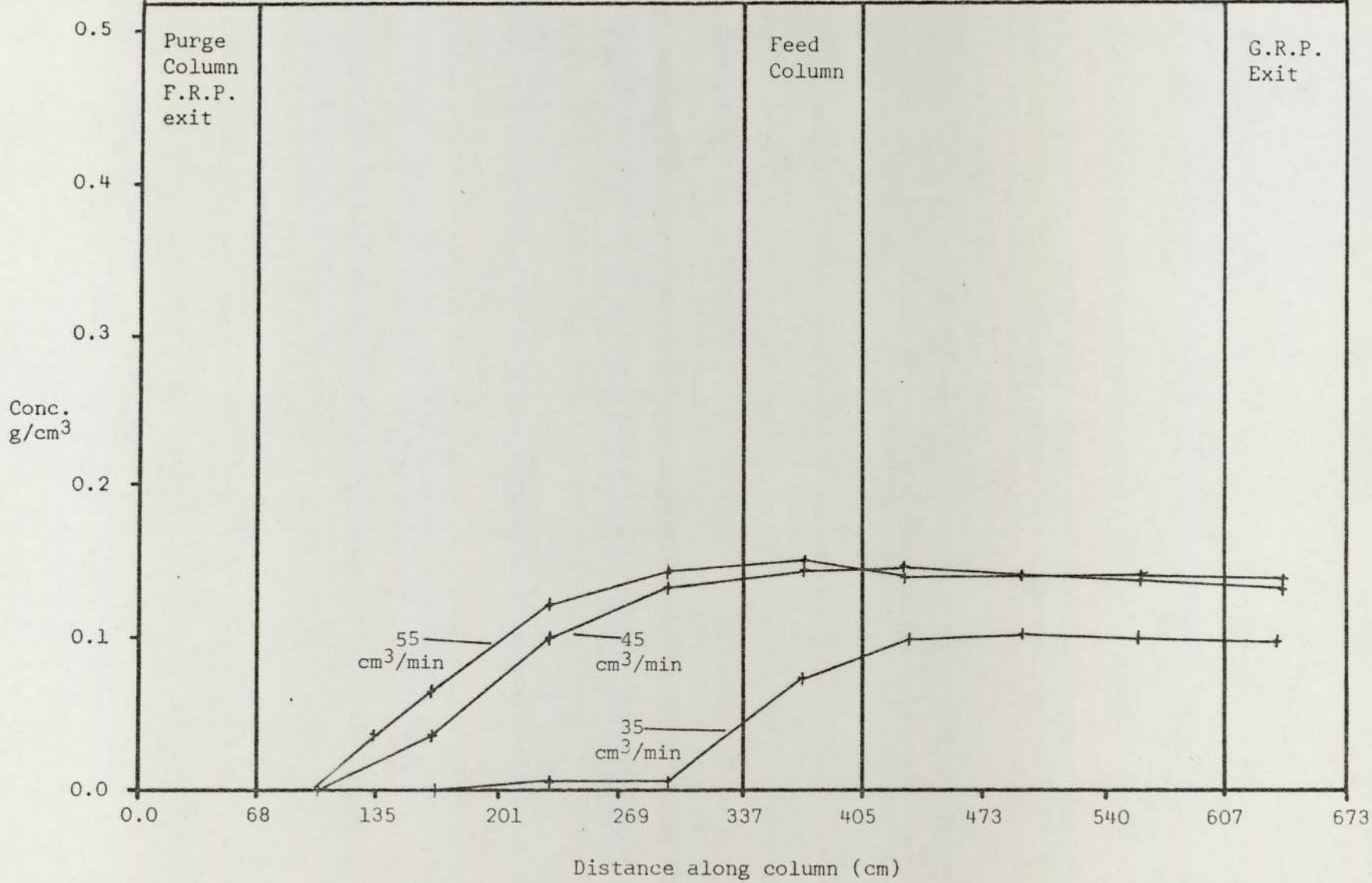
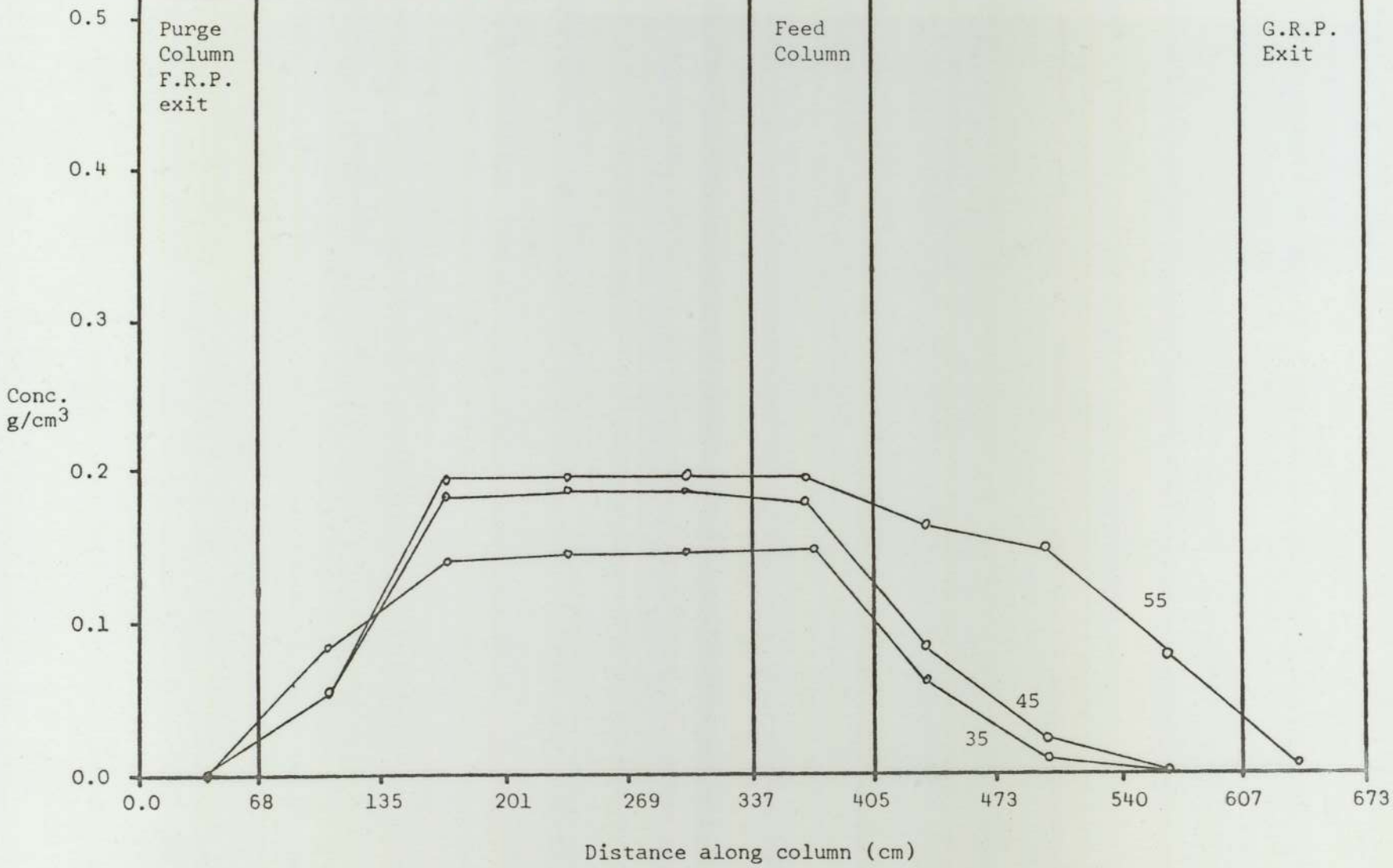


Fig. 6.11 Individual fructose profiles for varying feed flow rate experiments



If the glucose profiles are studied (Fig. 6.10) the results for Run 20-35-105-30 show that very little glucose travels with the stationary phase towards the pre-feed section. As the feed rate increases and the quantity of sugar being input to the system rises so the amount of water available for hydration of the sugar molecules decreases. This is thought to give rise to concentration effects as described in section 6.2 which makes the structure of the glucose molecule more favourable for complexing with the calcium ions: thus a movement with the stationary phase. No detectable glucose contaminates the fructose product however which is 99.9% pure.

If the fructose profiles are studied, Fig. 6.11, it is seen that as the feed rate increases more fructose tends to travel with the mobile phase into the post-feed section. This is due to the increase in linear velocity of the mobile phase. This is enough to begin to break the complex formed between the fructose and the calcium ion and move the equilibrium position of the fructose towards the mobile phase. In the fructose profile of run 20-55-105-30, Fig. 6.9, it may be seen that the leading edge of the profile has travelled into the glucose product exit column, thus contaminating the glucose product which was only 94% pure.

Summarising, at an eluent to feed ratio of 3:1 at 20% solids feed total separation is achieved in approximately four column lengths. As the feed rate increases by 28% and the eluent to feed ratio drops to 2.33:1 the concentration effect causes some glucose to travel with the mobile phase. Two pure products are still obtained and approximately seven column lengths are used to achieve separation. As the feed rate increases by a further 22% and the eluent to feed ratio drops to 2:1, glucose travels with the stationary phase owing to the concentration effect and fructose travels with the mobile phase

due to the increase in linear velocity of this phase.

6.4 Effect of changing the switch period

In an attempt to assess the possible throughput of the rig three experiments were performed with a decreasing switch period i.e. 30 mins, 20 mins and 15 mins. The flow rates were correspondingly increased so that the $\frac{L_e}{P}$ ratio in the pre-feed and post feed sections remained the same for each run. All the experiments were conducted using an eluent to feed ratio of 3:1 and a feed concentration of 20% solids. The conditions of the experiments Run 20-35-105-30, Run 20-52-160-20 and Run 20-70-210-15 are shown with the results obtained in Table 6.4. The end of run concentration profiles are shown in Figs. 6.12, 6.13 and 6.14 with the separate glucose profiles shown in Fig. 6.15 and fructose profiles in Fig. 6.16.

6.4.1 Results and discussion

Runs 20-35-105-30 and 20-52-160-20 were both completed in 9 cycles. During run 20-70-210-15 problems were encountered with the equipment. This occurred when the sequencing of the valves took place. As was explained in section 3.5.1 the feed pump was a twin-headed micro-pump operating at 76 strokes per minute. The two heads were set at 180^{o} out of phase with each other to create a smoother flow regime. This effectively doubled the number of strokes of the pump per minute. While operating at the high feed flowrate of 70 cm^3 per minute the pressure build up in the feed line as the sequencing took place i.e. one feed valve closed and the next feed valve opened, was too great for the actuating air pressure to open the next feed valve. Consequently the feed line blew off

TABLE 6.4 RUN CONDITIONS FOR VARYING SWITCH PERIOD EXPTS

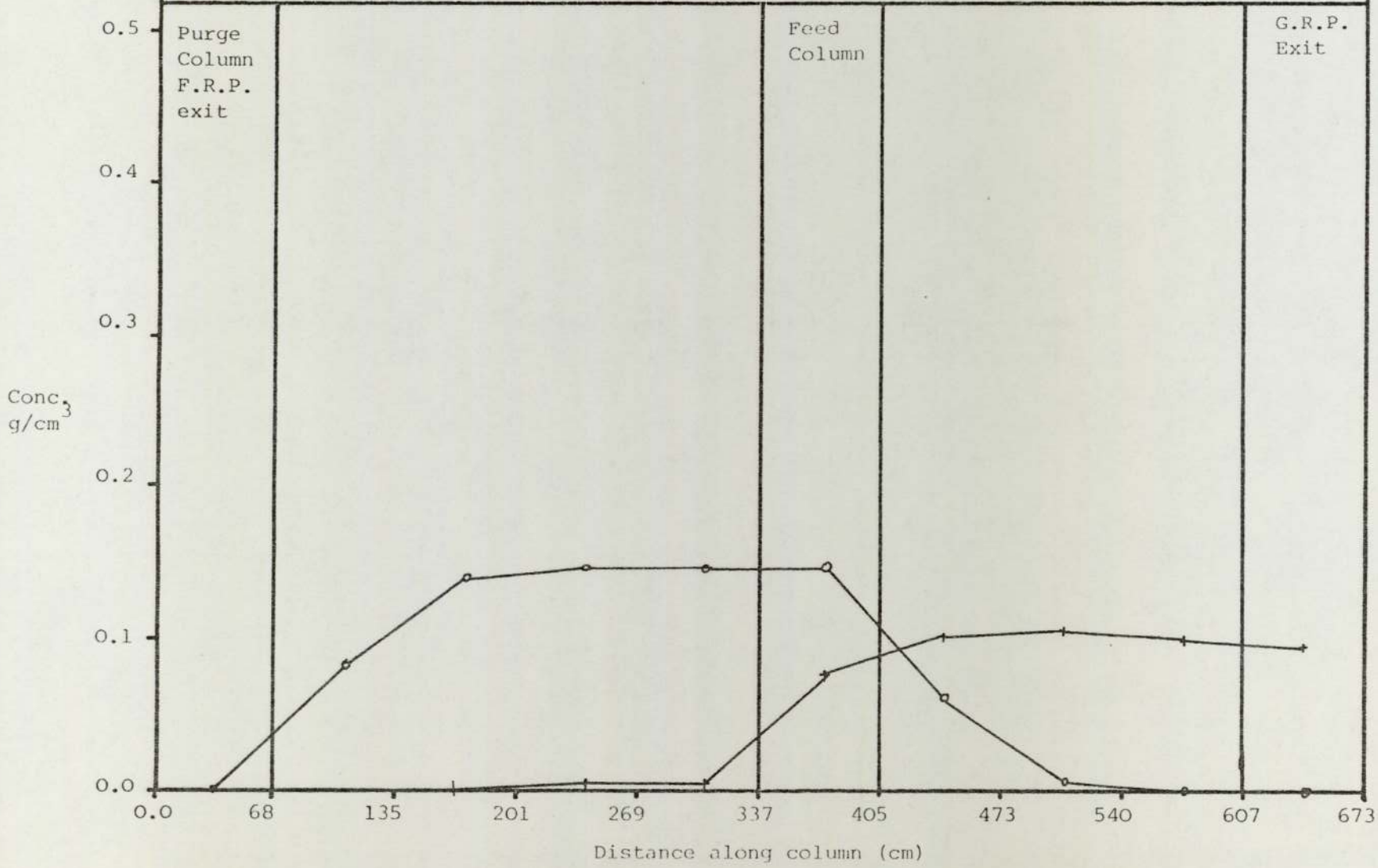
RUN NUMBER	AVERAGE FLOW RATES			FEED CONC. % w/v		SWITCH PERIOD min	AVERAGE PRESSURE DROP			L/P RATIOS		
	ELUENT cm^3/min	FEED cm^3/min	PURGE cm^3/min	F	G		ELUENT kNm^{-2}	FEED kNm^{-2}	PURGE kNm^{-2}	PRE FEED	POST FEED	AV
20-35-105-30	105	35	550	10.0	10.0	30	190	61	61	0.165	0.475	0.320
20-52-160-20	160	52	550	9.7	10.0	20	286	122	68	0.165	0.475	0.320
20-70-210-15	210	70	550	10.0	10.2	15	503	279	136	0.165	0.475	0.320

RESULTS FOR VARYING SWITCH PERIOD EXPTS

RUN NUMBER	FRUCTOSE PRODUCT			GLUCOSE PRODUCT			FEED T'PUT kg/hr	
	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v		
20-35-105-30	99.9	98	0.64	99.9	101	2.37	0.42	
20-52-160-20	99.9	96	0.81	99.9	96	2.38	0.66	
20-70-210-15	85	105	1.22	99.9	69	1.40	0.84	(RUN INCOMPLETE)

Fig. 6.12 Experimental Concentration Profile for Run 20-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins + Glucose o Fructose



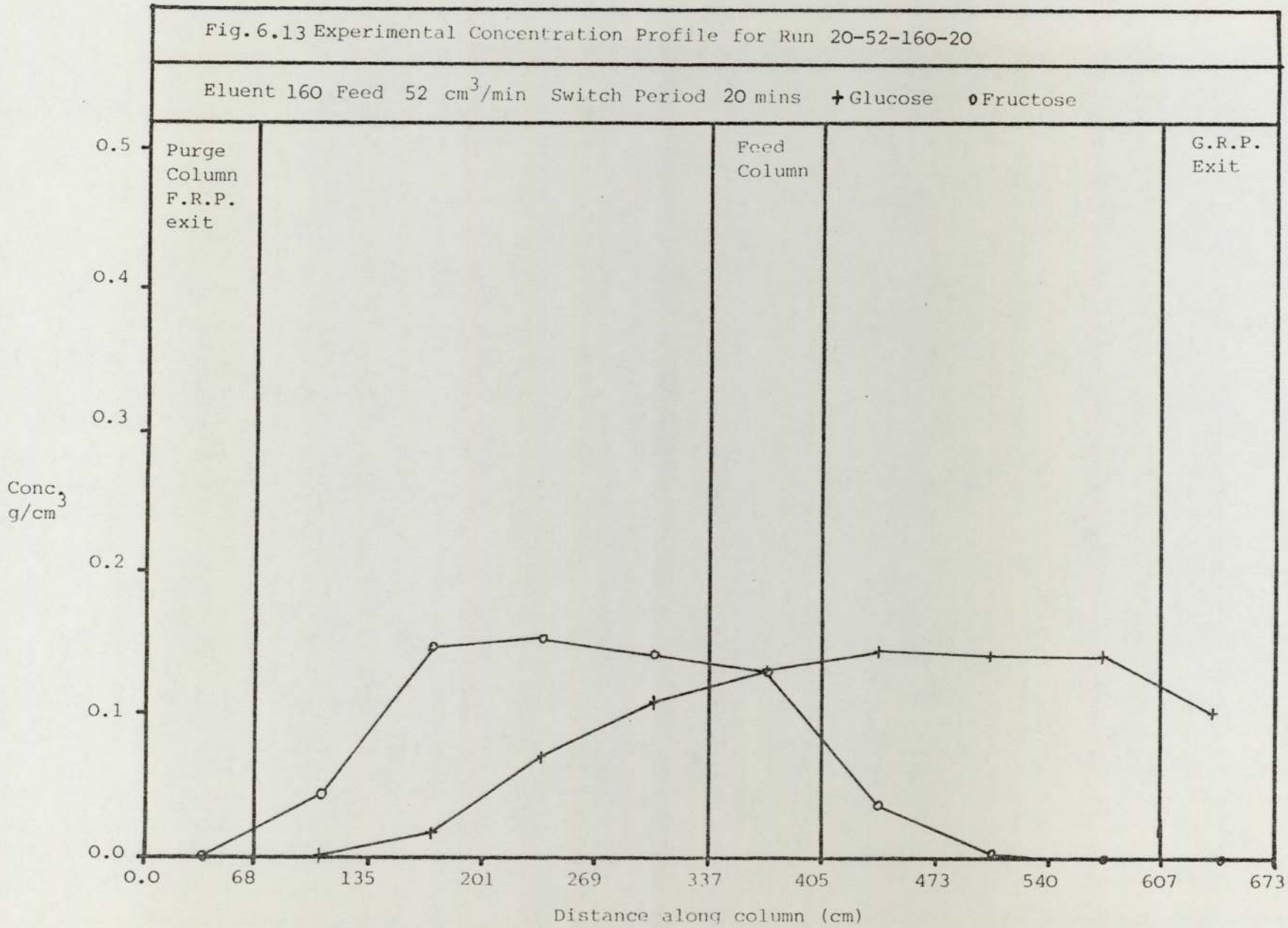
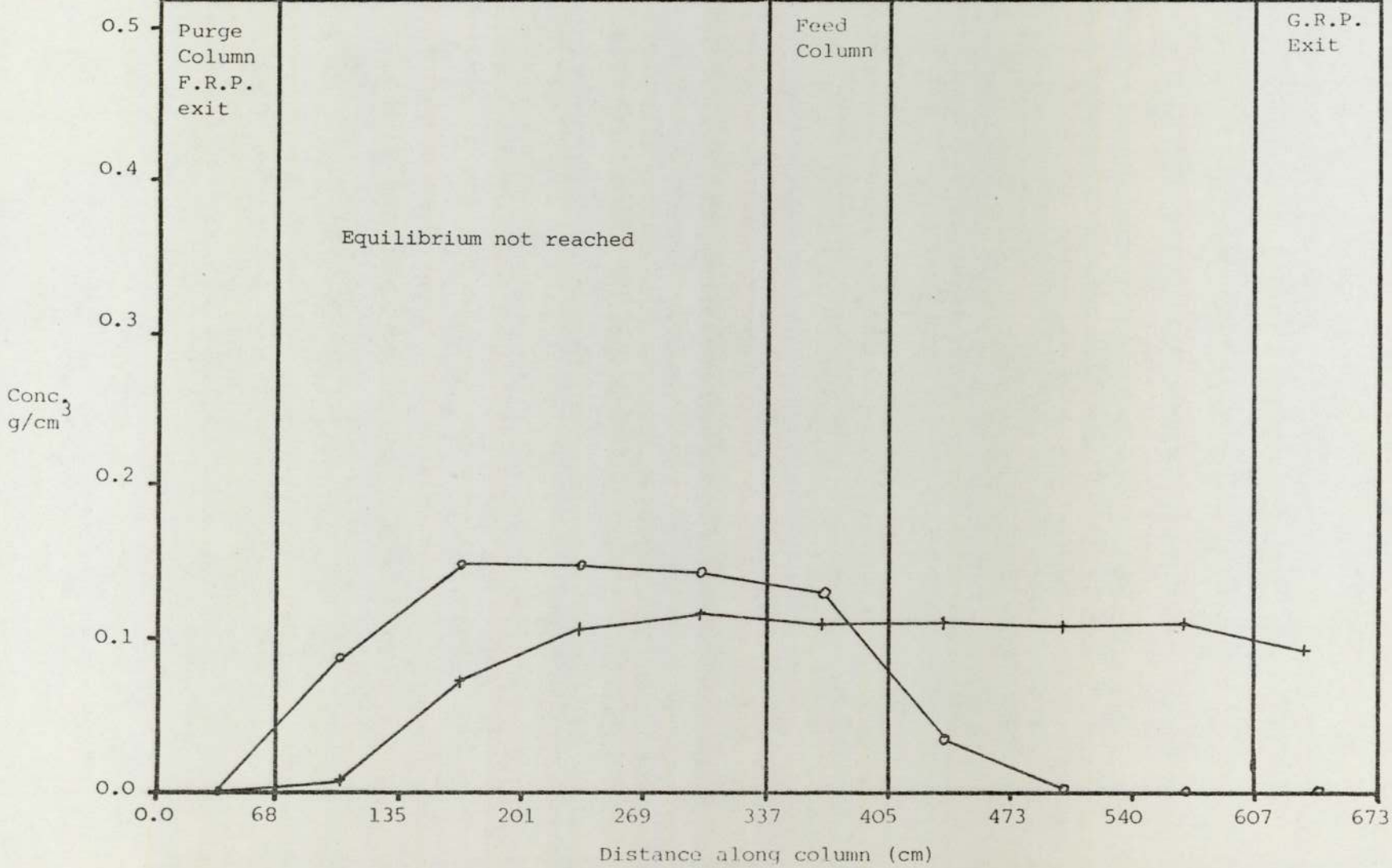


Fig. 6.13 Experimental Concentration Profile for Run 20-52-160-20
Eluent 160 Feed 52 cm3/min Switch Period 20 mins + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
G.R.P. Exit
Conc. g/cm3
0.0
0.1
0.2
0.3
0.4
0.5
0.0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

Fig.6.14 Experimental Concentration Profile for Run 20-70-210-15

Eluent 210 Feed 70 cm^3/min Switch Period 15 mins + Glucose ▫ Fructose



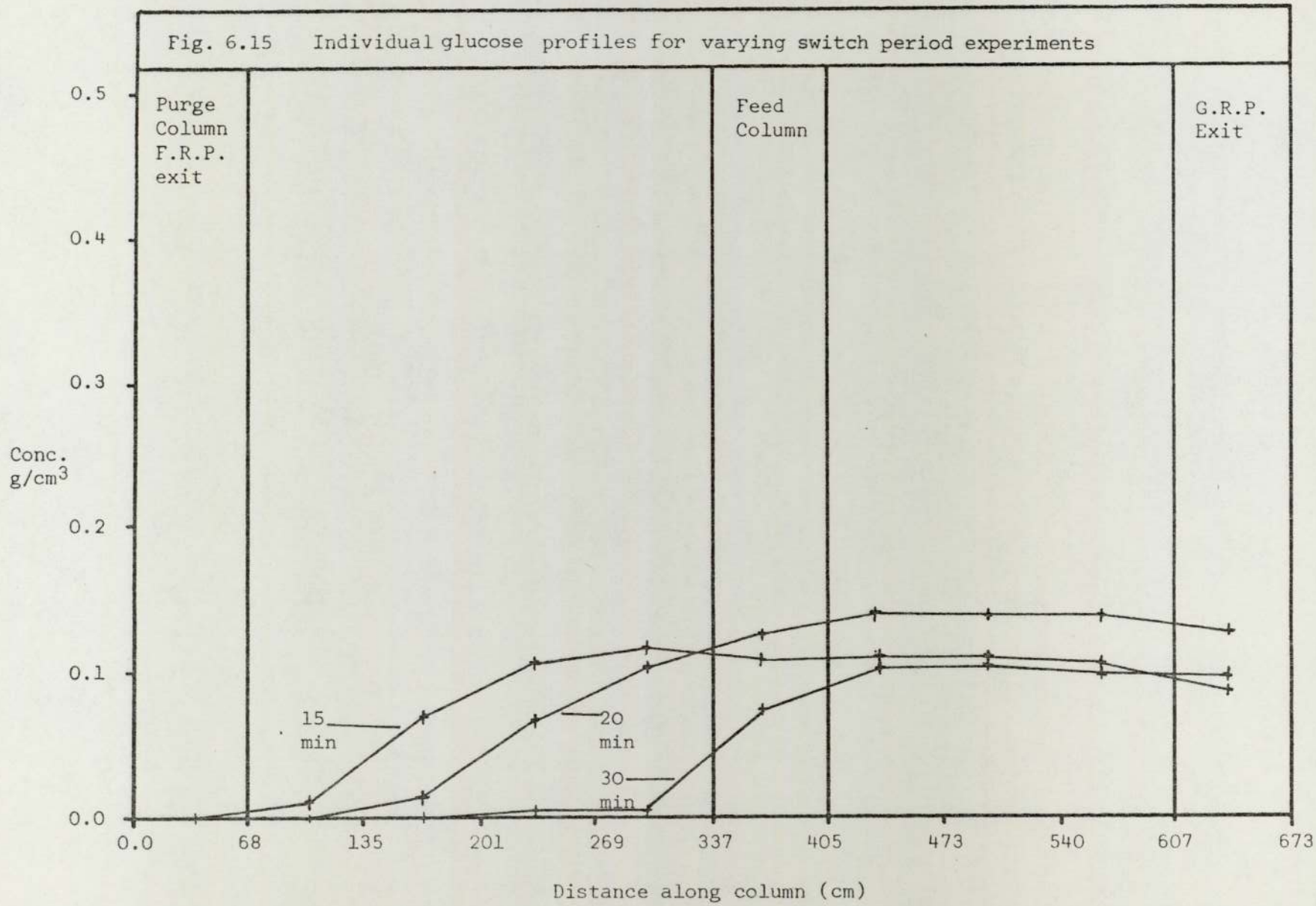


Fig. 6.15 Individual glucose profiles for varying switch period experiments
Purge Column F.R.P. exit
Feed Column
G.R.P. Exit
0.5
0.4
0.3
0.2
0.1
0.0
Conc. g/cm3
15 min
20 min
30 min
0.0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

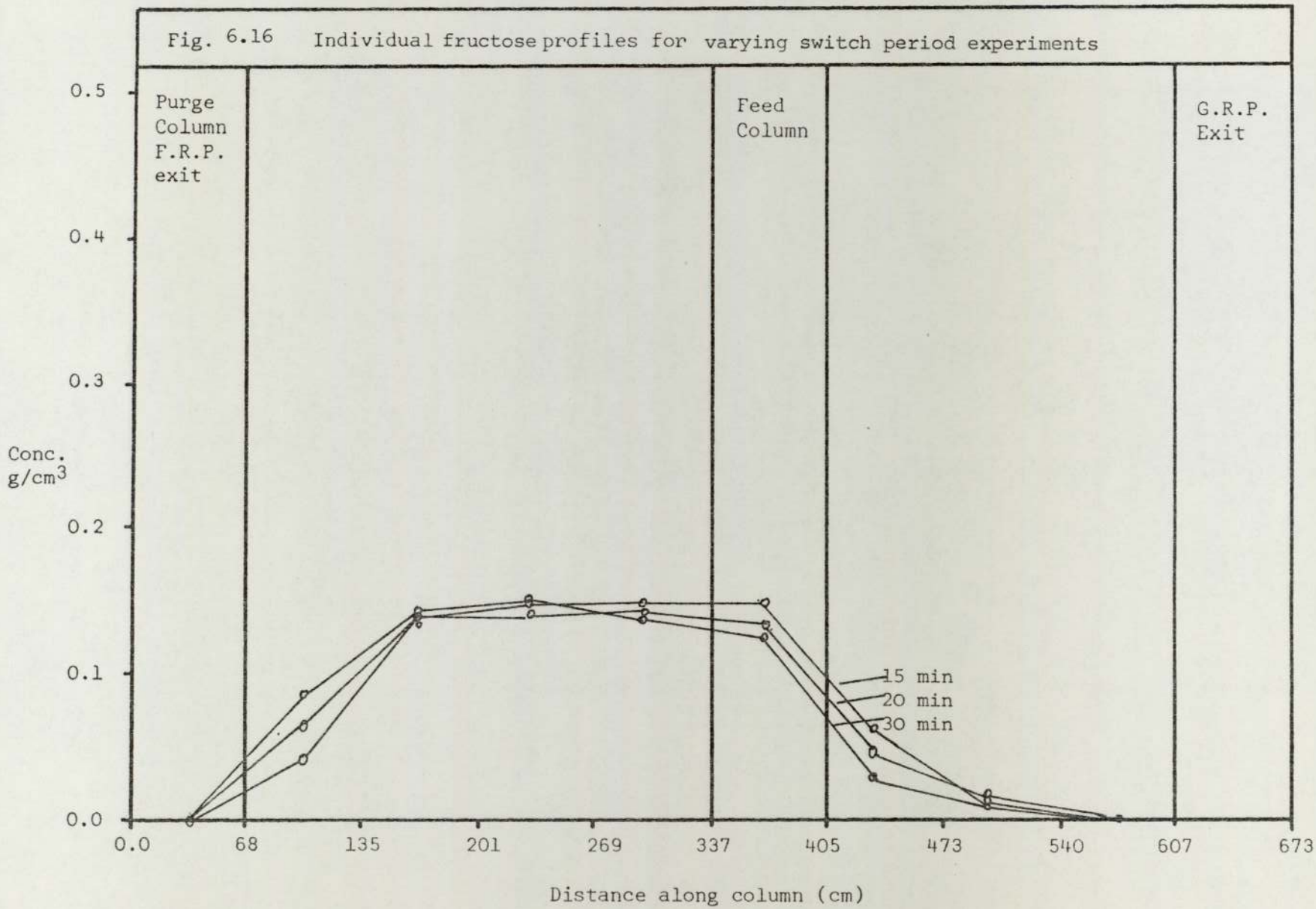


Fig. 6.16 Individual fructose profiles for varying switch period experiments

consistently and the run was abandoned after five cycles. However the columns were purged out and their concentration profile was plotted, Fig. 6.14. From the product concentration measurements and mass balance calculations it was evident that pseudo-equilibrium had not been reached. However some trends were apparent and certain conclusions could be drawn.

From Table 6.4 it may be seen that two pure products resulted from the first two runs with a switch period of 30 minutes and 20 minutes respectively. In the third run with the 15 minute switch time after five cycles glucose had contaminated the fructose product. In all cases the fructose product concentration had increased with increase in feed rate. This is because the volume of purge water per cycle had decreased.

If the fructose profiles are studied, Fig. 6.16 it may be seen that their basic shapes are the same. It is reasonable to assume that had equilibrium been established, the on column concentration of the fructose in the third run would have increased to a similar level as the first two. In all cases very little fructose has travelled preferentially with the mobile phase into the post feed section.

If the glucose profiles are studied, Fig. 6.15, it may be seen that as the switch time decreases more glucose travels preferentially with the stationary phase. This is true even though the eluent to feed ratio remains constant. The mass input of feed per minute however alters markedly with change in the switch time. At 30 minutes per switch period it is 7 gm sugar per minute, at the 20 minute switch period it is 10.5 gm per minute and at the 15 minute switch it is 14 gm per minute. This is clearly approaching

an overloading situation. There is such a high quantity of sugar being fed on to a feed column which only has a finite volume. As the sequencing of the columns takes place more and more glucose is moved with the stationary phase into the pre feed section. This is not being carried towards the glucose outlet at a high enough rate so that at the next switch it has not all been removed and thus a concentration build up occurs. It is thought that if the run 20-70-210-15 had reached equilibrium a greater contamination of the fructose product would have resulted.

6.5 A three-component feed system

To complete the research programme it was decided to test the rig's ability to refine a synthetic feedstock with a similar carbohydrate content to a commercial by-product.

In the manufacture of dextran by Fisons Ltd., Holmes Chapel, a fructose-rich effluent is produced with a carbohydrate composition as shown below. The effluent has a 70% w/v sugar solids content. The typical analysis of the Fisons effluent is:

Fructose	68% of total sugar present
dextran	22% of total sugar present
glucose	8% of total sugar present
Reducing sugars	2% of total sugar present

The commercial feedstock was contaminated by iron, copper, lead, zinc and calcium ions. To avoid any interference with the ion exchange resin in the column the synthetic feedstock was prepared. Three experimental runs were then performed using this feed mixture. Dextran is a polyglucose molecule produced by the fermentation of sucrose.

Chromatograms performed on a single column operating in the batch mode showed that dextran was totally excluded from the pore volume in the resin and moved only in the void volume. Dextran would therefore emerge in the eluent phase with glucose.

The objectives of the three experimental runs were:

1. To recover a substantially pure fructose product.

2. To find the maximum throughput while maintaining fructose purity.

3. To obtain high product concentrations.

4. To assess the rig's ability to operate at high pressure.

Objective 4 was essential because it was known that dextran, being polymeric formed a non-Newtonian fluid. This was found by Chuah (17) to exhibit a very high viscosity which caused a much higher pressure drop than the two-component glucose/fructose feed mixture.

From previous experience it was decided to use a fixed switch period of 30 minutes. The feed rate of the syrup would be varied. The eluent flow rate was therefore set at 105 $cm^3 min^{-1}$. The conditions and results of the three experimental runs D70-35-105-30, D70-45-105-30, D70-55-105-35 are shown in Table 6.5.

6.5.1 Results and discussion

It will be seen from Table 6.5 that all product concentrations increased with feed rate increase. At the highest feed flow rate 2.3 kg of sugars were being processed per hour. In this particular run D70-55-105-30 the pressure drop across the whole SCCR increased to 1,760 kNm^{-2}. During cycle 6 of the run this proved a limiting pressure for the apparatus. The available air supply was unable to

TABLE 6.5 RUN CONDITIONS FOR THE THREE COMPONENT SYSTEM

RUN NUMBER	AVERAGE FLOW RATES			FEED CONC. % w/v			SWITCH PERIOD min	AVERAGE PRESSURE DROP			L/P RATIOS		
	ELUENT cm^3/min	FEED cm^3/min	PURGE cm^3/min	F	G	D		ELUENT kNm^{-2}	FEED kNm^{-2}	PURGE kNm^{-2}	PRE FEED	POST FEED	AV
70-35-105-30D	105	35	550	69.0	9.0	22.0	30	748	326	95	0.165	0.475	0.320
70-45-105-30D	105	45	550	69.0	10.0	21.2	30	1006	551	130	0.165	0.540	0.350
70-55-105-30D	105	55	550	68.5	10.1	22.0	30	>1360	748	170	0.165	0.660	0.415

RESULTS FOR THE THREE COMPONENT SYSTEM

RUN NUMBER	FRUCTOSE PRODUCT			GLUCOSE/DEXTRAN PRODUCT				FEED T'PUT kg/hr	
	PURITY %	MASS BALANCE %	PRODUCT CONC. % w/v	PURITY %	MASS BALANCE %	PRODUCT CONC. G % w/v	PRODUCT CONC. D % w/v		
70-35-105-30D	99.9	104	3.2	99.9	98	1.01	3.8	1.47	
70-45-105-30D	99.9	101	4.4	99.9	99	1.12	4.5	1.89	
70-55-105-30D	99.9	92	5.1	99.9	92	1.40	5.0	2.31	(RUN INCOMPLETE)

Fig. 6.17 Experimental Concentration Profile for Run D70-35-105-30

Eluent 105 Feed 35 cm^3/min. Switch Period 30 mins. + Glucose o Fructose Δ Dextran

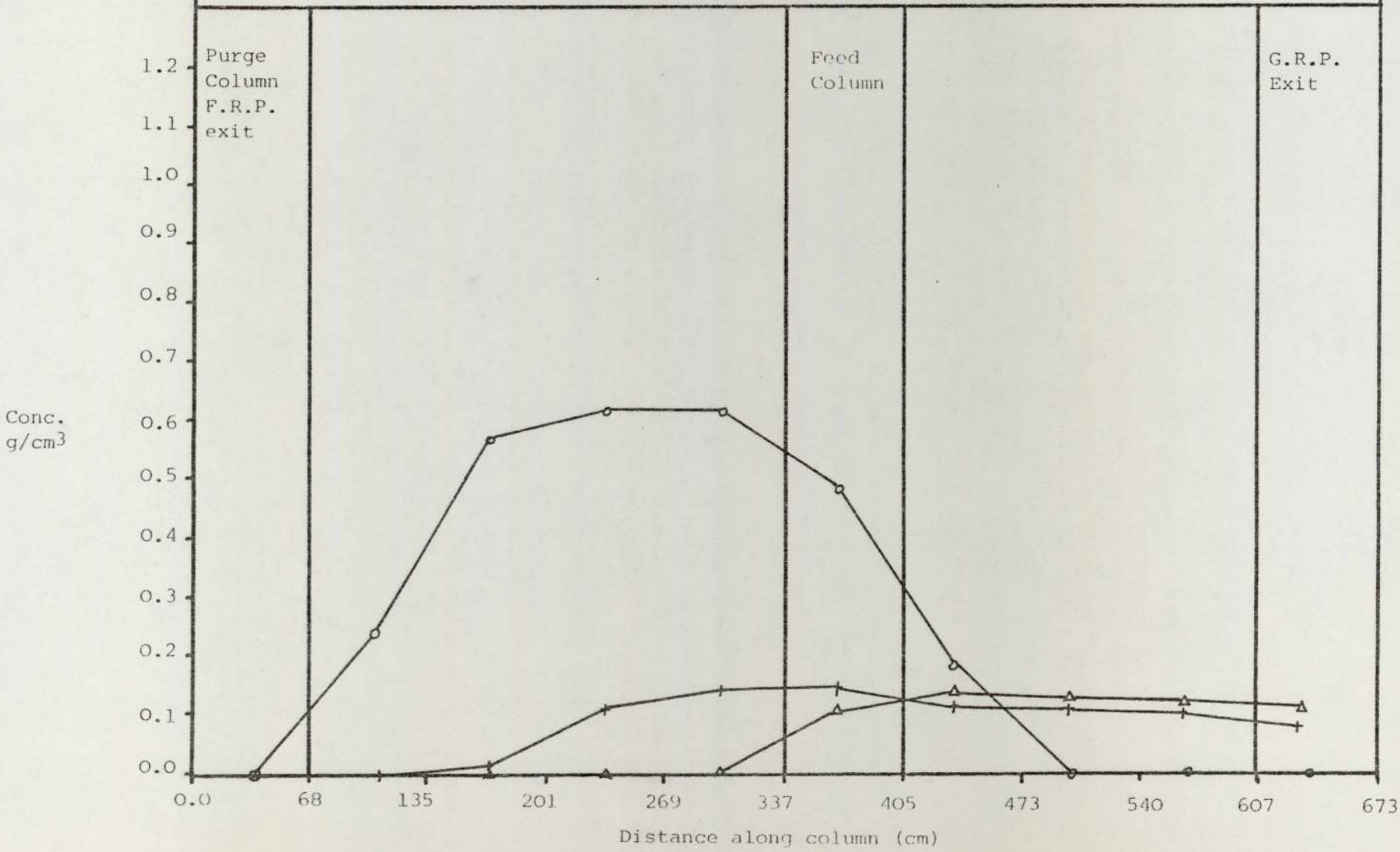


Fig. 6.18 Experimental Concentration Profile for Run D70-45-105-30

Eluent 105 Feed 45 cm^3/min. Switch Period 30 mins. +Glucose ○Fructose △Dextran

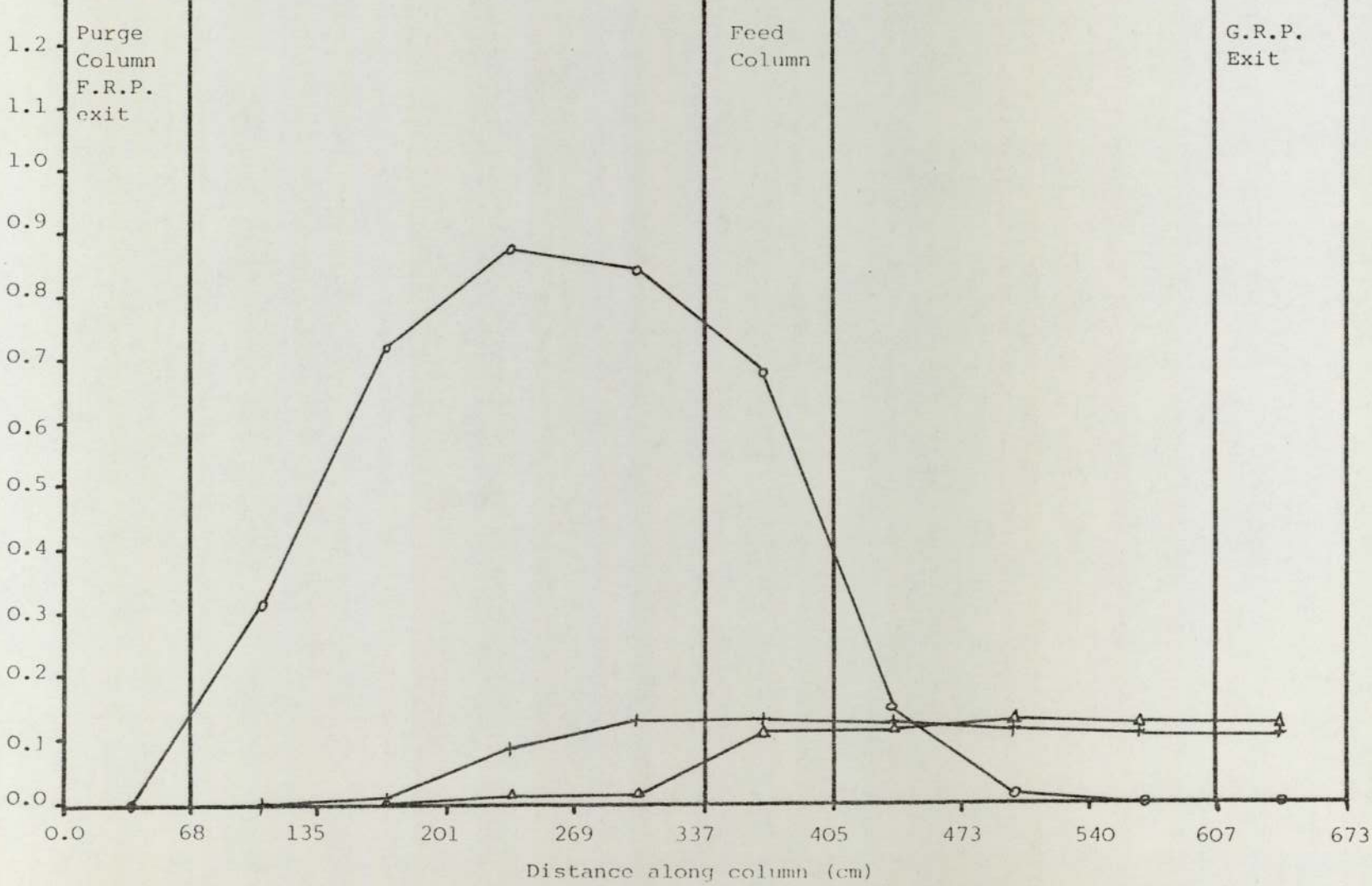
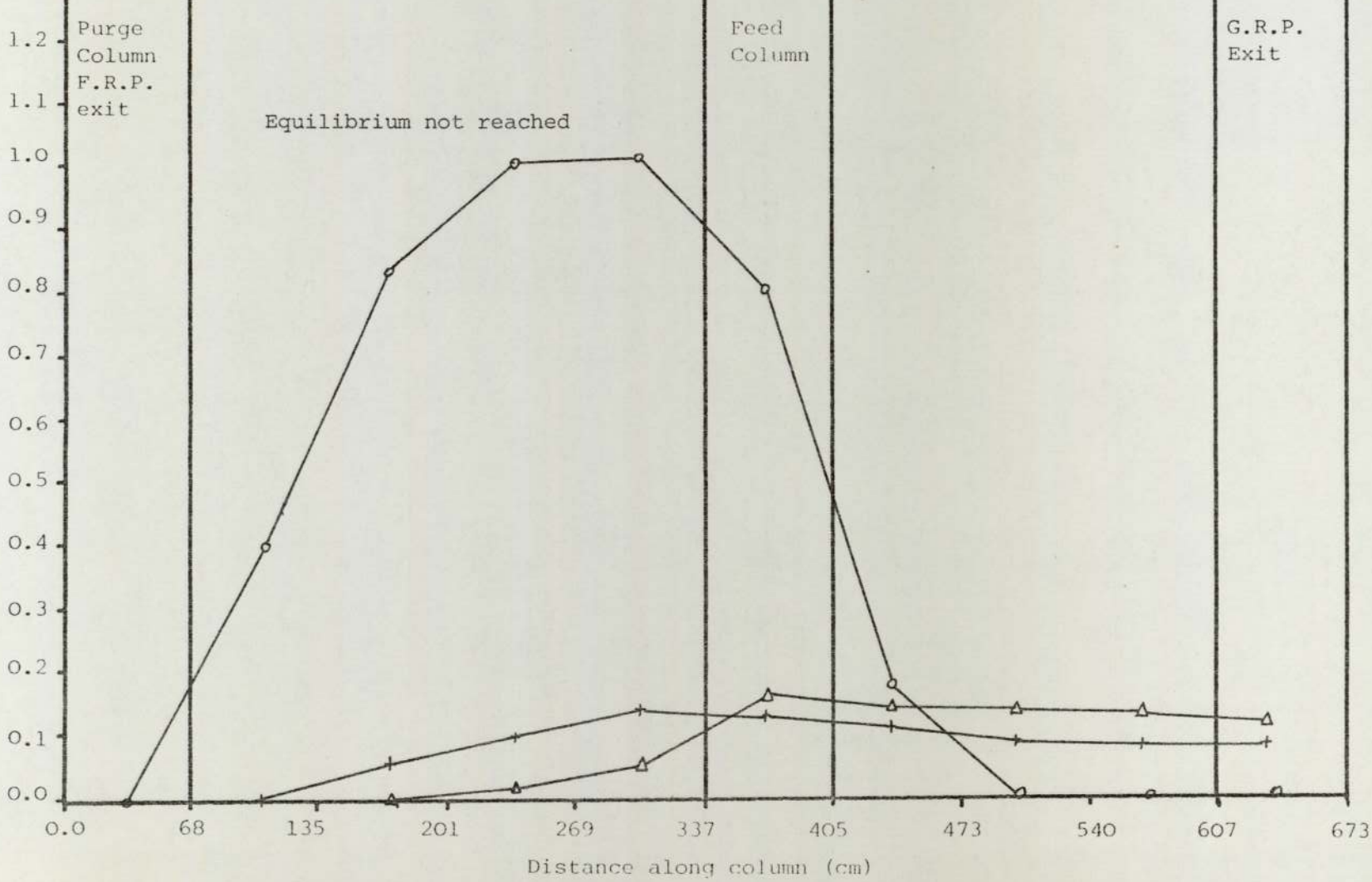


Fig. 6.19 Experimental Concentration Profile for Run D70-55-105-30

Eluent 105 Feed 55 cm^3/min. Switch Period 30 mins. + Glucose o Fructose Δ Dextran



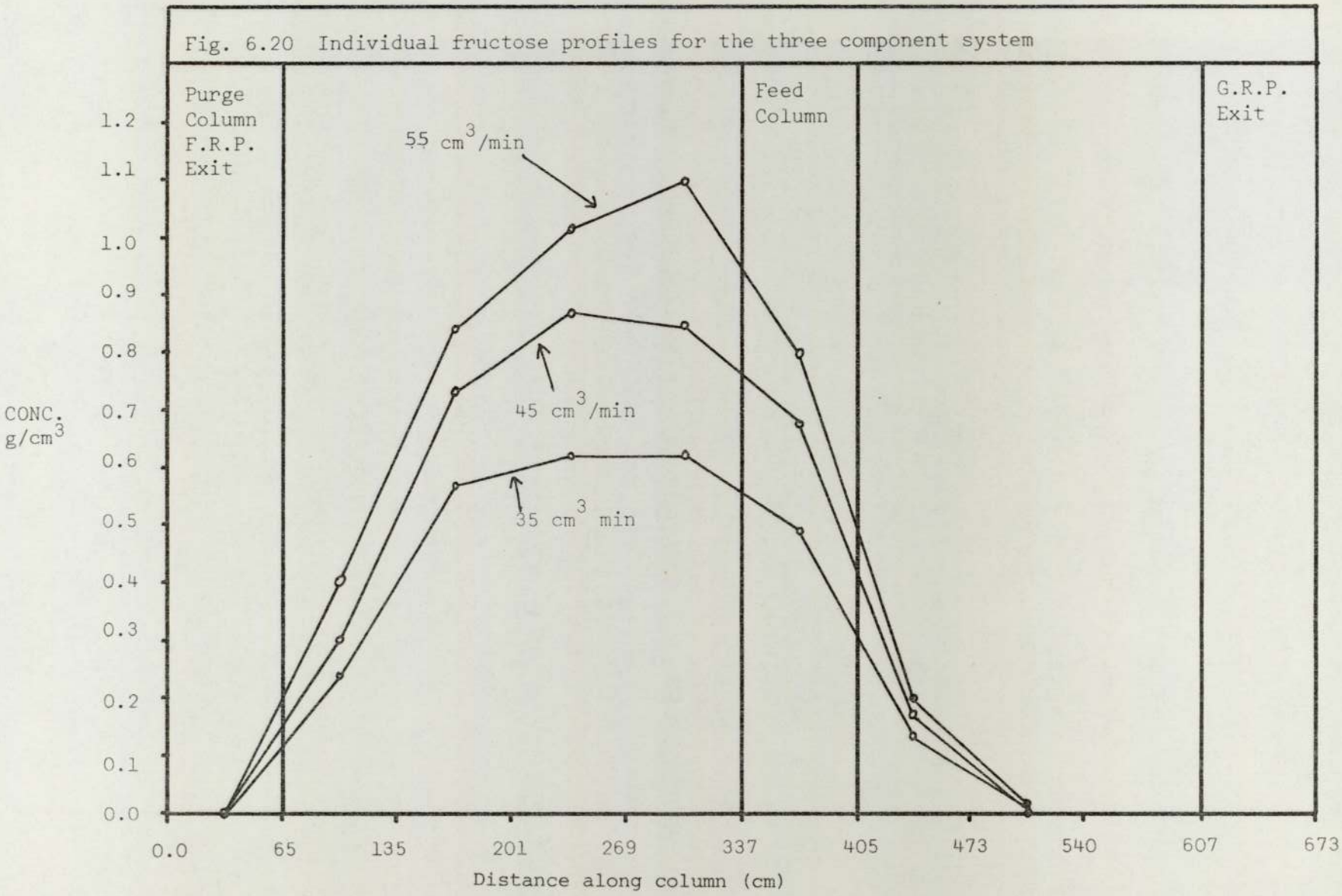


Fig. 6.20 Individual fructose profiles for the three component system

open the valves against this process pressure. Thus the run was abandoned and the columns were purged out. The concentration profiles obtained for the three runs were plotted in Figs. 6.17, 6.18 and 6.19. From the mass balance calculations carried out on the third experiment it was evident that pseudo-equilibrium had not been reached. Thus the profile obtained in Fig. 6.19 was not an equilibrium profile but certain trends and conclusions can be made.

When the fructose profiles are compared, Fig. 6.20, it is evident that the mass of fructose in a column increases with increased feed rate. This indicates that the resin is able to chemisorb the fructose quite adequately under these conditions. This is confirmed when it is seen that very little fructose travels with the mobile phase into the post feed section. A fact that becomes evident is that the purge flow rate of 500 $cm^3 min^{-1}$ is capable of removing all the retained fructose at this high concentration.

When the glucose and dextran profiles are studied it may be seen in all cases that very little dextran travels with the stationary phase into the pre feed section. This is to be expected as the species has been shown to be totally excluded from the interior of the resin particles. The small amounts of dextran that are present aredue to the semi-continuous nature of the column sequencing.

The glucose which has travelled with the stationary phase into the pre feed section is small but increases as the feed rate increases. Again it is thought to be the effect of the amount of sugar present retarding its progress through the column due to the hydration phenomena and the physical flow phenomena. No trace of glucose was detected in the fructose products.

In summary, total purity may be attained with an eluent to feed ratio of 2:1. At this high throughput pressure drop problems were encountered with the SCCR unit. These may be overcome in one of two ways:

1. Increase the available air supply pressure with the necessary safety precautions or

2. Raise the temperature of operation to lower the viscosity of the dextran and reduce the pressure drop. Batch experimental evidence has shown (6) that as temperature increases a decrease in resolution between fructose and glucose occurs. Research is necessary to assess this on a continuous basis and when the pressure drop problems are overcome change of switch period and consequently higher flow rates may be attempted to increase the throughput above 2.3 $kghr^{-1}$.

6.6 Radial distribution

All the columns had samples points fitted. On some columns there were groups of sample points at 90^{o} to each other on the same radial plane. These four points were placed at three different positions along the columns' length. To check radial distribution, during the commissioning experimental run the rig was stopped and various samples both along the bed and into the bed were taken and subsequently analysed. This was accomplished by using hypodermic needles of various lengths for insertion to different depths. The results obtained are shown in Table 6.6.

6.6.1 Results and discussion

No unacceptable maldistribution was found to occur.

DISTANCE INTO BED cm	DISTANCE FROM COLUMN INLET		
	5 cm, % w/v G	35 cm, % w/v G	60 cm, % w/v G
At Wall	9.23	9.65	9.71
1.27	9.67	9.60	9.75
3.81	9.45	9.69	9.76
At Bed Centre	9.65	9.64	9.69

TABLE 6.6 RADIAL DISTRIBUTION DATA

At the plane 5 cm into the bed, the various concentrations across the bed do vary by 4.9% but with no apparent trend. This variation grows less significant the further one samples from the column inlet. These results suggest a satisfactory packing technique and distribution system.

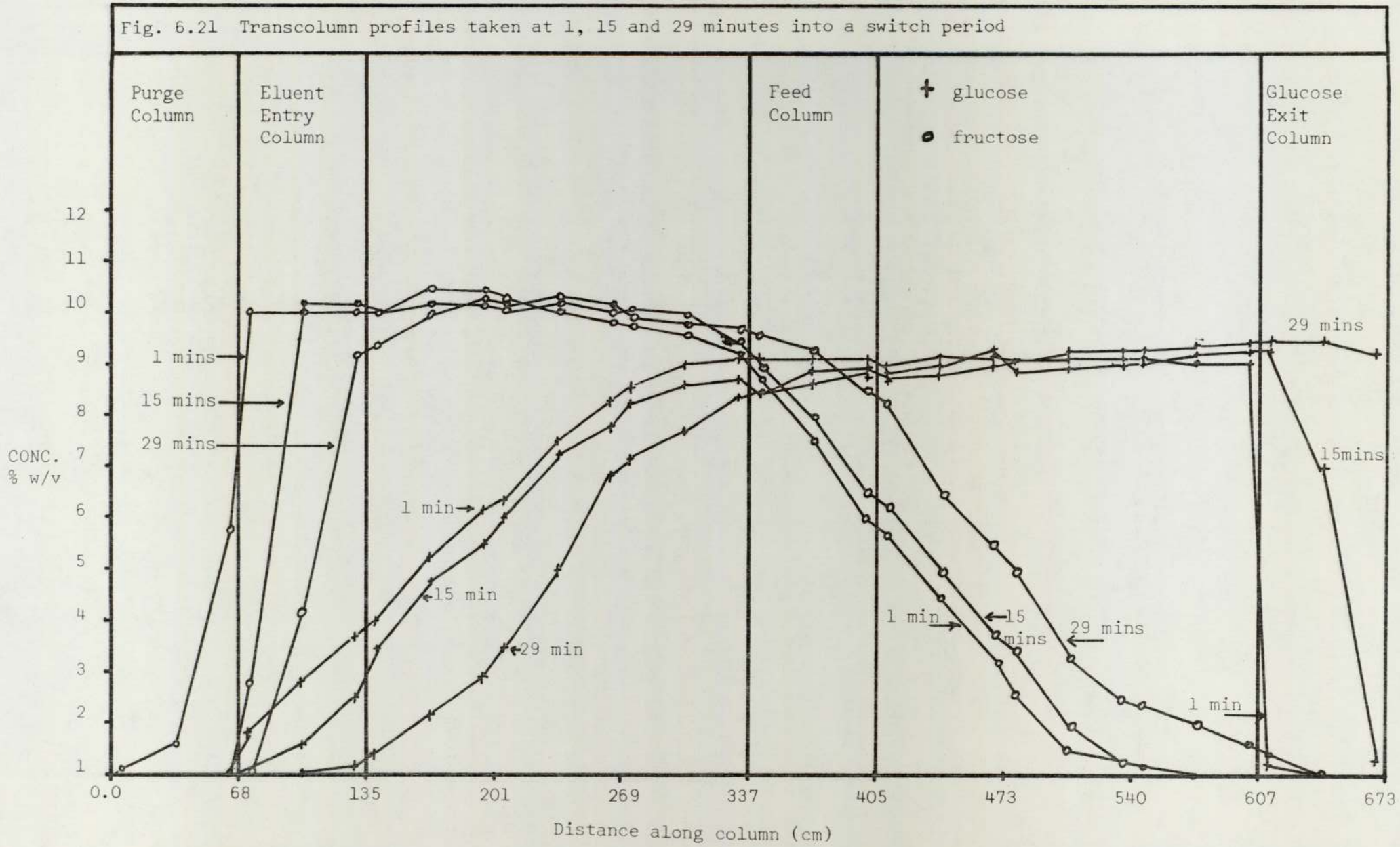
6.7 The transcolumn profile

To obtain an indication of how the concentration profile changed with time during a 30 minute switch period, use was made of the longitudinal sample points on the columns. Once the pseudo-equilibrium state had been established the pumps were stopped 1 minute into a switch period. Samples of on-column fluid were then withdrawn from 30 points from the eluent inlet column along the column length to the purge outlet section. The pumps were restarted and stopped again when fifteen minutes of the period had elapsed. Samples were again removed from the same sample points. This procedure was repeated at twenty-nine minutes into the switch period. The ninety samples were analysed and the concentration profiles constructed as shown in Fig. 6.21.

6.7.1 Results and discussion

The most noticeable effects occur at each end of the separating section in columns 10 and 8 in Fig. 6.21. In column 9, the purge section it is seen that after 15 minutes of purging there is no trace of fructose in the column - thus it is ready for the next sequence.

Fig. 6.21 Transcolumn profiles taken at 1, 15 and 29 minutes into a switch period



In column 10, the eluent entry column there is initially a high concentration of fructose similar to that in column 1. There is also a significant quantity of glucose. With the eluent entering both sugars are being eluted through the bed. After 29 minutes almost all of the glucose has been removed which leaves almost pure fructose to be purged off as the fructose product at the next sequence.

When both the glucose and fructose profiles are observed it can be seen that as the switch period progresses they both move with the mobile phase in a similar way to a slug of feed moving through a batch column.

In column 8 the glucose exit column is initially very low in sugar concentration. This is because in the previous switch period it was the purged column and the void volume is full of pure water as all the fructose had been removed. As this water is forced out the on column concentration of the glucose increases. At the end of the switch period it is constant throughout the column length.

6.7.2 Increased product concentrations

The results of the transcolumn profile, Fig. 6.21, confirmed a belief by the author that product concentrations could be significantly improved by collection of a fraction of each product stream. This was supported by a concentration profile of each product stream as a thirty minute switch period elapsed. These are shown in Figs. 6.22 and 6.23 and in Table 6.7. During a consequent switch period each product stream was collected in two fractions of fifteen minutes each. The glucose product rose in concentration from 3% solids to 6% solids and the fructose product from 0.87%

Fig. 6.22 Glucose Product Concentration Profile

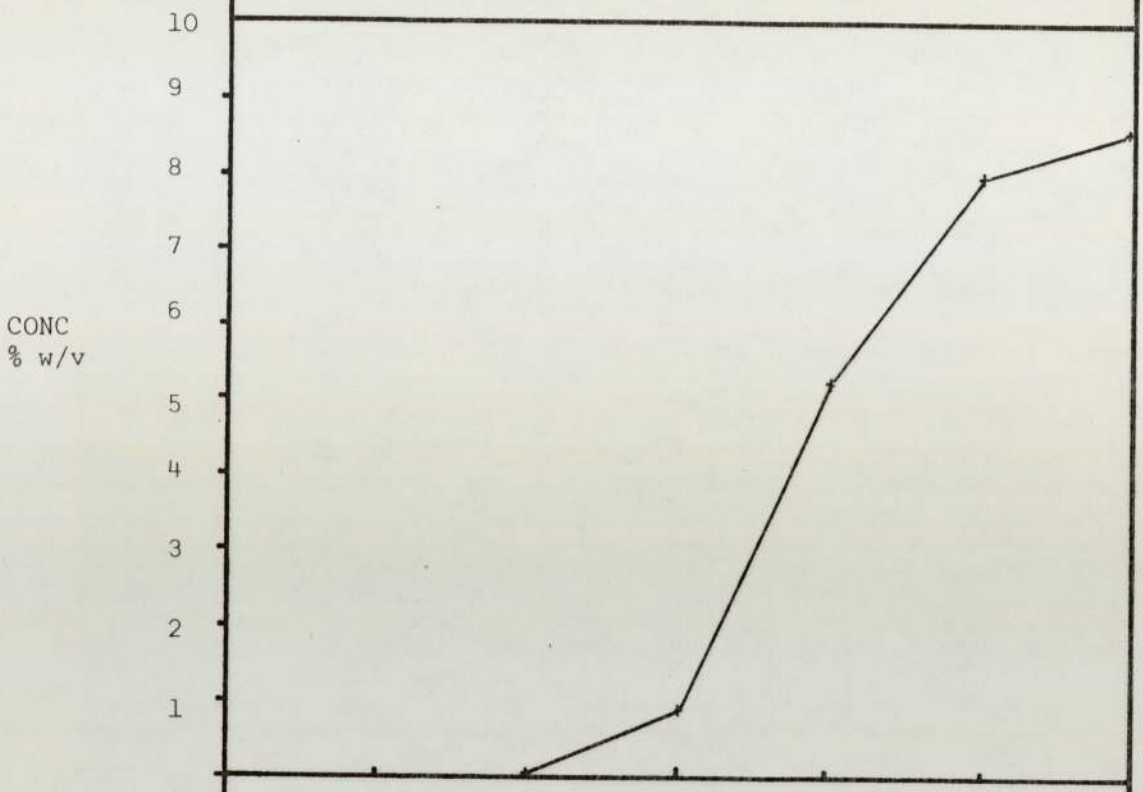
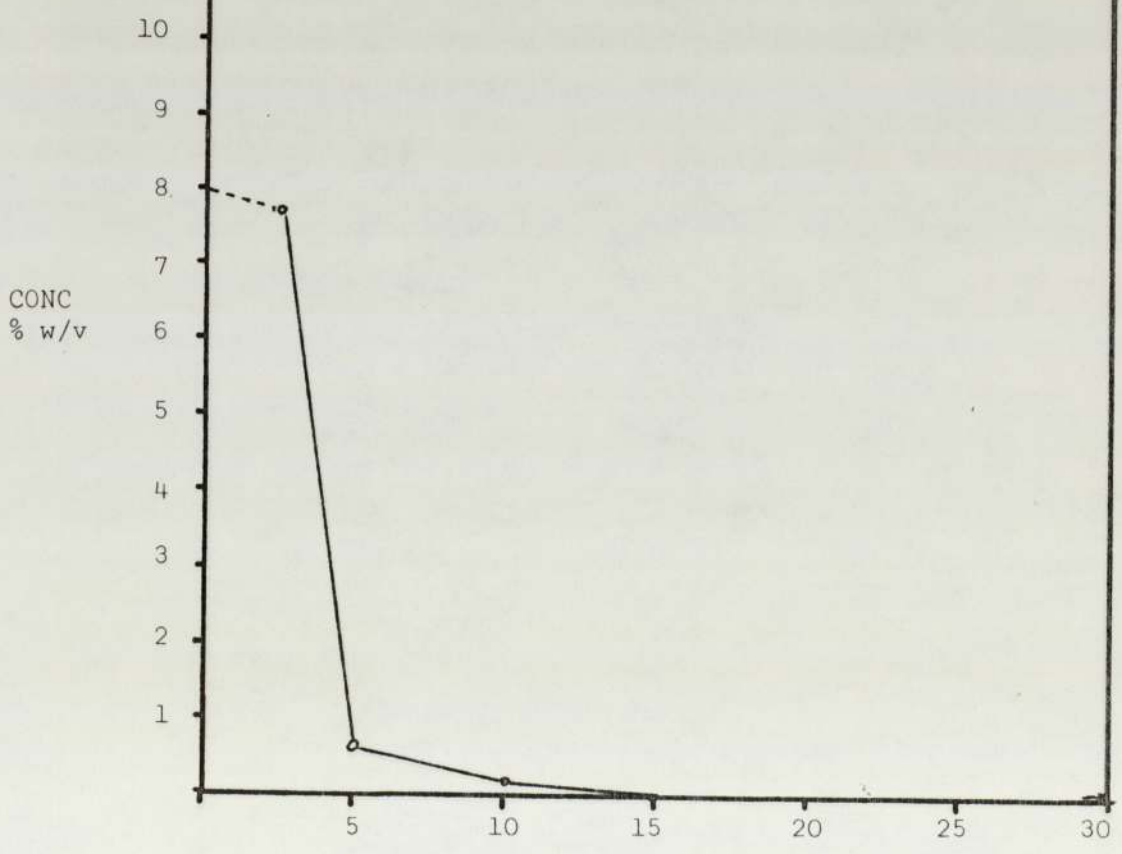


Fig. 6.23 Fructose Product Concentration Profile



TIME min	Glucose Outlet Concns. % w/v	Fructose Outlet Concns. % w/v
2.5	-	7.8
5	-	0.6
10	-	0.25
15	0.9	-
20	5.2	-
25	8.0	-
30	8.6	-

TABLE 6.7 - Glucose and Fructose Outlet Profiles

solids to 2.1% solids. If the process were used on a commercial basis, higher concentrations would be possible if some loss of product or recycle were tolerable.

6.8 Comparison of batch and continuous operation

Since the conclusion of this research the SCCR has been adapted by Miss S. Thawait (88) to operate in the batch mode. This has been achieved by removing all transfer valves and maintaining the controlling camshaft in the same position. This allows a slug of feed to be pumped on to the column and then eluted through the ten linked sections. The product issuing from the end of the column was monitored and concentration profiles constructed. Using this information repetitive batch experiments have been performed.

6.8.1 Results and discussion

Within the column length available total separation was not achieved. A recycled fraction was always necessary to obtain two pure products. The eluent flow rate was fixed at 120 $cm^3 min^{-1}$. The volume of feed applied per batch was either 2% or 4% of the total column volume.

A direct comparison between the batch and continuous mode of operation in terms of sugar throughput was made at various feed concentrations, see Table 6.8.

The throughputs quoted take no account of recycle which was used on the batch experiments although the concentrations are those of the pure products.

Feed Conc %w/v		Batch Mode			Continuous Mode		
F %w/v	G %/wv	Glucose Conc. % w/v	Fructose Conc. % w/v	T'put Kg/hr	Glucose Conc. % w/v	Fructose Conc. % w/v	T'put Kg/hr
10	10	1.74	1.13	0.114	2.37	0.64	0.63
20	20	4.40	2.88	0.229	3.60	1.29	0.84
30	30	6.74	2.30	0.310	5.60	1.79	1.26

TABLE 6.8 - Comparison of Batch and Continuous Mode of Operation

It will be seen that the product concentrations of the batch experiments are high than those of the continuous experiments. If the flow scheme suggested in section 6.7.2 were adopted it is believed that the product concentrations obtained in the continuous experiments could be approximately doubled. This would compare with if not exceed those of the batch work. These experiments confirm the belief that the throughput of the continuous system is about four times greater than that of an equivalent batch system.

CHAPTER 7

MATHEMATICAL MODELLING AND COMPUTER SIMULATION OF THE SCCR

7.0 Introduction

Attempts to simulate truly continuous counter-current mass transfer operations such as distillation have proved successful. Many of these models employ the equilibrium plate concept and use mass balance and equilibrium relationships to predict product compositions and the number of stages necessary for a particular degree of separation.

For truly continuous chromatography of the moving bed type, a probabilistic model was proposed by Sciance and Crosser (89) to relate the degree of separation and operating conditions to column length for a binary system. For the case when the feed is introduced into the midpoint the following equations were proposed:

$$\ell n\ (u_Z)_A = \frac{\ell\ k''_A}{2u}\ (K_A - \psi) \qquad 7.1$$

$$\ell n\ (1 - (u_Z)_B) = \frac{-\ell\ k''_B}{2u}\ (K_B - \psi) \qquad 7.2$$

Where A = faster moving component

B = slower moving component

$(u_Z)_A$ = Bottoms/feed mass flow rate ratio of A

$(u_Z)_B$ = Tops/feed mass flow rate ratio of B

k'' = rate constant of desorption

u = average mobile phase velocity

ψ = mobile phase/stationary phase velocity ratio

ℓ = required column length

Data on k''_A and k''_B are difficult to determine experimentally and published values are rare. This has restricted the application

of the model.

To model moving column gas-liquid chromatography, Al-Madfai (79) used the random walk approach which Giddings (15) developed for a batch operation. He obtained an expression for predicting plate height as follows:

$$H = d_p \frac{2 D_m}{u} + \frac{2 \gamma_1 \gamma_2}{(\gamma_1+\gamma_2)^2} \frac{(u + u_L)^2}{u\gamma_2 - u_L\gamma_1} \qquad 7.3$$

Where H = plate height

d_p = average particle diameter

D_m = mobile phase diffusivity

u = mobile phase velocity

γ_1 = rate of transfer of molecules from gas to liquid

γ_2 = rate of transfer of molecules from liquid to gas

u_L = stationary phase velocity

Al-Madfai also related the number of theoretical plates required to resolve binary mixtures in a batch column and by a continuous chromatographic system.

$$\frac{N_{cc}}{N} = 3(\alpha-1) \qquad 7.4$$

where N_{cc} = Number of 'continuous' plates

N = Number of 'batch' plates

α = separation factor of components.

A similar result was obtained by Rony (90) in his attempt to relate the two types of chromatography.

A model developed by Barker and Lloyd (69) to simulate a continuous gas-liquid chromatographic system used the transfer unit concept. They developed equations to predict the number of overall transfer units in the moving column chromatograph for the rectifying and stripping section of the separating length:

$$(N_G)_R = \frac{1}{V_G/(K_o V_L - 1)} \ln \left[\frac{E_1/K_o V_L - C_i(V_G/K_o V_L - 1)}{E_1/K_o V_L - C_{ii}(V_G/K_o V_L - 1)} \right] \quad 7.5$$

$$(N_G)_S = \frac{1}{(1-V_G/K_o V_L)} \ln \left[\frac{E_2/K_o V_L - C_i\,(1-V_G/K_o V_L)}{E_2/K_o V_L - C_{ii}(1-V_G/K_o V_L)} \right] \quad 7.6$$

where

$(NG)_R$, $(N_G)_S$ = the number of overall gas phase transfer units in the rectifying and stripping sections respectively

E_1, E_2 = the mass flow rate of solute leaving in products 1 and 2 respectively

C_i, C_{ii} = gas phase solute concentrations at points i and ii in the column

V_G, V_L = the gas and liquid volumetric flow rates

K_o = the partition coefficient

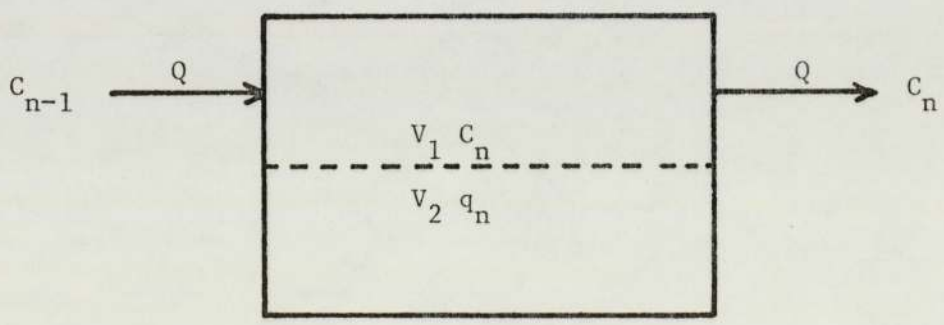
7.1 The approach employed for the SCCR simulation

In most systems which have been successfully simulated, a truly continuous counter-current mode of operation was used.

To simulate the semi-continuous system an additional variable: time, must be introduced to accommodate the sequential switch periods.

In his work Ching (6) developed a model which attempted to describe the SCCR mode of operation. The concept used was that of the equilibrium stage or plate. The separating length is considered to consist of a number of theoretical plates. Each plate comprises a volume of mobile phase and a volume of stationary phase. If a mass balance is calculated over a plate a differential equation may be derived which, upon integration, predicts the concentration of the mobile phase leaving that plate. This calculation is repeated for every theoretical plate over a small time increment. When the number of time increments simulated is equal to the length of the switch period, the sequencing action of the stationary phase is imposed by stepping the concentration profile obtained backwards by one column and continuing the simulation. The model considers only one solute and the profile of the second solute is determined by duplicating the calculation at each plate with a different variable name. The assumption is made that there is no interaction between the two components.

7.2 The model



Theoretical plate n.

A mass balance over the theoretical plate n gives:

$$QC_{n-1} = QC_n + V_1 \frac{dC_n}{dt} + V_2 \frac{dq_n}{dt} \qquad 7.7$$

Where

Q = mobile phase volumetric flow rate

C = solute concentration in mobile phase

q = solute concentration in stationary phase

V_1 = volume of mobile phase in a plate

V_2 = volume of stationary phase in a plate

The equilibrium distribution coefficient is defined as $K_d = \frac{q_n}{C_n}$.

Substituting this relationship into eqn. 7.7

$$QC_{n-1} = QC_n + (V_1 + V_2K_d) \frac{dC_n}{dt} \qquad 7.8$$

This equation may now be integrated providing Δt is sufficiently small to allow C_{n-1} to be considered constant:

$$C_n = C_{n-1} \left(1 - \exp\left(\frac{-Q\Delta t}{V_1 + V_2K_d}\right)\right) + C_n^o \exp\left(\frac{-Q\Delta t}{V_1 + V_2K_d}\right) \qquad 7.9$$

The first term on the right hand side of the equation represents the material transfer from the previous plate n-1 and the second term represents the material already present on plate n.

For the feed plate the mass balance yields a similar equation with additional terms for the feed stream entering the plate:

$$C_n = \frac{QC_{n-1} + FC_f}{Q + F} \left(1 - \exp\left(\frac{-Q\Delta t}{V_1 + V_2 K_d}\right)\right)$$

$$+ C_n^o \exp\left(\frac{-Q\Delta t}{V_1 + V_2 K_d}\right) \qquad 7.10$$

Where F = volumetric flow rate of feed

C_f = feed concentration

7.3 Improvements in the model

In the existing model the assumption was made that glucose was not retained by the resin. Its equilibrium distribution coefficient, K_{dg}, was therefore equated to zero. Batch experimental work described in section 4.4 has proved that this is not the case. When the elution volume of dextran, a truly non-retained component is compared with that of glucose, a significant difference is observed. This reflects a difference in the equilibrium distribution coefficient. With dextran no distribution occurs because all the component remains in the mobile phase and travels through the column in the void volume. Glucose molecules are small enough to penetrate the intra-particle volume as well as the void volume, thus a distribution does occur between the mobile and stationary phase and consequently the equilibrium distribution coefficient K_{dg} does have a positive value. This is an important factor in the mass balance calculation.

A significant improvement was made in the way in which the model calculated the pseudo-equilibrium concentration profile. The existing model took a numerical average of the individual plate outlet concentrations, C_n, for each plate and multiplied this value by the number of theoretical plates in the column. This took no account of the component associated with the stationary phase of each

plate. The new routine that has been introduced considered each theoretical plate individually. Using relationships between the mobile phase plate volume, the stationary phase plate volume, the equilibrium concentration of the component and the equilibrium distribution coefficient the total amount of the component associated with each plate was calculated. These values were added together for as many plates as there were in the column. These data were then output to give a far more realistic end of run profile.

A further improvement was introduced after batch experimental work. From section 4.4 it was observed that the number of plates per column was different for glucose and fructose. In the previous model it was assumed that the two values were the same. To overcome this discrepancy the program was adapted to simulate each sugar individually with the appropriate number of plates.

Extra routines were introduced to the model to include a third component dextran - the non-sorbed species in the feed stock. This was accomplished using further mass balance equations in the feed and separating sections of the program.

A flowsheet for the program used in the simulation is shown in Fig. 7.1. A listing of the program and a sample of the printout is provided in Appendix 1.

7.4 Simulation of experimental runs

The model performed approximately forty million calculations per simulation. The memory core available on the I.C.L. 1904S computer at Aston University was not large enough for a program of this size. Instead the University of Manchester

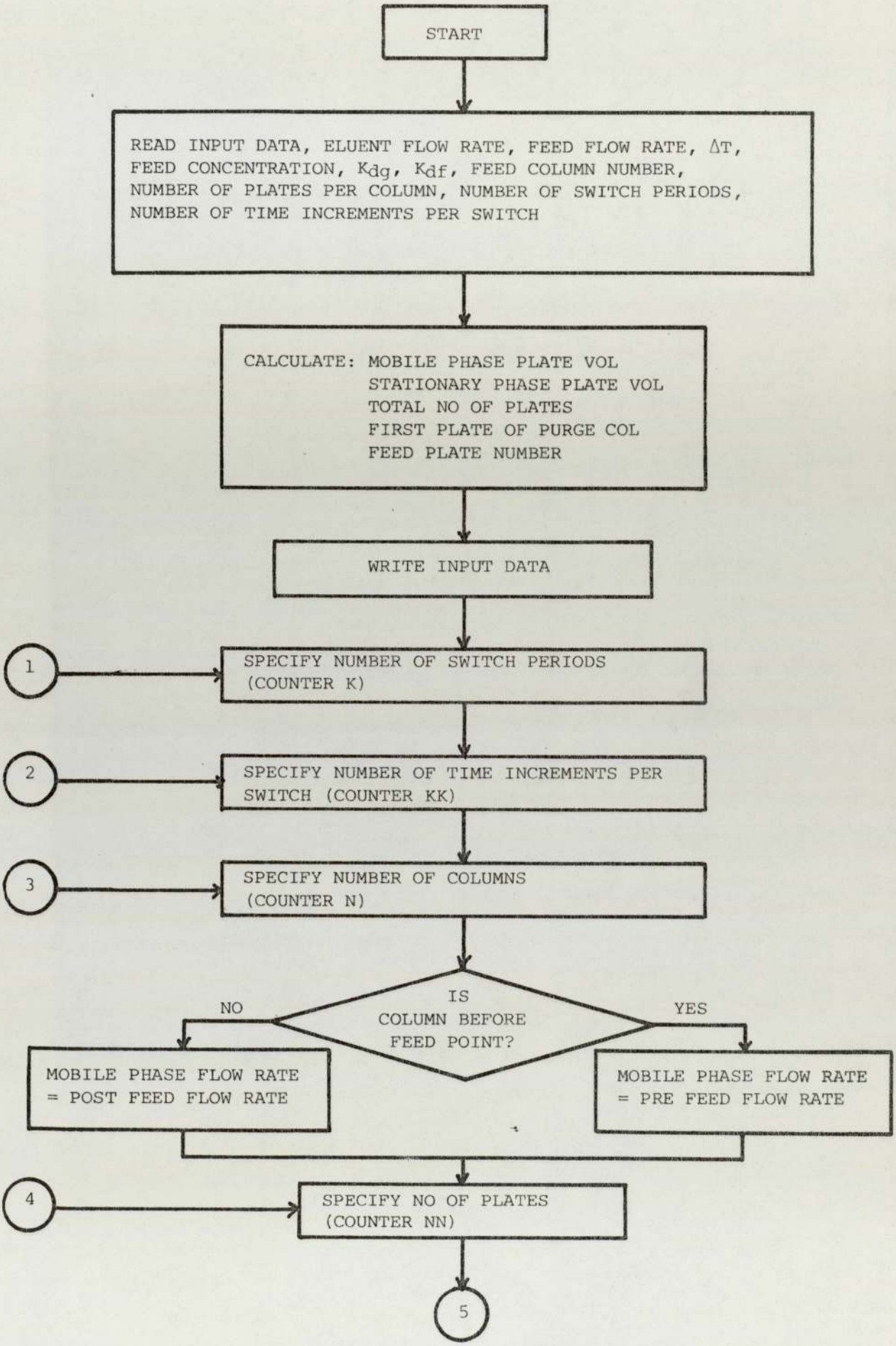
Regional Computer Centre was used. The simulations were performed on a CDC 7600 computer. Those runs in which it was suspected that pseudo-equilibrium had not been reached were not simulated.

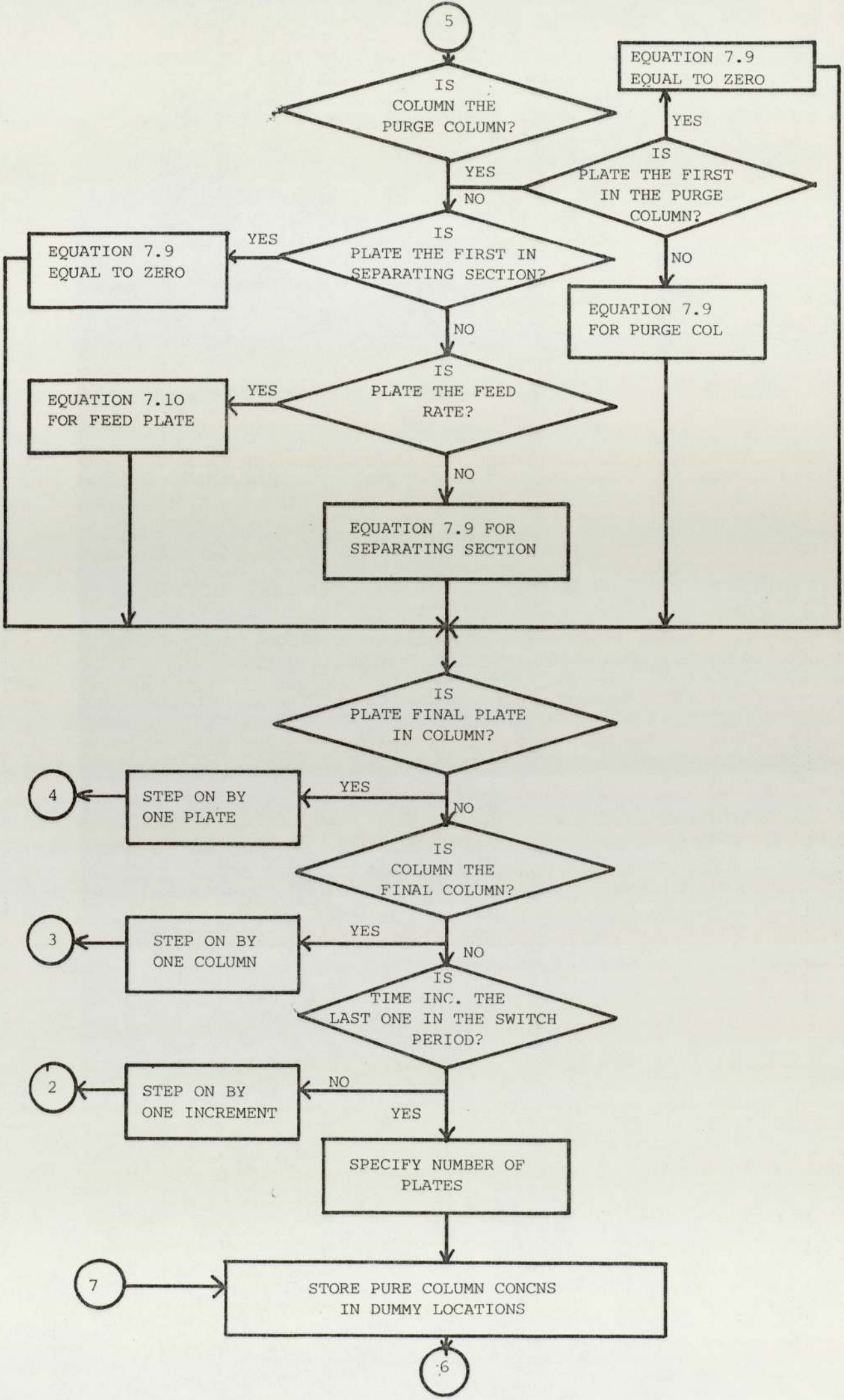
The experimental and simulated profiles of those runs simulated are shown in Figs. 7.2-7.9.

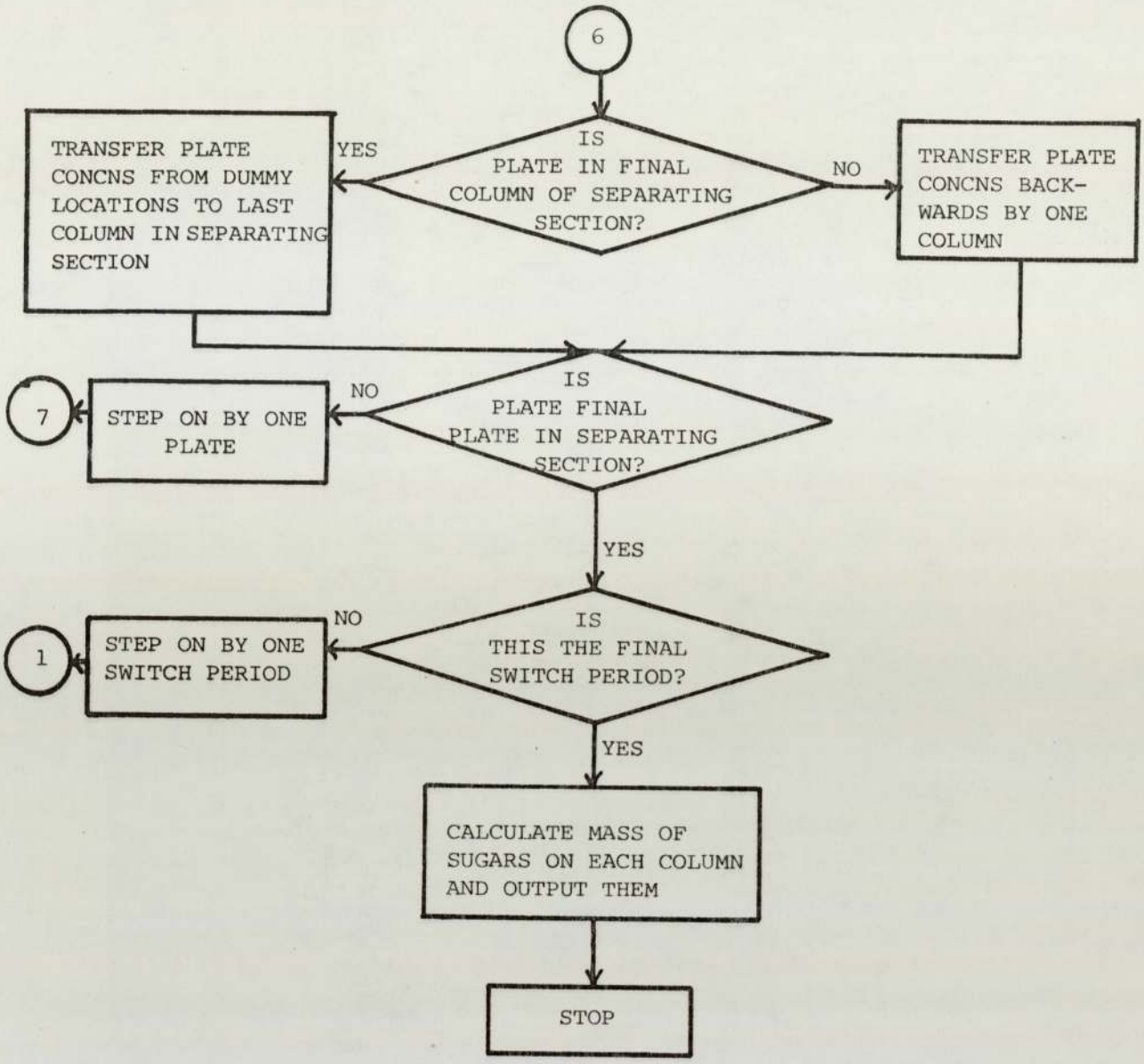
7.4.1 Results and discussion

In initial attempts at simulation of the experimental runs the values of the equilibrium distribution coefficients found during column characterisation (section 4.4) were used. The resulting profiles were very inaccurate. This was thought to be a result of the variation of the K_d's with the on-column concentration of the sugars. The values found during the characterisation were measured at virtually infinite dilution whereas in the actual operation of the SCCR on-column concentrations could reach that of the feed material. In an effort to obtain some meaningful results estimates were made of the distribution coefficients and these were input as data to the model. The profiles obtained in Figs. 7.2-7.9 are the results of the "best fit" distribution coefficients. It was observed from the results that the glucose K_d varied between 0.28 and 0.35 in the simulation compared to 0.15-0.22 at infinite dilution. The fructose K_d varied between 0.50-0.65 in the simulation and 0.37-0.52 at infinite dilution. From this it appears that the distribution of glucose changes more markedly with concentration than that of fructose. This gives support to the theory of glucose complex formation in its K_d means a higher concentration of glucose associated with the stationary phase.

FIG. 7.1 FLOW CHART FOR COMPUTER SIMULATION OF THE SCCR




5
IS COLUMN THE PURGE COLUMN?
YES
NO
IS PLATE THE FIRST IN THE PURGE COLUMN?
YES
NO
EQUATION 7.9 EQUAL TO ZERO
EQUATION 7.9 FOR PURGE COL
IS PLATE THE FIRST IN SEPARATING SECTION?
YES
NO
EQUATION 7.9 EQUAL TO ZERO
IS PLATE THE FEED RATE?
YES
NO
EQUATION 7.10 FOR FEED PLATE
EQUATION 7.9 FOR SEPARATING SECTION
IS PLATE FINAL PLATE IN COLUMN?
YES
NO
4
STEP ON BY ONE PLATE
IS COLUMN THE FINAL COLUMN?
YES
NO
3
STEP ON BY ONE COLUMN
IS TIME INC. THE LAST ONE IN THE SWITCH PERIOD?
NO
YES
2
STEP ON BY ONE INCREMENT
SPECIFY NUMBER OF PLATES
7
STORE PURE COLUMN CONCNS IN DUMMY LOCATIONS
6


6
IS PLATE IN FINAL COLUMN OF SEPARATING SECTION?
YES
NO
TRANSFER PLATE CONCNS FROM DUMMY LOCATIONS TO LAST COLUMN IN SEPARATING SECTION
TRANSFER PLATE CONCNS BACK-WARDS BY ONE COLUMN
IS PLATE FINAL PLATE IN SEPARATING SECTION?
NO
7
STEP ON BY ONE PLATE
YES
IS THIS THE FINAL SWITCH PERIOD?
NO
1
STEP ON BY ONE SWITCH PERIOD
YES
CALCULATE MASS OF SUGARS ON EACH COLUMN AND OUTPUT THEM
STOP

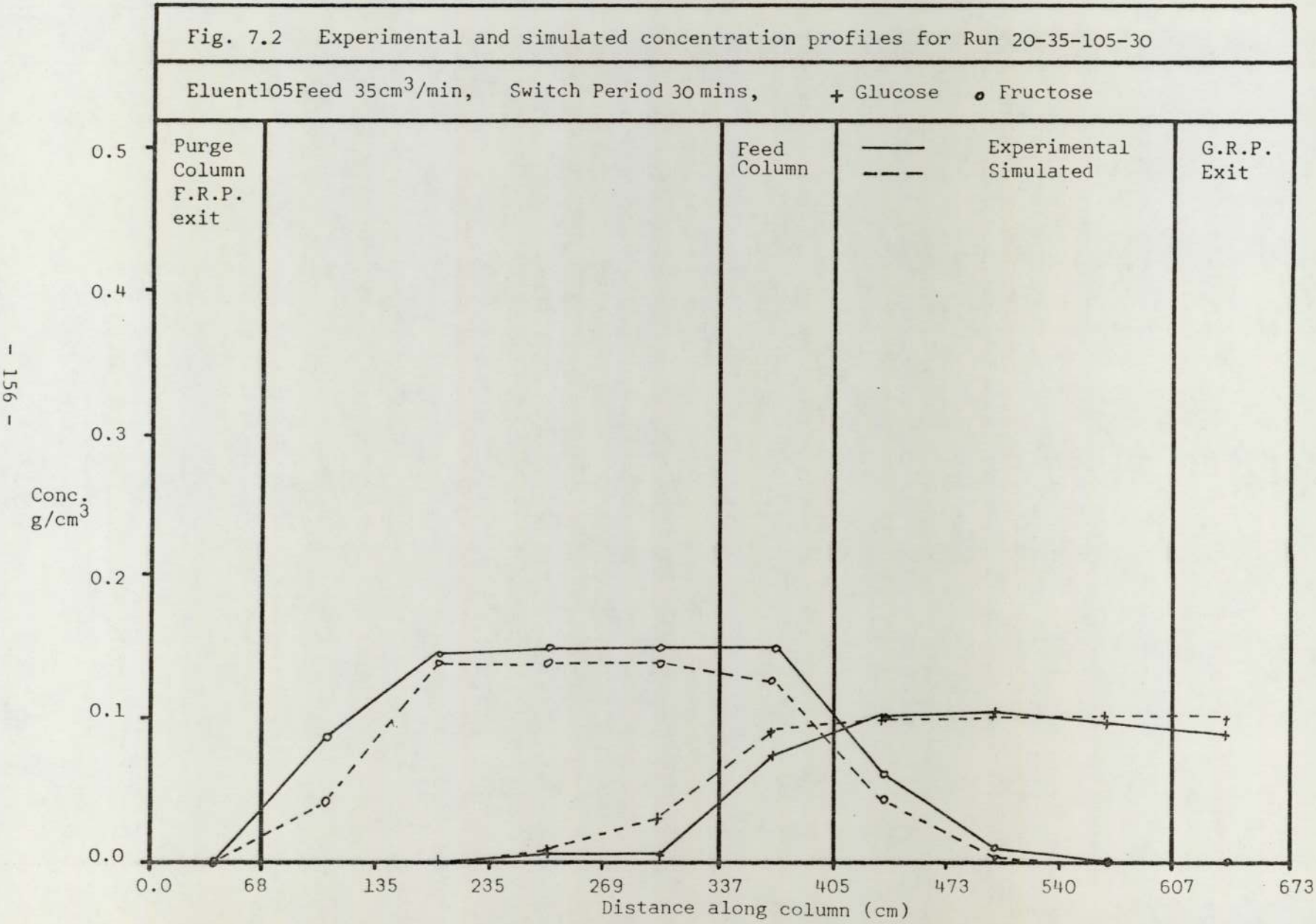
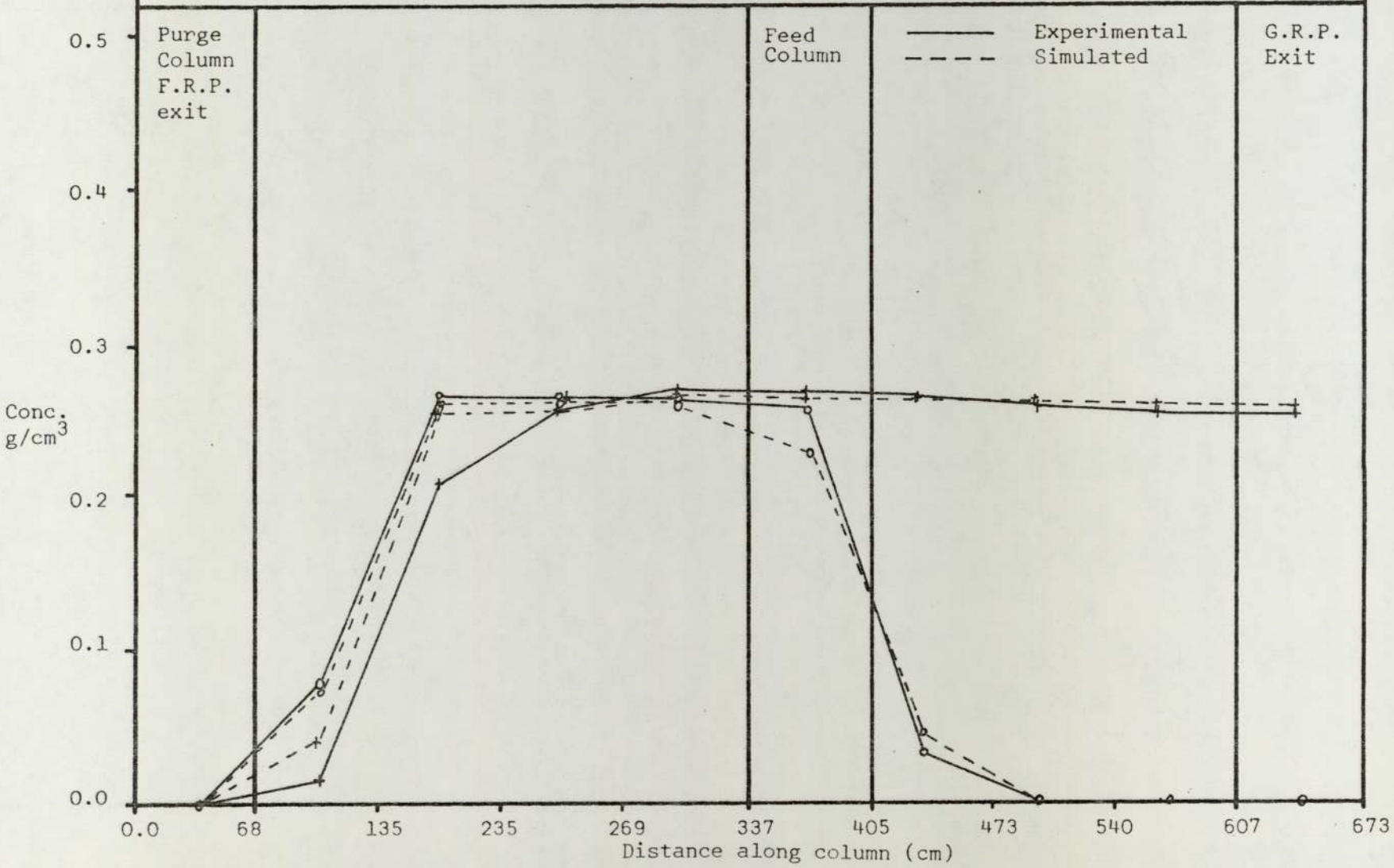


Fig. 7.2 Experimental and simulated concentration profiles for Run 20-35-105-30
Eluent105Feed 35cm3/min, Switch Period 30 mins, + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
Experimental
Simulated
G.R.P. Exit
Conc. g/cm3
0.0
0.1
0.2
0.3
0.4
0.5
0.0
68
135
235
269
337
405
473
540
607
673
Distance along column (cm)

Fig. 7.3 Experimental and simulated concentration profiles for Run 40-35-105-30

Eluent105 Feed 35 cm^3/min, Switch Period 30 mins, + Glucose o Fructose



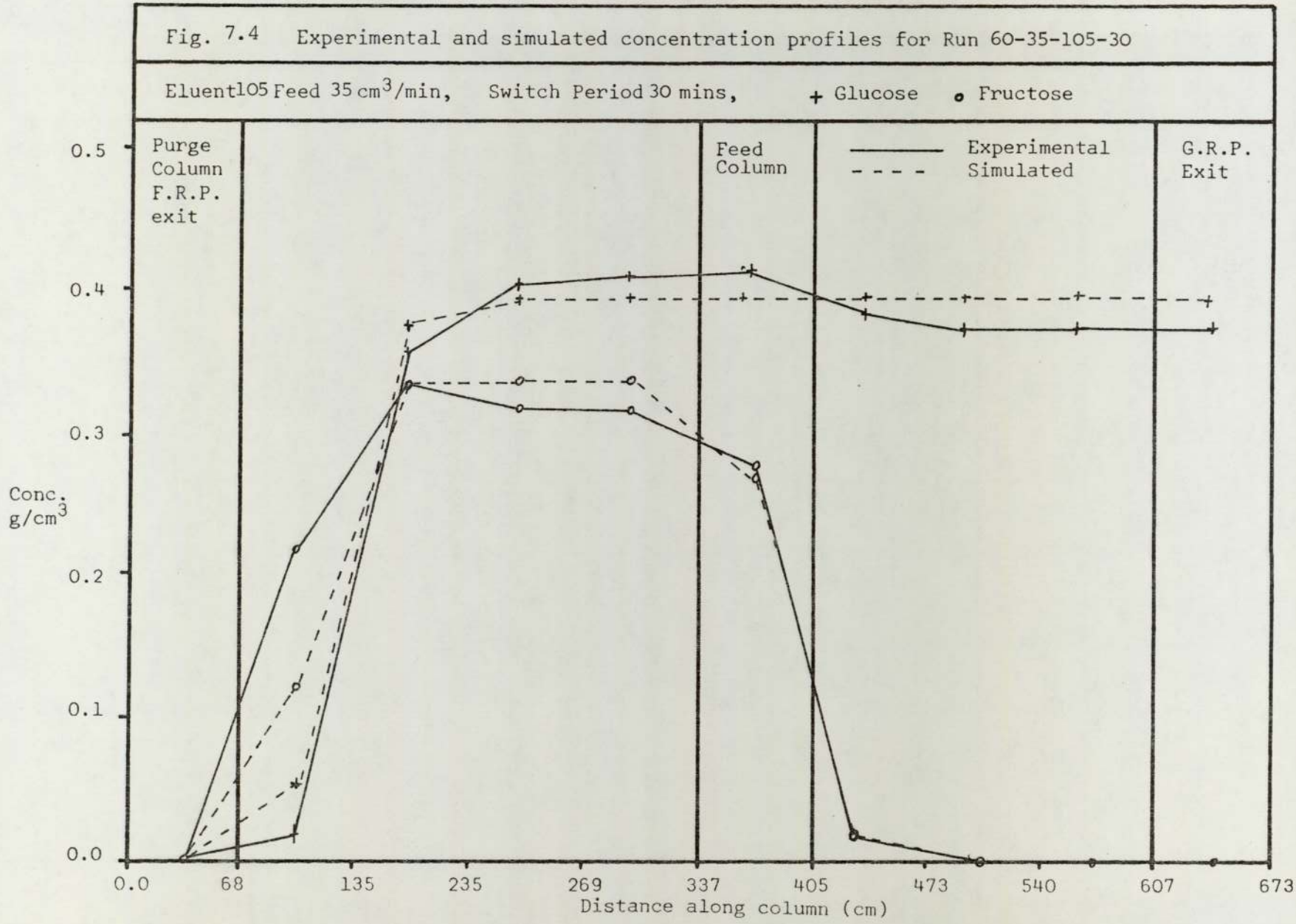


Fig. 7.4 Experimental and simulated concentration profiles for Run 60-35-105-30
Eluent105 Feed 35 cm3/min, Switch Period 30 mins, + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
Experimental
Simulated
G.R.P. Exit
Conc. g/cm3
0.0
0.1
0.2
0.3
0.4
0.5
0.0
68
135
235
269
337
405
473
540
607
673
Distance along column (cm)

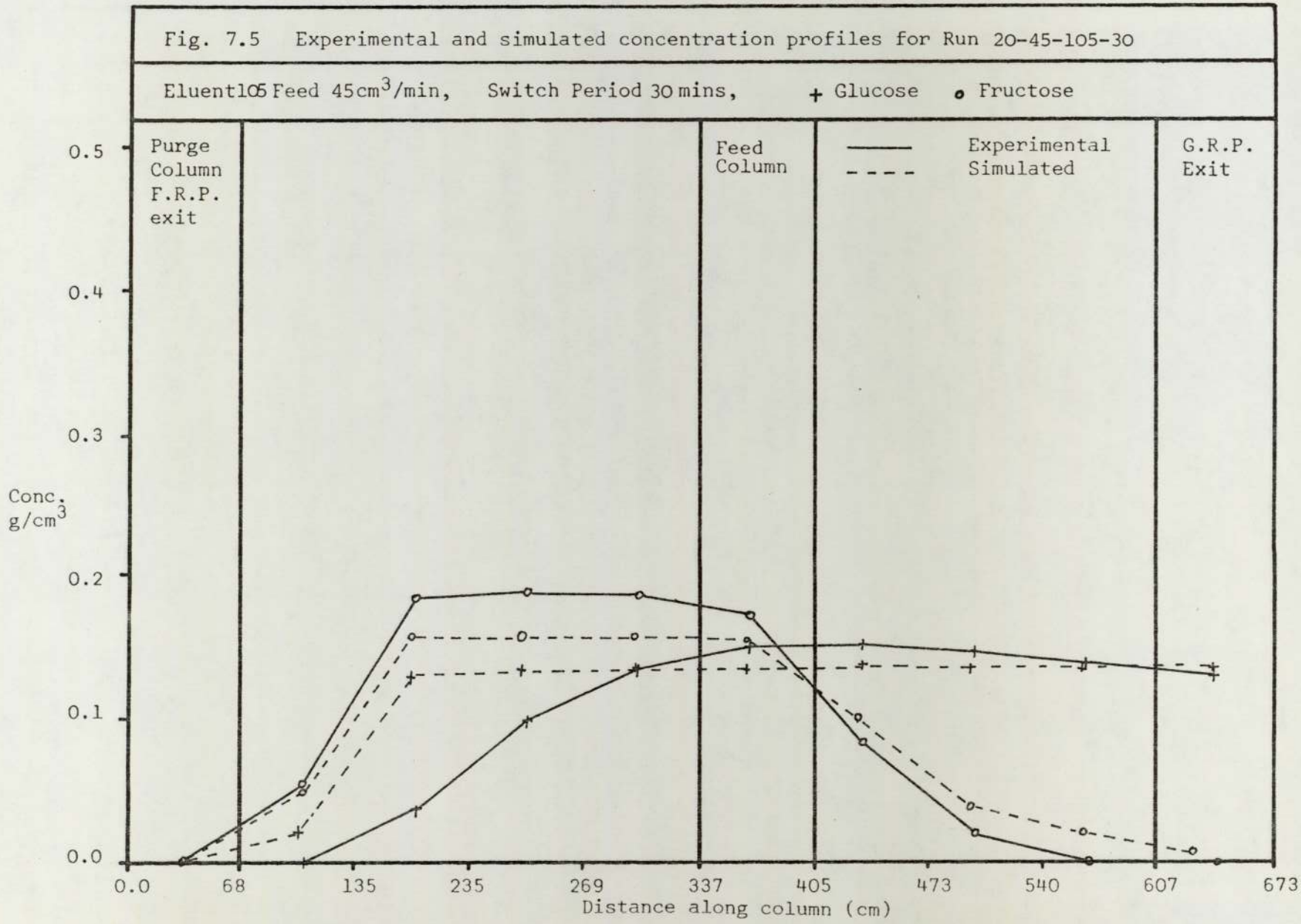


Fig. 7.5 Experimental and simulated concentration profiles for Run 20-45-105-30
Eluent105 Feed 45cm3/min, Switch Period 30 mins, + Glucose o Fructose
Purge Column F.R.P. exit
Feed Column
Experimental
Simulated
G.R.P. Exit
Conc. g/cm3
0.0
0.1
0.2
0.3
0.4
0.5
0.0
68
135
235
269
337
405
473
540
607
673
Distance along column (cm)

Fig. 7.6 Experimental and simulated concentration profiles for Run 20-55-105-30

Eluent105Feed 55cm^3/min, Switch Period 30 mins, + Glucose o Fructose

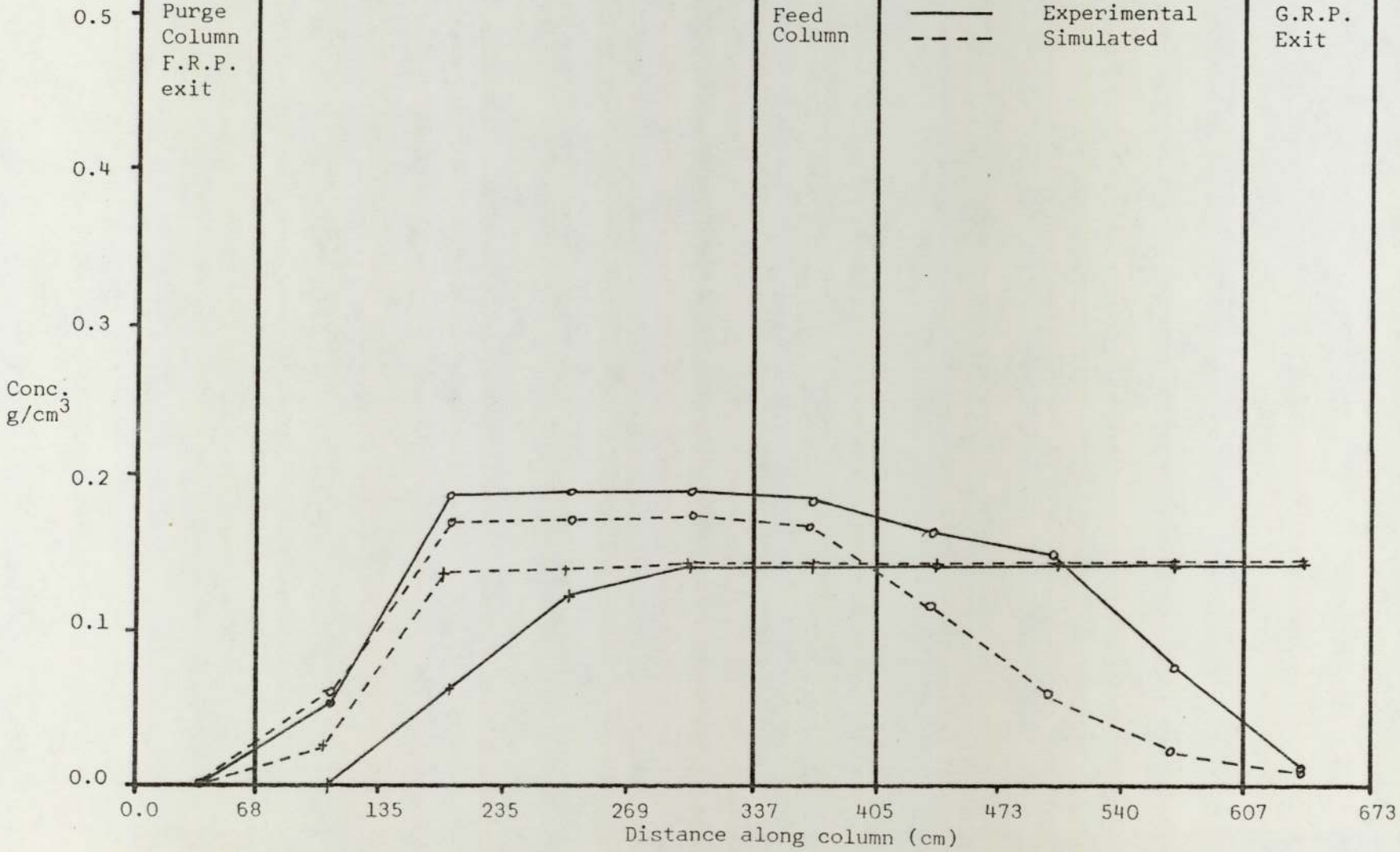


Fig. 7.7 Experimental and simulated concentration profiles for Run 20-54-164-20

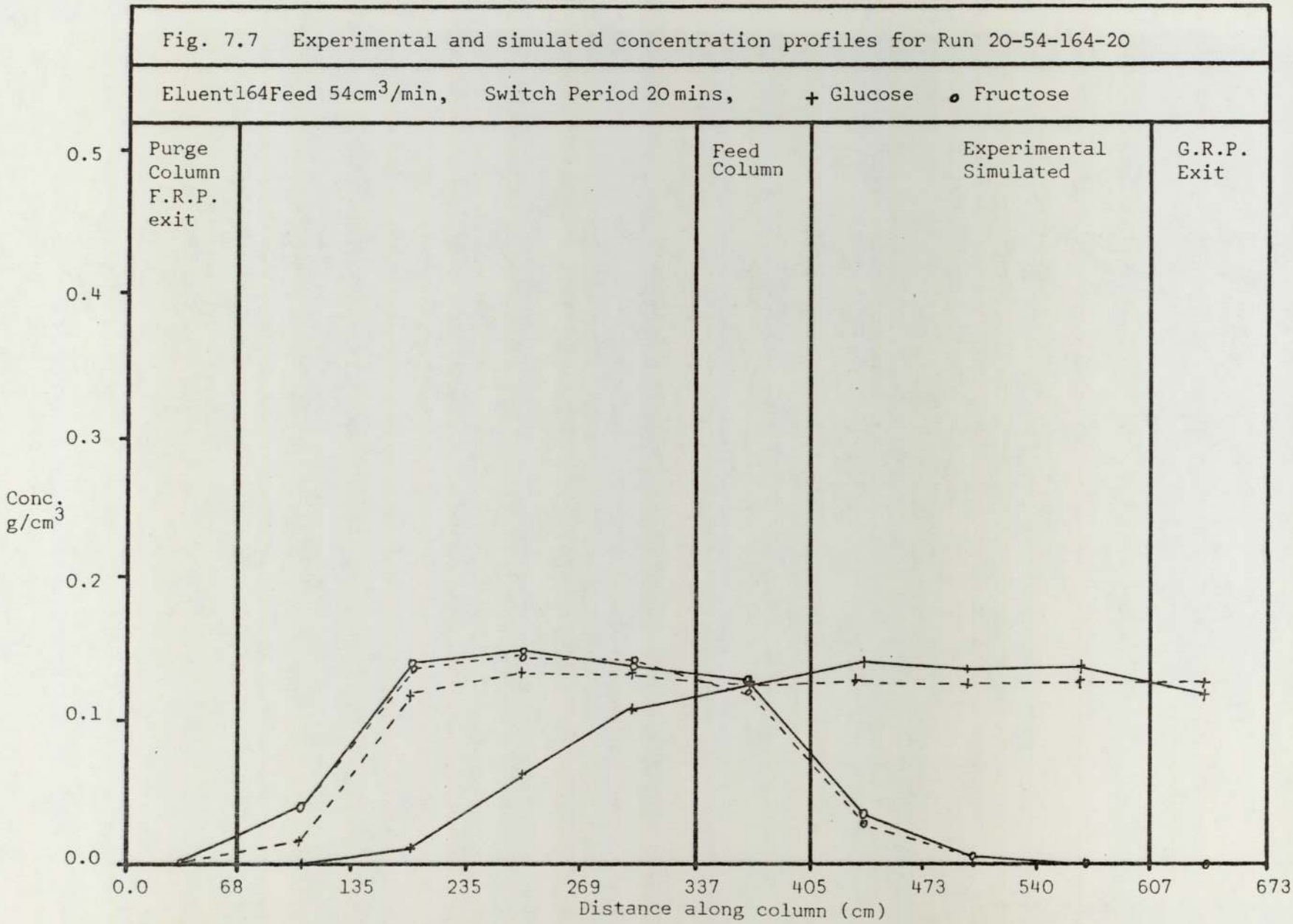


Fig. 7.8 Experimental and Simulated Concentration Profile for Run D70-35-105-30

Eluent 105 Feed 35 cm^3/min. Switch Period 30 mins. + Glucose o Fructose ▲ Dextran

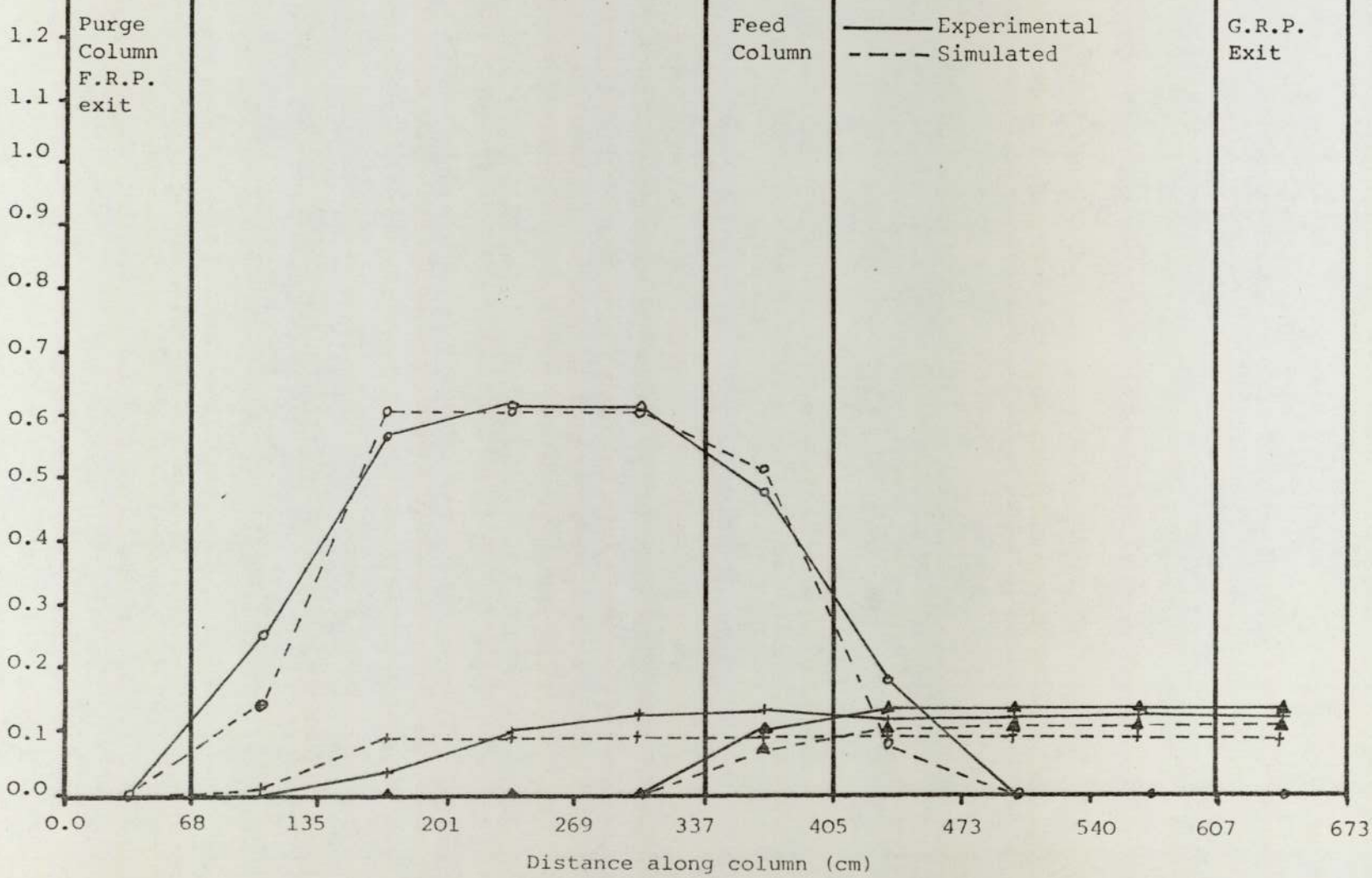
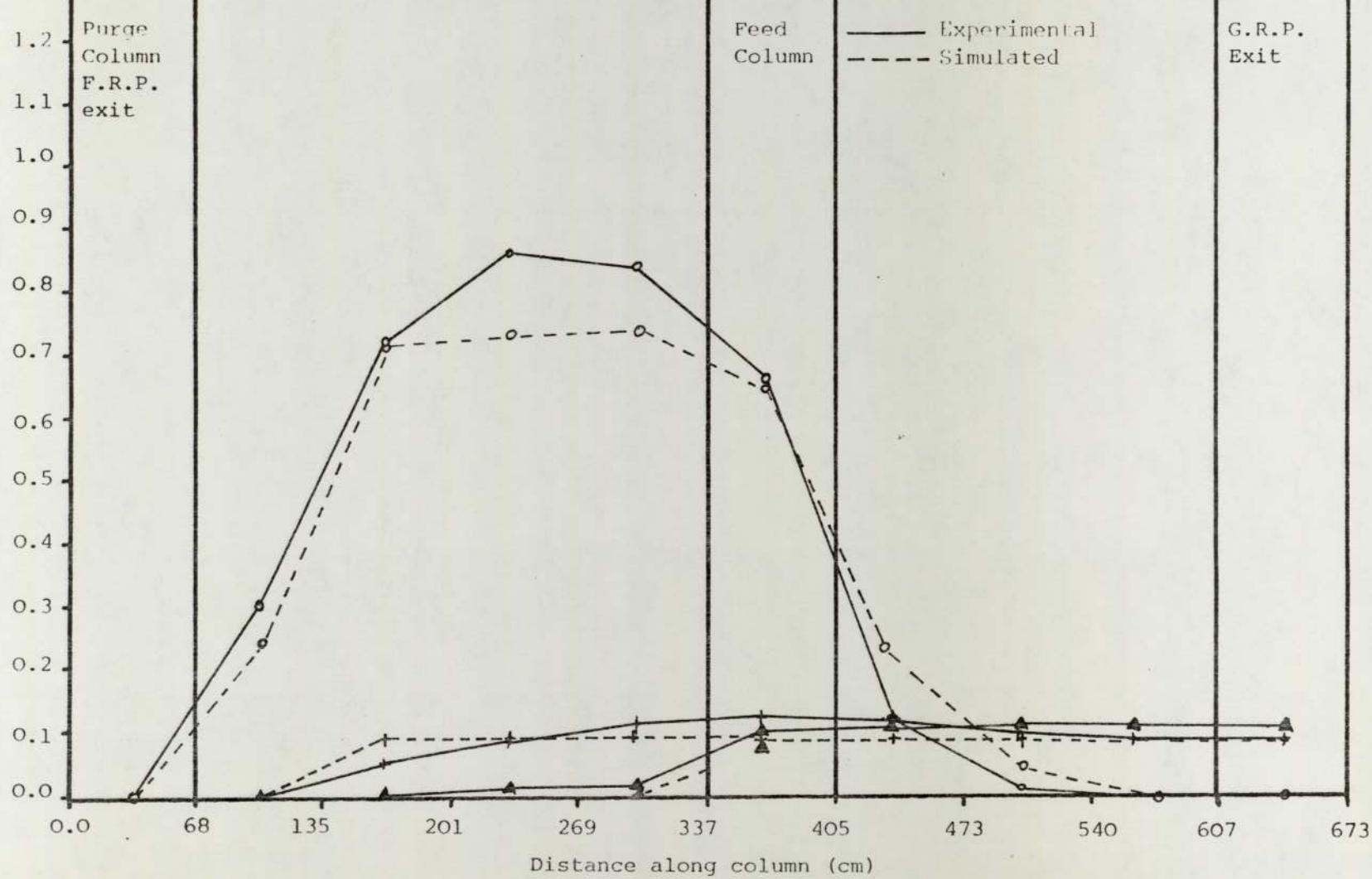


Fig. 7.9 Experimental and Simulated Concentration Profile for Run D70-45-105-30

Eluent 105 Feed 45 cm^3/min. Switch Period 30 mins. + Glucose o Fructose ▲ Dextran



If the simulated profiles are compared with the experimental profiles in Figs. 7.2-7.9 the area where the largest differences generally occur is near the eluent entry column. It is here where the on-column concentration is at its lowest and also where the greatest variation through a switch period occurs (see trans-column profile Fig. 6.21). Thus the greatest change of distribution coefficient occurs in this column. This highlights the inaccuracy of using only one value of the equilibrium distribution coefficient for the whole column length.

When a relationship is available concerning the K_d variation with on-column concentration, the model should be further improved by including a subroutine for calculating a K_d value for each component in each theoretical plate throughout the length of the separating section.

It is also open to discussion whether the instantaneous equilibrium, which the model assumes across the plate, actually occurs. Plate volumes used in the simulations are, for glucose, approximately 200 cm^3 which theoretically comprises 90 cm^3 mobile phase and 110 cm^3 stationary phase.

From the results of the simulations it was found that the model reached pseudo-equilibrium in five cycles whereas in practice approximately seven cycles are necessary which again serves to highlight a difference in the theoretical model and the real case.

As suggested in section 8.2 new data may enable the formulation of a model based on the height of a transfer unit concept.

7.5 The effect of feed-point location

To obtain an indication of how changing the location of the feed-point would affect product purity the model was used to simulate a change in this parameter.

During the experimental research programme the feed was always introduced into the fifth column after the eluent entry column. This meant that there were four column lengths in the pre-feed section and five column lengths in the post-feed section. The purge column was always isolated.

As a result of the glucose moving with the stationary phase at high concentrations and contaminating the fructose rich product, it was thought that a feed entry point nearer to the glucose exit column would result in two pure products under a wider range of conditions than those previously used.

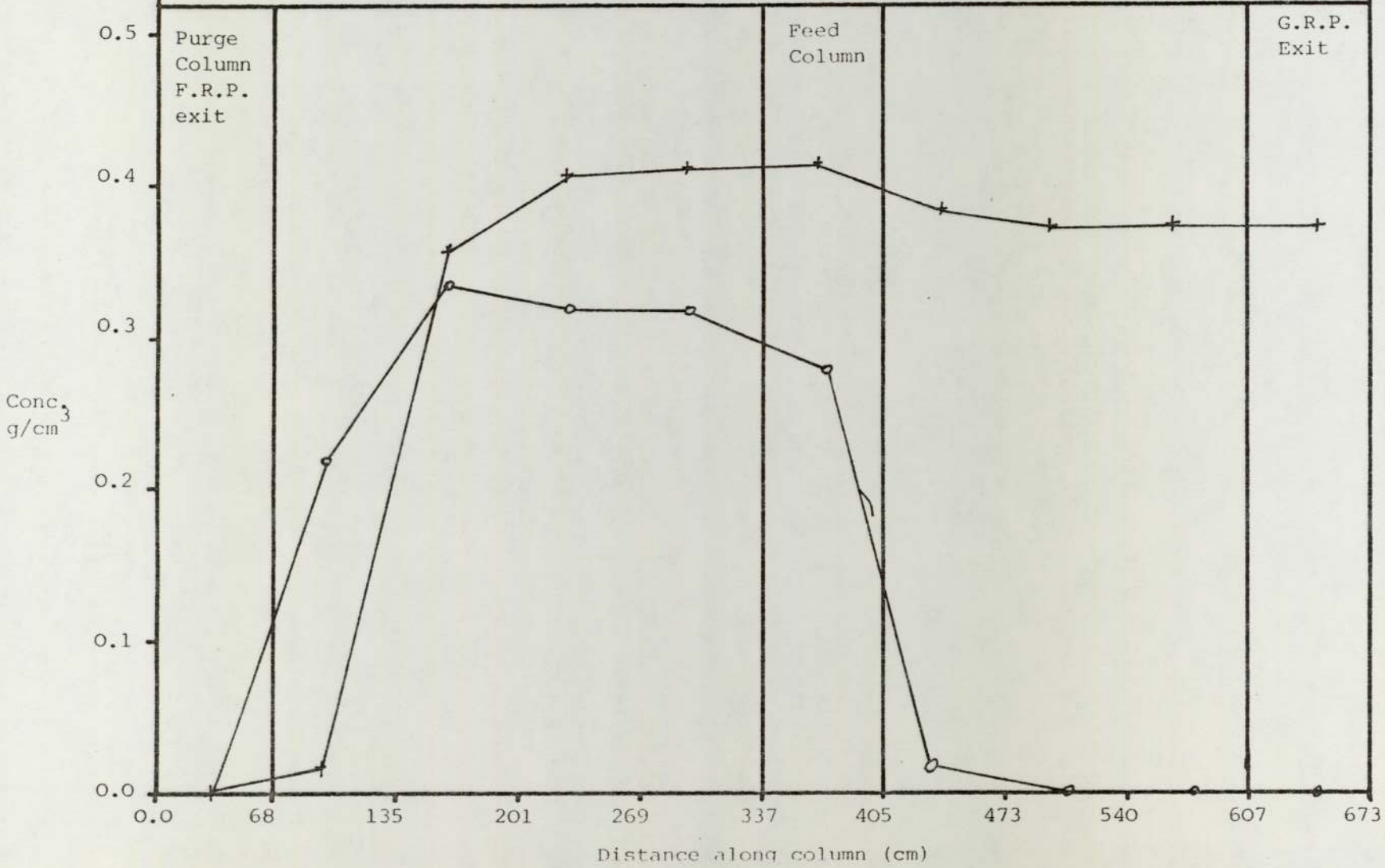
Consequently run 60-35-105-30 was selected as the base run, where the conc. was 60% w/v solids. The experimental run with the feed entering at the fifth column gave a pure glucose rich product but the fructose rich product was contaminated with glucose. The 'cross-over' point of the profiles was located in the second column. The simulation program was used to predict the effect of using the sixth, seventh and eighth columns as feed entry points.

7.5.1 Results and discussion

The experimental equilibrium concentration profile is shown in Fig. 7.10 and the simulated profiles shown in Figs. 7.11-7.13. The glucose profiles have been isolated in Fig. 7.14 and the fructose profiles in Fig. 7.15.

Fig. 7.10 Experimental Concentration Profile for Run 60-35-105-30

Eluent 105 Feed 35 cm^3/min Switch Period 30 mins + Glucose o Fructose



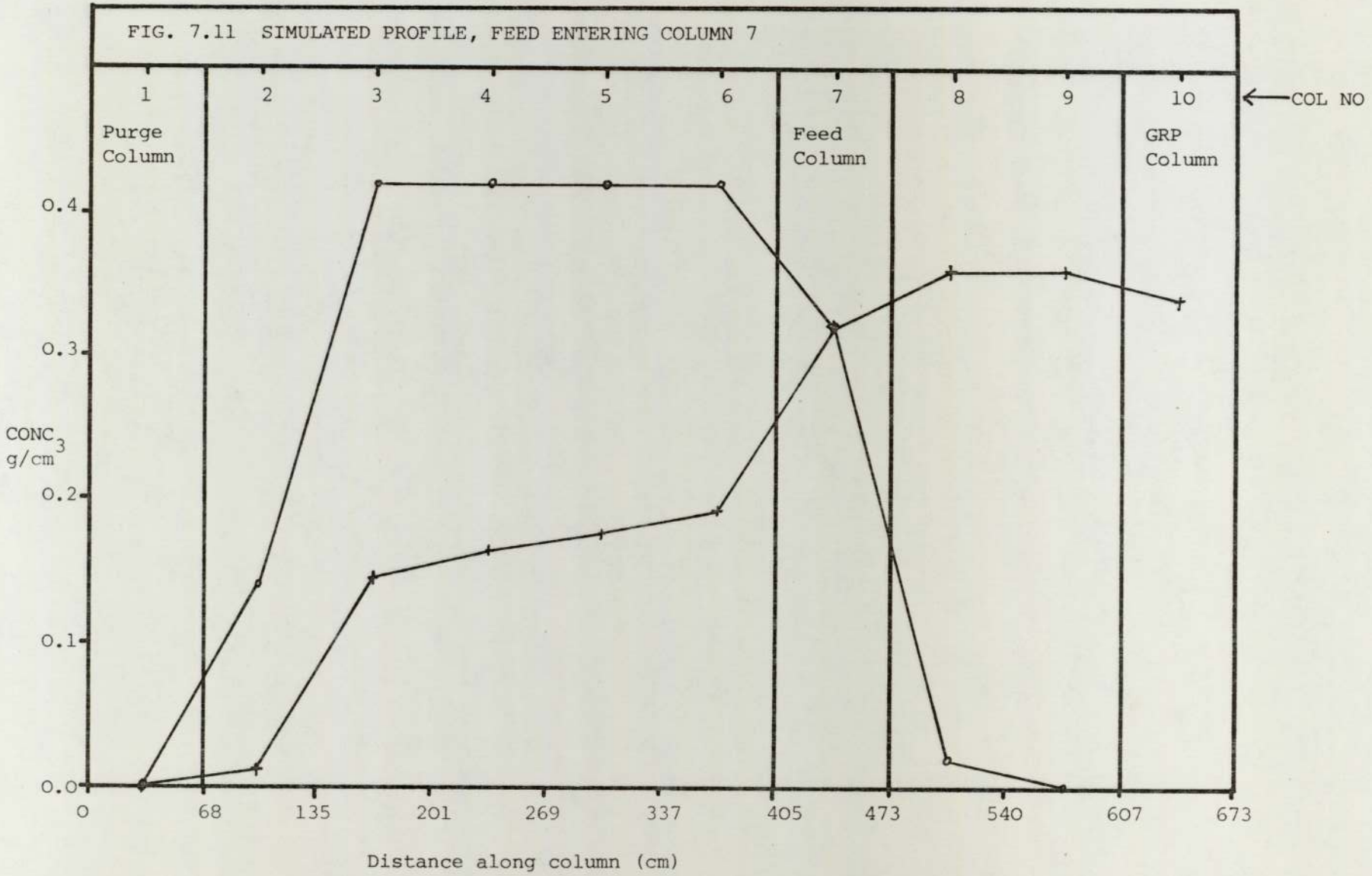


FIG. 7.11 SIMULATED PROFILE, FEED ENTERING COLUMN 7
1
2
3
4
5
6
7
8
9
10
COL NO
Purge Column
Feed Column
GRP Column
CONC g/cm3
0.0
0.1
0.2
0.3
0.4
0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

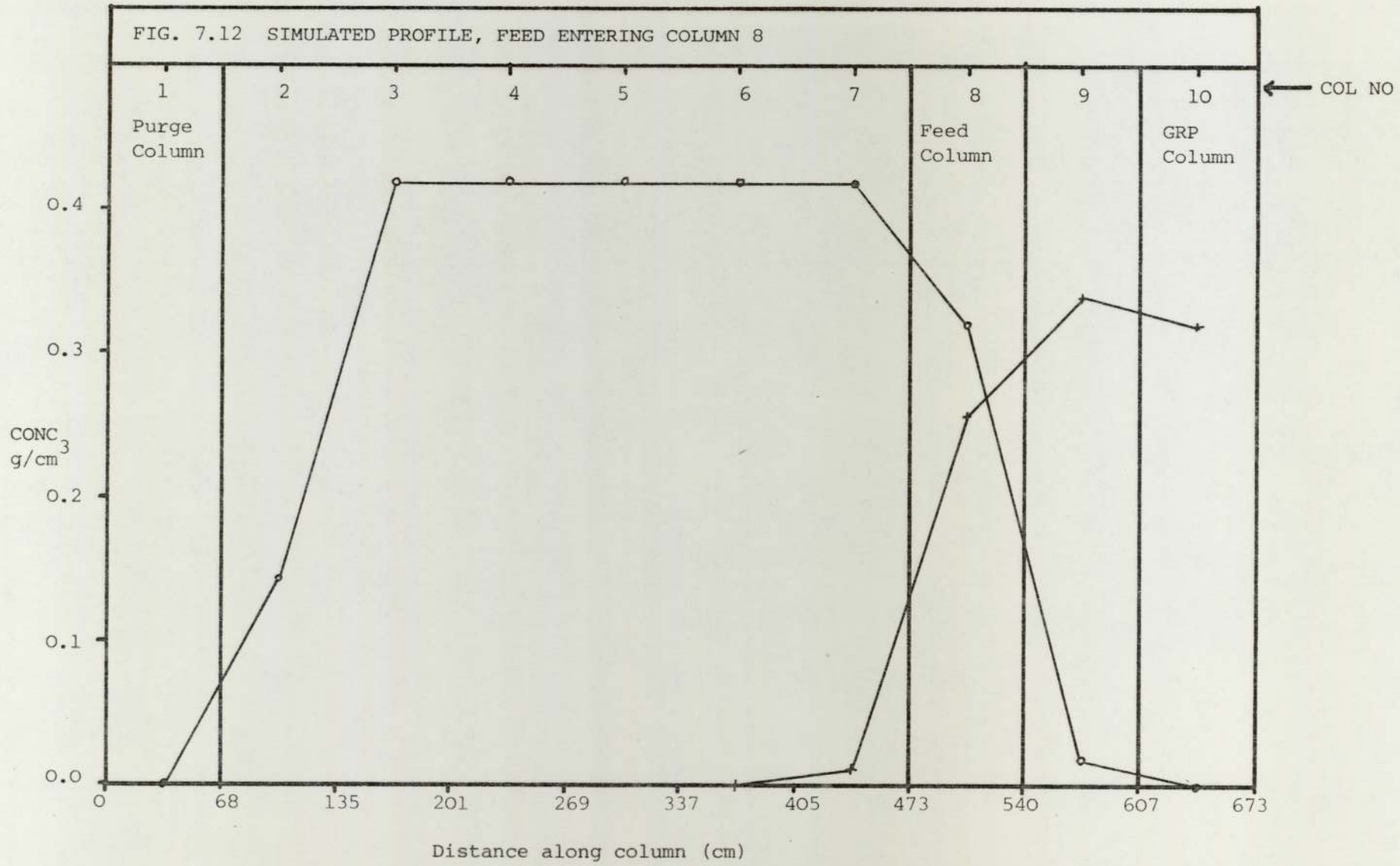


FIG. 7.12 SIMULATED PROFILE, FEED ENTERING COLUMN 8
1
2
3
4
5
6
7
8
9
10
COL NO
Purge Column
Feed Column
GRP Column
0.4
0.3
0.2
0.1
0.0
CONC g/cm3
0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

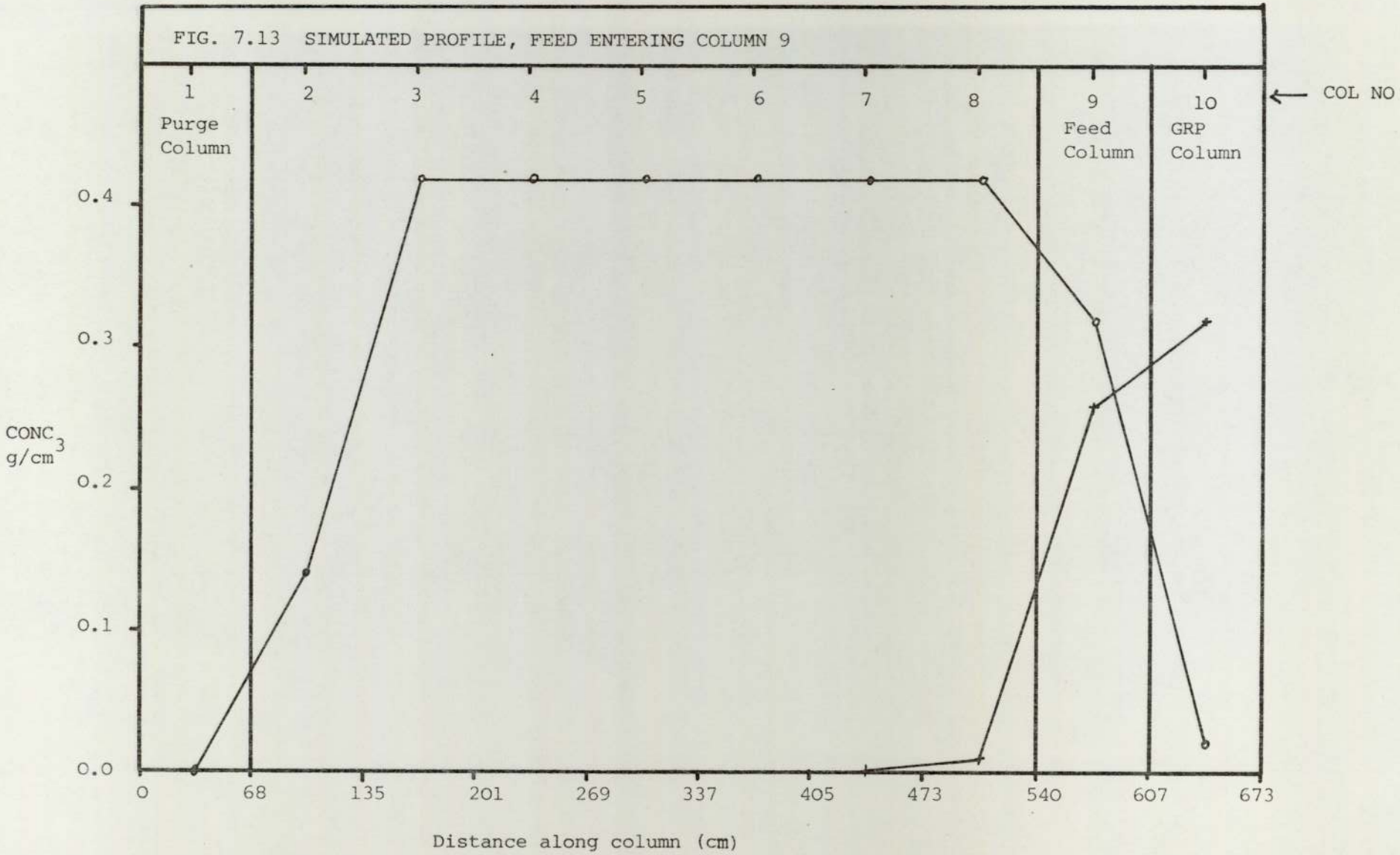


FIG. 7.13 SIMULATED PROFILE, FEED ENTERING COLUMN 9
1
Purge
Column
2
3
4
5
6
7
8
9
Feed
Column
10
GRP
Column
COL NO
0.4
0.3
0.2
0.1
0.0
CONC
g/cm3
0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

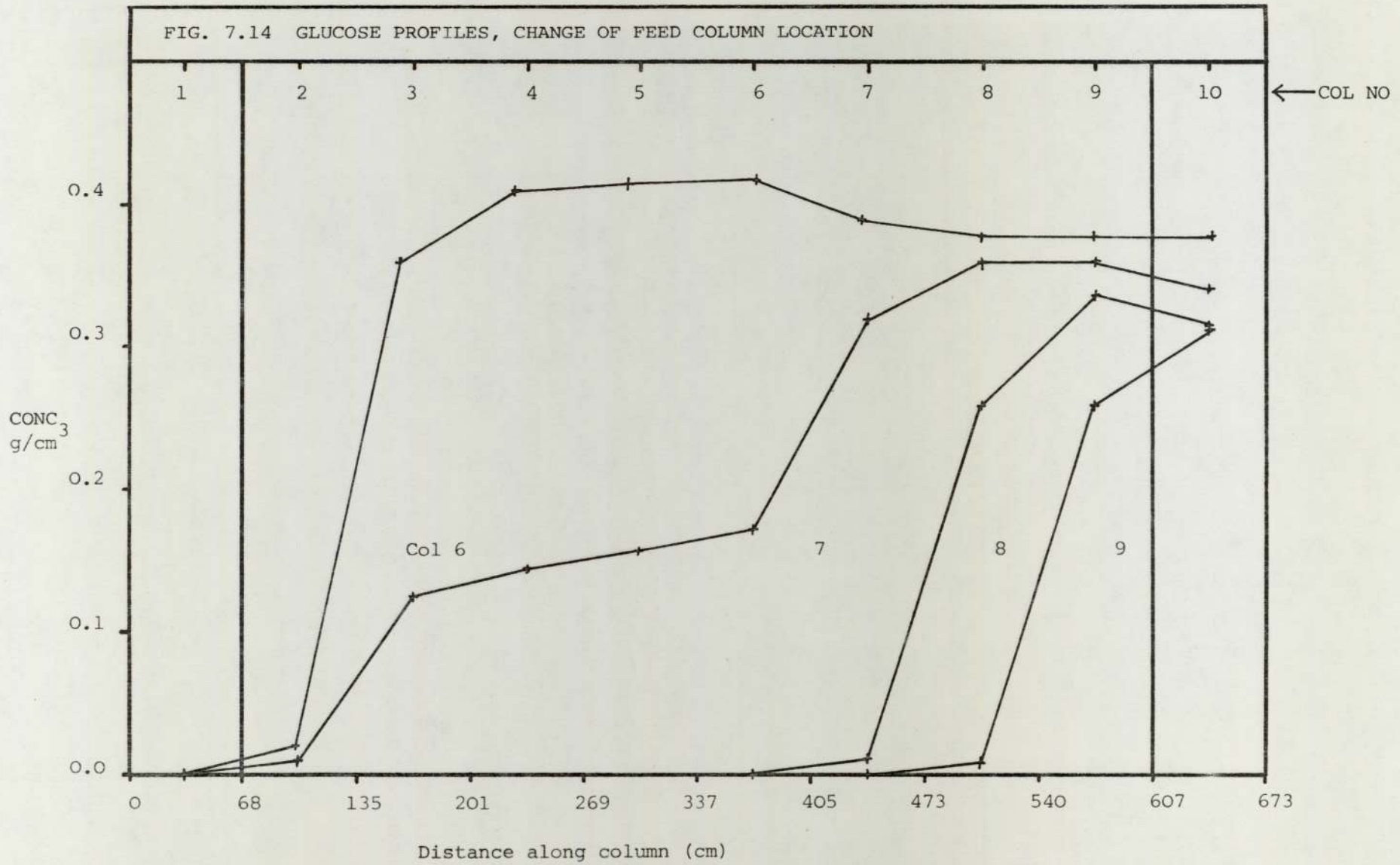


FIG. 7.14 GLUCOSE PROFILES, CHANGE OF FEED COLUMN LOCATION
1
2
3
4
5
6
7
8
9
10
COL NO
0.4
0.3
0.2
0.1
0.0
CONC g/cm3
Col 6
7
8
9
0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

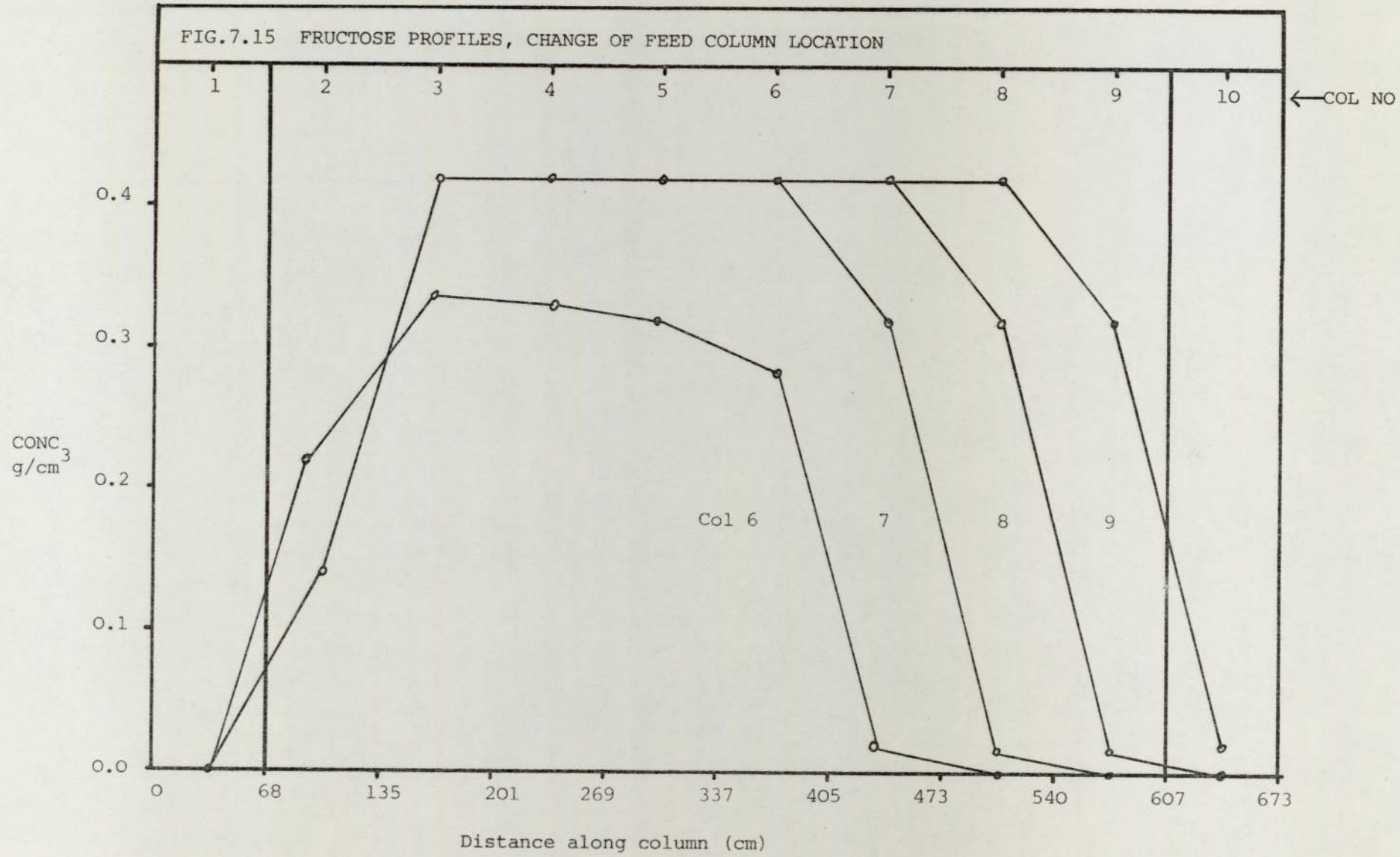


FIG.7.15 FRUCTOSE PROFILES, CHANGE OF FEED COLUMN LOCATION
1
2
3
4
5
6
7
8
9
10
←COL NO
0.4
0.3
0.2
0.1
0.0
CONC g/cm3
Col 6
7
8
9
0
68
135
201
269
337
405
473
540
607
673
Distance along column (cm)

The model predicts that it is possible by change of feed point location to obtain two pure products. This is true when columns six or seven are used. If the feed enters the eighth column fructose will contaminate the glucose rich product as seen in Fig. 7.13.

If the fructose profiles alone are studied, see Fig. 7.15, it may be seen that as the feed entry point moves closer to the glucose exit column the leading edge of the profiles move correspondingly to the right. It is also apparent that fructose only travels one column length beyond the feed column into the post-feed section the on-column concentration reaches a constant value. This concentration decreases at the eluent entry column.

If the glucose profiles are studied, see Fig. 7.14 it may be seen that as the feed entry point is moved nearer to the glucose exit column, the amount of glucose which travels with the stationary phase decreases. This would be expected to happen for the following reason: since the pre-feed section increases in length as the feed entry point is moved nearer to the glucose outlet column, the amount of water in this section increases. This allows the sugars to become more nearly hydrated thus the weak complexes between the glucose and calcium ions are less likely. The on-column concentration in this section is then only due to the fructose present. It is also important to note that as the feedpoint is changed the glucose on-column concentration in the post-feed section tends to decrease. This suggests that more of the glucose remains in the mobile phase and thus its distribution is changing. If this were the case in practice

then the K_d values would also be changing. As discussed earlier the model uses a single value of the K_d throughout the length of the separating section. Thus these simulations should only be used as an indication of what might actually happen until the exact nature of the K_d-concentration relationship is determined.

CHAPTER 8

CONCLUSIONS AND RECOMMENDATIONS

8.1 The research has demonstrated, probably for the first time, that two 99.9% pure products can be obtained from an initial feedstock of a 50% w/w mixture of fructose and glucose. This separation was achieved on a continuous once through basis using the SCCR which was designed and built for this research. It has shown that virtually total product purity has resulted using equipment where the hold up volume is insignificant compared to the total column volume. This appears to confirm the theory that the equipment built by Ching (6) failed to give total purity because of this factor.

In the reported programme of experimental work, equi-concentrations of fructose and glucose were used in the feedstock. There is a potential flexibility of the SCCR to be used for feedstocks of various proportions of sugars to be refined or enhanced. Various product purities may be achieved with a suitable feedstock. As less pure products are required so throughput may be increased.

The highest feed flow rate used which resulted in two pure products was 54 $cm^3 min^{-1}$. This was at a concentration of 20% w/v sugar solids representing a mass throughput of sugar of 0.66 kgh^{-1}.

The maximum mass throughput of sugar in the binary fructose glucose experiments was 1.26 kgh^{-1}. The resultant products were 99.9% pure glucose and 77.0% pure fructose.

The work has shown that approximately seven cycles of operation are necessary for the pseudo-equilibrium state to be reached.

Analysis of the experimental results has led to the following observations: As the concentration of sugar on the column increases more glucose tends to travel into the pre-feed

section with the stationary phase. This is thought to occur because insufficient water molecules are available to allow total hydration of the sugar molecules. In attempts to attain this thermodynamically favourable state it is thought that structural changes may occur in the glucose molecule which result in the formation of a weak complex between the glucose and the calcium ions. Thus they are retarded by the stationary phase. This gives rise to a different distribution of the glucose between the mobile and stationary phases and consequently a change in the distribution coefficient K_{dg}.

The ion exchange resin was able to accommodate all concentrations of feed that were used. It retarded the fructose satisfactorily at all times. The movement of fructose into the post feed section of the separating length has proved to be a function of the mobile phase velocity only.

The limiting pressure drop through the separating section for the on-off valves to operate satisfactorily is approximately 1420 kNm^{-2}. Above this pressure the valves are unable to open at the end of a switch period.

The simulation work has shown that the equilibrium plate concept is only able to give an indication of performance of the SCCR. This is due to a lack of K_d-concentration data. The assumption of instantaneous equilibrium across a theoretical plate is thought to be invalid in the real situation. It is also thought that the use of a single distribution coefficient along the whole of the separating length is a major reason for the model's inaccuracy.

However the average on column concentration of the sugars can be predicted with some degree of accuracy. It is near the eluent entry point, where conditions are changing significantly that improvements need to be made.

8.2 Recommendations for future work

To counteract the effect of glucose migrating with the stationary phase and contaminating the fructose product an investigation into the location of the feed entry column should be carried out. If the feed was introduced nearer to the glucose outlet point it may be possible to minimise the contamination.

To eliminate the problem completely more water needs to be made available. This may be achieved by using a higher eluent to feed ratio. A disadvantage of this suggestion is the more dilute products which would result. A novel suggestion for consideration is the inclusion of a second pure water inlet to the SCCR. This may be located in the column immediately before the feed column. This would ensure adequate water for hydration of the sugar molecules and prevent glucose travelling with the stationary phase. Its flow rate would need to be controlled accurately to prevent purging the fructose from the resin and carrying it to the post feed section. If this were suitable, the fructose product would maintain its purity and not suffer from dilution.

The work needs to be supported by batch experiments to gather data of the exact effects of concentration on the distribution of the components between the two phases. This is essential if improvements to the existing model are to be made. Ideally a routine

needs to be included in the simulation program to predict a K_d value for each theoretical plate. Alternatively the data may allow the formulation of a new approach to the modelling. It may be possible to use the concept of the 'height of a transfer unit' where the packed height necessary for a particular concentration change is predicted.

In continuing attempts to increase the concentration of the products, investigation into the use of the dilute fraction of the fructose product as an eluent may be considered. This would ensure complete recovery of fructose whilst maintaining a higher concentration.

To reduce the dilution which occurs in the purging of the fructose, other mechanisms may be considered. A change in temperature of the purge liquid or perhaps a change in its pH. Care needs to be taken that the equilibrium of the ions associated with the ion exchange resin itself is not destroyed. Again small scale batch work is necessary to assess the feasibility of these suggestions.

To obtain greater throughputs a higher actuating air pressure may be used to increase the operating pressure of the valves which appear to be the limiting mechanical factor. All necessary safety precautions should be observed.

Doubtless the search will continue for an even more suitable stationary phase. The higher the resolution of a column packing the higher the possible throughput. Work is proceeding (91) on tests with an anionic exchange resin which retards the glucose rather than the fructose. It is hoped that product concentrations of the fructose will be again increased.

A constant temperature enclosure has since been built around the SCCR and thus experiments may now be performed at an increased temperature to lower viscosities and hence increase the potential throughput.

A higher temperature is used in practice to prevent any biological growth in the process liquids or the refining equipment. This, along with high feed concentrations and low pH are parameters thought to be used for such control in actual sugar refining processes (92).

With the constant temperature enclosure now constructed experiments can be performed to assess the ability of the resin in the columns to invert and separate a sucrose feedstock into fructose and glucose products.

If the budget allows the semi-continuous counter-current technique may be applied to other feedstocks. Different mobile and stationary phases may be used, depending on the material to be separated or refined. The technique could have great potential as an industrial unit operation.

APPENDIX 1

COMPUTER PROGRAM LISTING AND SAMPLE OF OUTPUT

PROGRAM LISTING

```
C
C
      DIMENSIONG(500),F(500),AG(500),AF(500),GMASS(500),GCUM(500),
     1FMASS(500),FCUM(500) ,GCONC(500),FCONC(500)
      REAL KD1,KD2
      READ(1,1)CFLOW,FFLOW,SFLOW,DT
    1 FORMAT(4F7.2)
      READ(1,2)GFEED,FFEED,KD1,KD2
    2 FORMAT(4F5.2)
      READ(1,3)NFEED,NNBED,KTOTAL,KKINK
    3 FORMAT(4I4)
      CFLOWM=CFLOW*60.0
      SFLOWM=SFLOW*60.0
      FFLOWM=FFLOW*60.0
      SWP=KKINK*DT/60.0
      KCYC=KTOTAL/10
      CFEED=(GFEED+FFEED)*100.0
      V1=2800.0/NNBED
      V2=3200.0/NNBED
      WRITE (2,5)
      WRITE(2,6)
      WRITE(2,7)
      WRITE(2,8)
      WRITE(2,51)CFEED,CFLOWM,FFLOWM,SFLOWM,SWP,NNBED,KD1,KD2
      WRITE(2,9)
      WRITE(2,12)KCYC
      WRITE(2,10)
    5 FORMAT(1H1,///,6X,'FEED',1X,'ELUENT',1X,'FEED',3X,'PURGE',2X,
     1'SWITCH',1X,'NO',6X,'KD1',3X,' KD2')
    6 FORMAT(6X,'CONC',1X,'FLOW',3X,'FLOW',3X,'FLOW',3X,'PERIOD',1X,
     1'OF')
    7 FORMAT(8X,'%',2X,'RATE',3X,'RATE',3X,'RATE',3X,'MINS',3X,
     1'PLATES')
    8 FORMAT(11X,'ML/MIN',1X,'ML/MIN',1X,'ML/MIN',8X,'/ COL'//)
    9 FORMAT(//,10X,'AVERAGE  CONCENTRATION OF SUGARS ON EACH
     1COLUMN')
   12 FORMAT(10X,'AFTER',I3,'CYCLES')
   10 FORMAT(//,9X,'COL NO',2X,'AV GLUCOSE CONCN',2X,'AV
     1 FRUCTOSE CONCN')
   51 FORMAT(6X,F4.1,1X,F5.1,2X,F4.1,3X,F5.1,2X,F4.1,3X,I2,4X,F4.2,
     14X,F4.2)
      DO 99 I=1,500
      G(I)=0.0
      F(I)=0.0
      AG(I)=0.0
      AF(I)=0.0
   99 CONTINUE
```

```
      NNNINE=NNBED*9+1
      NNFEED=(NFEED-1)*NNBED+1
      DO 100 K=1,KTOTAL
      ISTKK=KKINK*(K-1)+1
      LSTKK=KKINK*K
      DO 200 KK=ISTKK,LSTKK
      DO 300 N=1,10
      IF(N.LE.5)CFLOWC=CFLOW
      IF(N.GE.6)CFLOWC=CFLOW+FFLOW
      IF(N.LE.(NFEED-K))GO TO 300
      NNFST=NNBED*(N-1)+1
      NNLST=NNBED*N
      DO 400 NN=NNFST,NNLST
      IF(N.EQ.1)GO TO 80
      IF((N.EQ.2).AND.(NN.EQ.NNFST))GO TO 40
      IF(NN.EQ.NNFEED) GO TO 50
      GO TO 60
   40 G(NN-1)=0.0
      F(NN-1)=0.0
      GO TO 70
   50 A=CFLOWC* DT
      IF(F(NN-1).LT.0.1E-10)F(NN-1)=0.0
      IF(G(NN-1).LT.0.1E-10)G(NN-1)=0.0
      RR=EXP(-A/(V1+V2*KD1))
      SS=EXP(-A/(V1+V2*KD2))
      G(NN)=(1.-RR)*((CFLOW*G(NN-1)+FFLOW*GFEED)/CFLOWC)+RR*G(NN)
      F(NN)=(1.-SS)*((CFLOW*F(NN-1)+FFLOW*FFEED)/CFLOWC)+SS*F(NN)
      GO TO 150
   60 IF(G(NN-1).LT.0.1E-10)G(NN-1)=0.0
      IF(F(NN-1).LT.0.1E-10)F(NN-1)=0.0
   70 A=CFLOWC*DT
      RR=EXP(-A/(V1+V2*KD1))
      SS=EXP(-A/(V1+V2*KD2))
      G(NN)=(1.0-RR)*G(NN-1)+RR*G(NN)
      F(NN)=(1.0-SS)*F(NN-1)+SS*F(NN)
      GO TO 150
   80 IF(NN.EQ.NNFST)GO TO 90
      IF(G(NN-1).LT.0.1E-10)G(NN-1)=0.0
      IF(F(NN-1).LT.0.1E-10)F(NN-1)=0.0
      GO TO 95
   90 G(NN-1)=0.0
      F(NN-1)=0.0
   95 A=SFLOW*DT
      RR=EXP(-A/(V1+V2*KD1))
      SS=EXP(-A/(V1+V2*KD2))
      G(NN)=(1.0-RR)*((G(NN-1)))+RR*G(NN)
      F(NN)=(1.0-SS)*F(NN-1)+SS*F(NN)
  150 IF(K.EQ.KTOTAL.AND.KK.EQ.81000)GO TO 160
      GO TO 400
  160 CONTINUE
  400 CONTINUE
  300 CONTINUE
  200 CONTINUE
      DO 500 NN=1,NNBED
      AG(NN)=G(NN)
      AF(NN)=F(NN)
  500 CONTINUE
      DO 600 NN=1,NNTOT
      IF(NN.GE.NNNINE)GO TO 2010
      NNADJ=NN+NNBED
      G(NN)=G(NNADJ)
```

```
      F(NN)=F(NNADJ)
      GO TO 600
 2010 NNADJ=NN+1-NNNINE
      G(NN)=AG(NNADJ)
      F(NN)=AF(NNADJ)
  600 CONTINUE
  100 CONTINUE
   52 FORMAT(5X, I2, 2X, I6, 2X, I2, 2X, I3, 2X, F10.8, 2X, F10.8)
      FCUM(1)=0.0
      GCUM(1)=0.0
      DO 11 I=1, NNTOT
      GMASS(I)=G(I)*V1+G(I)*KD1*V2
      FMASS(I)=F(I)*V1+F(I)*KD2*V2
   11 CONTINUE
      L=NNBED
   27 DO 14 I=1, NNTOT
      IF(I.EQ.(L+1))GCUM(I)=GMASS(I)
      IF(I.EQ.(L+1))FCUM(I)=FMASS(I)
   14 CONTINUE
      L=L+NNBED
      IF(L.EQ.NNTOT)GO TO 15
      GO TO 27
   15 L=NNBED
      I=2
      GCUM(1)=0.0
      FCUM(1)=0.0
   13 DO 16 M=I, L
      GCUM(M)=GMASS(M)+GCUM(M-1)
      FCUM(M)=FMASS(M)+FCUM(M-1)
      GCONC(M)=GCUM(M)/2745.0
      FCONC(M)=FCUM(M)/2745.0
      KOLNO=M/NNBED
      IF(M.EQ.L) GO TO 18
      GO TO 16
   18 WRITE(2, 4)KOLNO, GCONC(M), FCONC(M)
    4 FORMAT(11X, I2, 8X, F7.5, 12X, F7.5)
   16 CONTINUE
      I=L+2
      L=L+NNBED
      IF(L.EQ.NNTOT)GO TO 17
      GO TO 13
   17 STOP
      END
####S
   1.75    0.58    9.00    2.00
 0.10 0.10 0.09 0.48
   6  25  90 900
****
```

User 2299-JF 6 AUG 81 16:46:18 Pdn 21 No of pages 3

OUTPUT FROM SCCR SIMULATION

FEED CONC %	ELUENT FLOW RATE ML/MIN	FEED FLOW RATE ML/MIN	PURGE FLOW RATE ML/MIN	SWITCH PERIOD MINS	NO OF PLATES / COL	KD1	KD2
40.0	105.0	35.0	540.0	30.0	25	.30	.53

AVERAGE CONCENTRATION OF SUGARS ON EACH COLUMN AFTER 9CYCLES

COL NO	AV GLUCOSE CONC GM/ML	AV FRUCTOSE CONC GM/ML
1	.04175	.07753
2	.26252	.26386
3	.27395	.26507
4	.27395	.26453
5	.27395	.26453
6	.27395	.04829
7	.27395	.00315
8	.27391	.00020
9	.25352	.00001

Nomenclature

A	eddy diffusion mass transfer resistance term in van Deemter equation
A'	column area
B	axial diffusion mass transfer resistance term in van Deemter equation
C_S	stationary phase mass transfer term in van Deemter equation
C_M	mobile phase mass transfer resistance term in van Deemter equation
C	solute concentration in mobile phase
C_1, C_2	gas phase solute concentration at points 1 & 2 in Barker & Lloyds H.T.U. model
C_{n-1}, C_n	concentrations of solute in mobile phase of plate n-1 and n in SCCR simulation model
C_n^o	initial concentration of solute in plate n used in the SCCR simulation model
C_f	feed concentration in SCCR simulation model
D	diffusion coefficient
d_p	mean particle diameter
D_m	mobile phase molecular diffusivity
D_S	stationary phase molecular diffusivity
d_c	column diameter
D_r	radial diffusion coefficient
F_f	feed flow rate in simulation model
F.R.P.	fructose rich product
G.R.P.	glucose rich product
HETP	height equivalent to a theoretical plate

Nomenclature (continued)

K_d	equilibrium distribution coefficient
k'	capacity factor
k''	rate constant of desorption
L	length of packed column
L_e	pre-feed mobile phase velocity
L_f	feed flow rate
L_3	post-feed mobile phase flow rate
L_4	purge flow rate
N	number of theoretical plates in a column
P	stationary phase flow rate
q	solute concentration in stationary phase
Q	mobile phase flow rate in SCCR simulation model
R_s	resolution
r_c	column radius
T_R	elution time of a component
T_o	elution time of a non-retained component
u	mobile phase velocity
V_R	elution volume of a component
V_M	mobile phase volume in a column
V_S	stationary phase volume in a column
V_o	void volume of a column
V_T	total volume of a column
V_1, V_2	volume of mobile and stationary phase in one theoretical plate used in SCCR simulation model
W	peak width at the base line

Nomenclature (continued)

Wh/e	peak width at peak height divided by 'e'
z	distance along column
α	relative retention factor
γ_1, γ_2	rate of transfer of molecules from gas to liquid and liquid to gas in Al Madfai's model
$\delta_1, \delta_2, \delta_3$	series of factors to correct theoretical Le/P operating limits of SCCR
ε	voidage
ρ	mobile phase density
μ	mobile phase viscosity

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